

Pressure-relieving and Depressuring Systems

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Introduction

The portions of this standard dealing with flares and flare systems are an adjunct to API Standard 537, which addresses design, operation, and maintenance of flare equipment. It is important for all parties involved in the design and use of a flare system to have an effective means of communicating and preserving design information about the flare system. To this end, API has developed a set of flare datasheets, which can be found in API Standard 537, Annex E. The use of these datasheets is both recommended and encouraged as a concise, uniform means of recording and communicating design information.

The Bibliography lists the documents that are referenced informatively in this standard, as well as other documents not cited in this standard but which contain additional useful information. Some of the content of the documents listed might not be suitable for all applications and therefore needs to be assessed for each application before use.

In this standard, quantities are expressed in the International System (SI) of units and the US customary (USC) units.

Pressure-relieving and Depressuring Systems

1 Scope

This standard is applicable to pressure-relieving and vapor depressuring systems. Although intended for use primarily in oil refineries, it is also applicable to petrochemical facilities, gas plants, liquefied natural gas (LNG) facilities, and oil and gas production facilities. The information provided is designed to aid in the selection of the system that is most appropriate for the risks and circumstances involved in various installations.

This standard specifies requirements and gives guidelines for the following:

- examining the principal causes of overpressure;
- determining individual relieving rates;
- selecting and designing disposal systems, including such component parts as piping, vessels, flares, and vent stacks.

This standard does not apply to direct-fired steam boilers.

2 Normative References

The following referenced documents are indispensable for the application of this document. For dated references, only the edition cited applies. For undated references, the latest edition of the referenced document (including any amendments) applies.

API Standard 520, *Sizing, Selection, and Installation of Pressure-relieving Devices, Part 1—Sizing and Selection*, Ninth Edition, 2014

API Standard 520, *Sizing, Selection, and Installation of Pressure-relieving Devices, Part 2—Installation*, Sixth Edition, 2015

API Standard 537, *Flare Details for Petroleum, Petrochemical, and Natural Gas Industries*, Third Edition, Addendum 1, 2020

3 Terms, Definitions, Acronyms, and Abbreviations

3.1 Terms and Definitions

For the purposes of this document, the following definitions apply.

3.1.1

accumulation

Pressure increase over the maximum allowable working pressure (MAWP) of the vessel during discharge through the pressure-relief device.

NOTE Accumulation is expressed in units of pressure or as a percentage of MAWP or design pressure. Maximum allowable accumulations are established by pressure design codes for emergency operating and fire contingencies.

3.1.2

administrative controls

Procedures intended to ensure that personnel actions do not compromise the overpressure protection of the equipment.

3.1.3**assist gas**

Flammable gas that is added to relief gas prior to the flare burner or at the point of combustion in order to raise the heating value.

NOTE In some designs, the assist gas can increase turbulence for improved combustion.

3.1.4**atmospheric discharge**

Release from pressure-relieving and depressuring devices to the atmosphere.

3.1.5**availability**

Fraction of time that a system (e.g. safety instrumented system, atmospheric-relief system, or flare-relief system) is able to perform the designated function if required for use.

3.1.6**backpressure**

Pressure that exists at the outlet of a pressure-relief device or depressuring valve as a result of the pressure in the discharge system.

NOTE The backpressure is the sum of the superimposed and built-up backpressures.

3.1.7**balanced pressure-relief valve**

Spring-loaded pressure-relief valve that incorporates a bellows or other means for minimizing the effect of backpressure on the operational characteristics (set pressure, closing pressure, and relieving capacity) of the valve.

3.1.8**blowdown drum**

Knockout drum with or connected to a stack open to atmosphere.

NOTE 1 The term "blowdown drum" is sometimes also used for knockout drums connected to flare or other disposal systems, but it is not used in this context in this standard.

NOTE 2 The term "blowdown" is sometimes used in the context of emergency depressuring of a plant or part of a plant, but it is not used in this context in this standard.

3.1.9**built-up backpressure**

Increase in pressure at the outlet of a pressure-relief device that develops as a result of flow after the pressure-relief device opens.

3.1.10**burnback**

Internal burning within the flare tip.

NOTE Burnback can result from air backing down the flare burner at purge or low flaring rates.

3.1.11**burning velocity****flame velocity**

Speed at which a flame front travels into an unburned combustible mixture.

3.1.12**burn-pit**

Open excavation, normally equipped with a horizontal flare burner that can handle liquid as well as vapor hydrocarbons.

3.1.13**burst pressure**

Value of the upstream static pressure minus the value of the downstream static pressure just before a rupture disk bursts.

NOTE If the downstream pressure is atmospheric, the burst pressure is the upstream static gauge pressure.

3.1.14**cold differential test pressure****CDTP**

Inlet static pressure at which a pressure-relief valve is adjusted to open on the test stand.

NOTE The CDTP includes corrections for the service conditions of backpressure or temperature or both.

3.1.15**combustion air**

Air provided to burn the combustible gases.

3.1.16**combustion efficiency**

The percentage of the total fuel stream entering the flare that is oxidized to form only the products of complete combustion and heat.

NOTE For example, in the case of hydrocarbons, these products are carbon dioxide, water, and heat.

3.1.17**confined fire**

Fire inside a building or a compact process module where the walls and/or surrounding equipment can reradiate and preheat the combustion air, causing higher heat fluxes than an unconfined (i.e. open) fire.

3.1.18**conventional pressure-relief valve**

Spring-loaded pressure-relief valve whose operational characteristics (set pressure, closing pressure, and relieving capacity) are directly affected by changes in the backpressure.

3.1.19**corrected hydrotest pressure**

Hydrostatic test pressure multiplied by the ratio of stress value at upset temperature to the stress value at test temperature.

NOTE 1 See 4.2.2 and C.7.

NOTE 2 In this definition, the hydrostatic test pressure is that specified by the pressure design code, whether or not the equipment has actually been hydrostatically tested.

3.1.20**deflagration**

Explosion in which the flame front is advancing at less than the speed of sound in the unburned combustible mixture.

NOTE See “detonation” (3.1.23).

3.1.21**design pressure**

Pressure, together with the design temperature, used to determine the minimum permissible thickness or physical characteristic of each component, as determined by the design rules of the pressure design code.

NOTE The design pressure is selected by the user to provide a suitable margin above the most severe pressure expected during normal operation at a coincident temperature, and it is the pressure typically specified on the purchase order. The design pressure is equal to or less than the MAWP (the design pressure can be used as the MAWP in cases where the MAWP has not been established).

3.1.22**destruction efficiency**

Mass percent of the original combustible vapor that is no longer present in the combustion products.

3.1.23**detonation**

Explosion in which the flame front is advancing at or above the speed of sound in the unburned combustible mixture.

NOTE See “deflagration” (3.1.20).

3.1.24**dispersion**

Dilution of a vent stream or products of combustion as the fluids move through the atmosphere.

3.1.25**elevated flare**

Flare where the burner is raised high above ground level to reduce radiation intensity and to aid in dispersion.

3.1.26**enclosed flare**

Enclosure with one or more burners arranged in such a manner that the flame is not directly visible.

3.1.27**flame-retention device**

Device used to prevent flame lift-off from a flare burner.

3.1.28**flare**

Device or system used to safely dispose of relief gases in an environmentally compliant manner through the use of combustion.

3.1.29**flare burner****flare tip**

Part of the flare where fuel and air are mixed at the velocities, turbulence, and concentration required to establish and maintain proper ignition and stable combustion.

3.1.30**flare header**

Piping system that collects and delivers the relief gases to the flare.

3.1.31**flashback**

Phenomenon occurring in a flammable mixture of air and gas when the local velocity of the mixture becomes less than the flame velocity, causing the flame to travel back through the mixture.

3.1.32**fuel-controlled fire**

Fire that always has at least enough air for combustion so the amount of fuel (i.e. the size of the pool or the fuel leak rate) controls the heat release rate.

3.1.33**ground flare**

Nonelevated flare.

NOTE A ground flare is normally an enclosed flare but can also be a ground multiburner flare or a burn-pit.

3.1.34**heat release**

Total heat liberated by combustion of the relief gases based on the lower heating value.

3.1.35**hydrate**

Solid, crystalline compound of water and a low-boiling-point gas (e.g. methane and propane), in which the water combines with the gas molecule to form a solid.

3.1.36**jet fire**

Fire created when a leak from a pressurized system ignites and forms a burning jet.

3.1.37**knockout drum**

Vessel in the effluent handling system designed to remove and store liquids.

3.1.38**lateral**

Section of pipe connecting the outlet(s) of single-source relief device(s) to a header where other sources are tied in.

NOTE The relief flow in a lateral is always from a single source, whereas the relief flow in a header can be from either single or multiple sources simultaneously.

3.1.39**lift**

Actual travel of the disc from the closed position when a valve is relieving.

3.1.40**lift-off**

Apparent separation of a stable or unstable flame above the flare burner, occurring if the fuel velocity exceeds the flame velocity.

3.1.41**liquid seal****water seal**

Device that directs the flow of relief gases through a liquid (normally water) on the path to the flare burner, used to protect the flare header from air infiltration or flashback, to divert flow, or to create backpressure for the flare header.

3.1.42**Mach number**

Actual fluid velocity divided by the velocity at which sound waves propagate through the fluid at the associated temperature.

3.1.43**manifold**

Piping system for the collection and/or distribution of a fluid to or from multiple flow paths.

3.1.44**maximum allowable working pressure****MAWP**

Maximum gauge pressure permissible at the top of a completed vessel in its normal operating position at the designated coincident temperature specified for that pressure.

NOTE 1 See “design pressure” (3.1.21).

NOTE 2 The MAWP is the least of the values for maximum allowed working pressure for any of the essential parts of the vessel.

3.1.45**molecular mass**

The SI term “relative molecular mass” is used rather than “molecular weight” in this standard.

3.1.46**noncondensable gas**

Gas or vapor that remains in the gaseous state at a given temperature and pressure.

3.1.47**open fire**

Fire for which the surroundings do not contribute to preheating the ventilation air or reradiation.

NOTE More heat will be lost to the surroundings as compared with confined fires.

3.1.48**operating pressure**

Pressure the process system experiences during normal operation, including normal variations.

3.1.49**overpressure (general)**

Condition where the MAWP, or other specified pressure, is exceeded.

3.1.50**overpressure (relieving device)**

Pressure increase over the set pressure of a relieving device.

NOTE In the context of a relieving device, overpressure is the same as “accumulation” (3.1.1) only when the relieving device is set to open at the MAWP of the vessel.

3.1.51**pilot-operated pressure-relief valve**

Pressure-relief valve in which the major relieving device or main valve is combined with and controlled by a self-actuated auxiliary pressure-relief valve (pilot).

3.1.52**pin-actuated device**

A nonreclosing pressure-relief device actuated by static pressure and designed to function by buckling or breaking a pin that holds a piston or a plug in place. Upon buckling or breaking of the pin, the piston or plug moves to the full open position.

3.1.53**pool fire**

Burning pool of liquid.

3.1.54**pressure design code**

Standard to which the equipment is designed and constructed.

EXAMPLE ASME BPVC, Section VIII: Rules for Construction of Pressure Vessels ^[17].

3.1.55**pressure-relief device**

Device actuated by inlet static pressure and designed to open during emergency or abnormal conditions to prevent a rise of internal fluid pressure in excess of a specified design value.

NOTE The device may be a pressure-relief valve, a rupture disk device, or a pin-actuated device.

3.1.56**pressure-relief valve**

Valve designed to open and relieve excess pressure and to reclose and prevent the further flow of fluid after normal conditions have been restored.

NOTE Other terms used for pressure-relief valve (PRV) include pressure safety valve (PSV), relief valve, safety valve, and safety-relief valve.

3.1.57**pressure system**

System of vessels, pipes, and other equipment operating with an internal pressure exceeding atmospheric.

3.1.58**process tank****process vessel**

Tank or vessel used for an integrated operation in petrochemical facilities, refineries, gas plants, oil and gas production facilities, and other facilities.

NOTE 1 See "storage tank" (3.1.76).

NOTE 2 A process tank or vessel used for an integrated operation can involve, but is not limited to, preparation, separation, reaction, surge control, blending, purification, change in state, energy content, or composition of a material.

3.1.59**purge gas**

Flammable gas or noncondensable inert gas added to the flare header to mitigate air ingress and burnback.

3.1.60**quenching**

Cooling of a fluid by mixing it with another fluid of a lower temperature.

3.1.61**radiation intensity**

Local radiant heat transfer rate from the flare flame, usually considered at grade level.

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3.1.62**rated capacity**

Rated flow of a pressure-relief device, determined in accordance with the pressure design code or regulation and supplied by the manufacturer.

NOTE The capacity marked on the device is the rated capacity on steam, air, gas, or water as required by the applicable code.

3.1.63**relief gas**

All gases and vapors sent to the flare tip, not including entrained air, and consisting of the sum of organic material, nitrogen, and any other gases added to the vent gas collection system, natural gas added as supplemental fuel, nitrogen added as purge gas, natural gas flowing to the flare pilots, and steam added at the flare tip.

NOTE Other terms used for relief gas include flared gas, vent gas, waste gas, and waste vapor.

3.1.64**relief valve**

Spring-loaded pressure-relief valve actuated by the static pressure upstream of the valve, which normally opens in proportion to the pressure increase over the set pressure.

NOTE A relief valve is normally used with incompressible fluids.

3.1.65**relieving conditions**

Inlet pressure and temperature on a pressure-relief device during an overpressure condition.

NOTE The relieving pressure is equal to the valve set pressure (or rupture disk burst pressure) plus the overpressure. The temperature of the flowing fluid at relieving conditions can be higher or lower than the operating temperature.

3.1.66**required relief rate**

Estimated flow rate that needs to be relieved to prevent the equipment pressure from exceeding the specified design value.

3.1.67**risk**

A measure of potential injury, environmental damage, or economic loss in terms of both the incident likelihood and the severity of the loss or injury.

NOTE API 752 ^[8] provides additional discussion of risk.

3.1.68**rupture disk device**

Nonreclosing pressure-relief device actuated by static differential pressure between the inlet and outlet of the device and designed to function by the bursting of a rupture disk.

NOTE A rupture disk device includes a rupture disk and a rupture disk holder.

3.1.69**safety instrumented system****SIS****high-integrity protection system****HIPS**

System composed of sensors, logic solvers, and final control elements for the purpose of taking the process to a safe state when predetermined conditions are met.

NOTE Other terms commonly used for an SIS include emergency shutdown system (ESD, ESS), safety shutdown system (SSD), and safety interlock system (see E.3.3.1).

3.1.70

safety integrity level

SIL

Discrete integrity level of a safety instrumented function in a safety instrumented system.

NOTE SILs are categorized in terms of probability of failure (see Annex E).

3.1.71

safety relief valve

Spring-loaded pressure-relief valve that can be used as either a safety valve or a relief valve depending on the application.

3.1.72

safety valve

Spring-loaded pressure-relief valve actuated by the static pressure upstream of the valve and characterized by rapid opening or pop action.

NOTE A safety valve is normally used with compressible fluids.

3.1.73

set pressure

Inlet gauge pressure at which a pressure-relief device is set to open under service conditions.

3.1.74

staged flare

Group of two or more flares or burners that are controlled so that the number of flares or burners in operation is proportional to the relief gas flow.

3.1.75

stoichiometric air/fuel ratio

Chemically correct ratio of air to fuel capable of perfect combustion with no unused fuel or air.

3.1.76

storage tank

storage vessel

Fixed tank or vessel that is not part of the processing unit in petrochemical facilities, refineries, gas plants, oil and gas production facilities, and other facilities.

NOTE 1 See "process tank" (3.1.58).

NOTE 2 These tanks or vessels are often located in tank farms.

3.1.77

superimposed backpressure

Static pressure that exists at the outlet of a pressure-relief device at the time the device is required to operate.

NOTE It is the result of pressure in the discharge system coming from other sources and can be constant or variable.

3.1.78

vapor depressuring system

Protective arrangement of valves and piping intended to provide for rapid reduction of pressure in equipment by releasing vapors.

NOTE The actuation of the system can be automatic or manual.

3.1.79**vent header**

Piping system that collects and delivers the relief gases to the vent stack.

3.1.80**vent stack**

Elevated vertical termination of a disposal system that discharges vapors into the atmosphere without combustion or conversion of the relieved fluid.

3.1.81**ventilation-controlled fire**

Fire that is in shortage of air such that the available air—the ventilation—controls the heat release rate.

3.1.82**vessel**

Container or structural envelope in which materials are processed, treated, or stored [e.g. pressure vessels, reactor vessels, and storage vessels (tanks)].

3.2 Acronyms and Abbreviations

BLEVE	boiling liquid expanding vapor explosion
BRL	Ballistic Research Laboratory
CCPS	Center for Chemical Process Safety
CDTP	cold differential test pressure
CUI	corrosion under insulation
DN	diameter nominal (nominal diameter)
ERPG	emergency response planning guideline
ESD	emergency shutdown system
ESS	emergency shutdown system
F&G	fire and gas
HIPS	high-integrity protection system
LAH	high liquid level alarm
LAHH	high-high liquid level alarm
LNG	liquefied natural gas
LOF	likelihood of failure
LOPA	layer of protection analysis
LPG	liquefied petroleum gas
LSHH	level switch high-high (high-high level shutdown)
MAWP	maximum allowable working pressure

MOC	management of change
NPS	nominal pipe size
PFD	process flow diagram or probability of failure on demand
PFP	passive fire protection
PHA	process hazard analysis
P&ID	piping and instrumentation diagram
PRD	pressure-relief device
PRV	pressure-relief valve
RPT	rapid phase transition
SI	International System of units
SIL	safety integrity level
SIS	safety instrumented system
SLT	superheat limit temperature
SSD	safety shutdown system
SRU	sulfur recovery unit
USC	US customary units
UTS	ultimate tensile strength
VOC	volatile organic compound
WHSG	waste heat steam generator

4 Causes of Overpressure and Their Relieving Rates

4.1 General

Pressure vessels, heat exchangers, operating equipment, and piping are designed to contain the system pressure. The design is based on the following:

- a) the normal operating pressure at operating temperatures;
- b) the effect of any combination of process upsets that are likely to occur during normal operations;
- c) the differential between the operating and set pressures of the pressure-relief device (PRD);
- d) the effect of any combination of supplemental internal loadings such as static head and external loadings such as earthquake and wind.

Section 4 discusses the principal causes of overpressure where the maximum allowable working pressure (MAWP), design pressure, or other specified pressure can be exceeded. Guidance in plant design to minimize the effects of these overpressure causes and guidance on estimating relieving rates is provided. Overpressure is the result of an unbalance or disruption of the normal flows of material and energy that causes the material or energy, or both, to

build up in some part of the system. Analysis of the causes and magnitudes of overpressure is, therefore, a special and complex study of material and energy balances in a process system.

The principal causes of overpressure listed in 4.4.2 through 4.4.16 are guides to generally accepted practices. The process systems designer shall define the minimum pressure-relief capacity required to prevent the pressure in any piece of equipment from exceeding the maximum allowable accumulated pressure. Annex B provides guidance on the use of a common relief device to protect multiple pieces of equipment from overpressure.

All equipment operations/status should be considered when establishing overpressure protection for the equipment, including, but not limited to, nonroutine situations such as start-up, shutdown, maintenance, standby, and out-of-service as defined by the user. Overpressure protection shall be considered when mass and/or energy exchange is possible for the equipment being considered. Equipment protection should consider the isolation practices as defined by the user. The equipment operations/status and protection methods should be consistent with isolation practice as defined by the user. Table 1 in 4.4 may be helpful in determining the need for overpressure protection for nonroutine situations.

The application of the principles outlined in Section 4 is unique for each processing system. Although efforts have been made to cover all major circumstances, the user is cautioned not to consider the conditions described as the only causes of overpressure. The treatment of overpressure in this standard can be only suggestive. Any circumstance that reasonably constitutes a hazard under the prevailing conditions for a system should be considered in the design. PRDs are installed to ensure that a process system or any of its components is not subjected to pressures that exceed the maximum allowable accumulated pressure. The practices evaluated in Section 4 should be used in conjunction with sound engineering judgment and with full consideration of federal, state, and local rules and regulations.

4.2 Overpressure Protection Philosophy

4.2.1 Hierarchy of Protective Measures

A hierarchy of measures should be used to ensure that equipment is not subject to excess pressure. Such a hierarchy first involves avoiding or reducing risks, then providing engineering controls, and finally providing administrative controls. Avoiding risks includes, for example, setting the MAWP of the equipment above the maximum pressure of all possible sources. Engineering controls include providing pressure relief on the vessel. Administrative controls include provision of block valves of the locked-open design. The user is cautioned that some systems may have unacceptable risk due to failure of administrative controls and resulting consequences due to loss of containment.

4.2.2 Use of Administrative Controls if Corrected Hydrotest Pressure Not Exceeded

It is the responsibility of the user to determine the overpressure scenarios upon which the pressure relief system is designed, and to determine the method of overpressure protection used to mitigate each scenario in accordance with the relevant codes. It is the responsibility of the user to determine whether administrative controls are included in the basis for the pressure relief system design.

If administrative controls are used to eliminate an overpressure scenario as a basis for the pressure relief system design, the user shall evaluate the potential overpressure in the event the administrative control fails, compare it to the equipment corrected hydrotest pressure, and consider additional risk reduction if the corrected hydrotest pressure can be exceeded.

This philosophy is applied to the following scenarios:

- a) closed outlets on vessels (see 4.4.2);
- b) inadvertent valve opening (see 4.4.9.2);
- c) check valve leakage or failure (see 4.4.9.3);
- d) heat transfer equipment failure (see 4.4.14).

The user is cautioned that some systems can have unacceptable risk due to failure of administrative controls and resulting consequences due to loss of containment. In these cases, limiting the overpressure to the normally allowable overpressure can be more appropriate. Note that the entire system, including all of the auxiliary devices (e.g. gasketed joints, instrumentation), should be considered for the overpressure during the failure of administrative controls.

Note that the corrected hydrotest pressure accounts for allowable stress differences of the material of construction between the overpressure scenario temperature (relieving temperature) and either the test temperature or the design temperature (usually higher than the test temperature). Details and examples of the corrected hydrotest pressure are given in C.7.

4.2.3 Double Jeopardy

The causes of overpressure are considered to be unrelated (i.e. independent) if no process or mechanical or electrical linkages exist among them or if the length of time that elapses between possible successive occurrences of these causes is sufficient to make their classification unrelated. The simultaneous occurrence of two or more unrelated causes of overpressure (also known as double or multiple jeopardy) is not a basis for design. Examples of double jeopardy scenarios are fire exposure simultaneous with heat exchanger internal tube failure, fire exposure simultaneous with failure of administrative controls to drain and depressure isolated equipment, or operator error that leads to a blocked outlet coincident with a power failure. On the other hand, instrument air failure during fire exposure may be considered single jeopardy if the fire exposure causes local air line failures. See 4.2.6 for additional guidance.

This standard describes single jeopardy scenarios that should be considered as a basis for design. The user may choose to go beyond these practices and assess multiple jeopardy scenarios, particularly for severe consequence events. Because such assessments are outside the basis for design, the user is not required to meet accumulations allowed by the pressure design code for these scenarios. Acceptance criteria are the sole responsibility of the user.

4.2.4 Latent Failures

Latent failures should normally be considered as an existing condition and not as a cause of overpressure when assessing whether a scenario is single or double jeopardy. For example, latent failures can exist in instrumentation that prevents it from functioning favorably during an overpressure condition. It is not double jeopardy to assume the absence of beneficial instrumentation response in combination with an unrelated overpressure cause. Likewise, it is not double jeopardy to assume a latent failure of a check valve allowing reverse flow during a pump failure.

4.2.5 Operator Error/Effect of Operator Response

Operator error is considered a potential source of overpressure.

The decision to take credit for operator response in determining maximum relieving conditions requires consideration of those who are responsible for operation and an understanding of the consequences of an incorrect action. A commonly accepted time range for the response is between 10 min and 30 min, depending on the complexity of the plant. The effectiveness of this response depends on the process dynamics.

When considering operator intervention, the user should ensure:

- a) there is an alarm of the abnormal event or condition independent of the initiating event, which will alert the operator to intervene;
- b) there is sufficient time for the operator to diagnose the alarm(s) and take appropriate action to intervene and mitigate the specific event of interest. Note that there may be instances of "alarm overload" where many alarms simultaneously sound such as in the event of a power failure;
- c) the operator is trained to take the mitigation action and has the authority to take the actions to mitigate the event; the training includes a written procedure associated with the alarm and corrective actions;
- d) the consequences of operator intervention failure are understood;

- e) the probability of alarm failure and/or ineffective operator intervention (e.g. incorrect action or mitigation does not work) is evaluated.

4.2.6 Role of Instrumentation in Overpressure Protection

Fail-safe devices, automatic start-up equipment, and other conventional instrumentation should not be a substitute for properly sized PRDs as protection against single jeopardy overpressure scenarios. There can be circumstances, however, where the use of PRDs is impractical and reliance on instrumented safeguards is needed. Where this is the case, if permitted by local regulations, a PRD might not be required.

EXAMPLE See ASME BPVC, Section VIII, Division 1, UG-140, *Overpressure Protection by System Design*.

The design shall comply with the local regulations and the user's risk tolerance criteria, whichever is more restrictive. If these risk tolerance criteria are not available to perform analyses per the guidance in Annex E, then as a minimum, the overall system performance including instrumented safeguards should provide safety integrity level 3 (SIL-3) performance in accordance with ISA 84.01 ^[91].

Although favorable response of conventional instrumentation should not be assumed when sizing individual process equipment pressure relief, in the design of some components of a relieving system, such as the collection header, flare, and flare tip, favorable response of some instrument systems can be assumed. The decision to base the design of such systems on excluded or reduced specific loads due to the favorable response of instrument systems should consider the number and reliability of applicable instrument systems. See Section 5 for more details on sizing disposal systems.

Some pipeline design codes (e.g. ASME B31.8 ^[19]) permit the use of a pressure limiting station to protect gas transmission lines, in place of, or in addition to, a PRD. A pressure limiting station is defined as "equipment that under abnormal conditions acts to reduce, restrict, or shutoff the supply of gas flowing into a system to prevent the gas pressure from exceeding a predetermined value." The design options for these pressure limiting stations are prescribed by ASME B31.8 and typically include a combination of multiple pressure regulators in addition to a shutdown valve.

4.3 Determination of Individual Relieving Rates

4.3.1 General Philosophy

The basis for determining individual relieving rates that result from various causes of overpressure is presented in the form of general considerations and specific guidelines. Good engineering judgment, rather than blind adherence to these guidelines, should be followed in each case. The results achieved should be economically, operationally, and mechanically feasible, but in no instance should the safety of a plant or its personnel be compromised.

The liquid or vapor rates used to establish relief requirements are developed by the net energy input. The two most common forms of energy are:

- a) heat input, which increases pressure through vaporization or thermal expansion, and
- b) direct pressure input from higher-pressure sources.

Overpressure can result from one or both of these sources.

The peak individual relieving rate is the maximum rate at which the pressure shall be reduced to protect equipment against overpressure due to any single cause. The probability of two unrelated failures occurring simultaneously is remote and normally does not need to be considered (see 4.2.3).

4.3.2 Effects of Pressure, Temperature, and Composition

Pressure and temperature should be considered to determine individual relieving rates, since they affect the volumetric and compositional behavior of liquids and vapors. Vapor is generated when heat is added to a liquid. The rate at which vapor is generated changes with equilibrium conditions because of the increased pressure in a

confined space and the heat content of streams that continue to flow into and out of the equipment. In many instances, a volume of liquid can be a mixture of components with different boiling points. Heat introduced into fluids that do not reach their critical temperature under pressure-relieving conditions produces a vapor that is rich in low-boiling components. As heat input is continued, the vapor is enriched with heavier components. Finally, if the heat input is sufficient, the heaviest components are vaporized.

During pressure relieving, the changes in vapor rates and relative molecular masses at various time intervals should be investigated to determine the peak relieving rate and the composition of the vapor. The composition of inflowing streams can also be affected by variations in time intervals and therefore requires study.

Relieving pressure can sometimes exceed the critical pressure (or pseudo-critical pressure) of the components in the system. In such cases, reference shall be made to compressibility correlations to compute the density-temperature-enthalpy relationships for the system fluid. If the overpressure is the result of an inflow of excess material, then the excess mass quantity shall be relieved at a temperature determined by equating the incoming enthalpy with the outgoing enthalpy.

In a system that has no other inflow or outflow, if the overpressure is the result of an extraneous excess heat input, the quantity to be relieved is the difference between the initial contents and the calculated remaining contents at any later time. The cumulative extraneous enthalpy input is equal to the total gain in enthalpy by the original contents, whether they remain in the container or are vented. By calculating or plotting the cumulative vent quantity versus time, the maximum instantaneous relieving rate can be determined. This maximum usually occurs near the critical temperature. In such cases, the assumption of an ideal gas can be too conservative, and Equation (9) (see 4.4.13.2.4.3) oversizes the pressure-relief valve (PRV).

4.3.3 Dynamic Simulation

Dynamic simulation can be used in pressure-relief system design to calculate transient pressure increases as indicated in 4.4.14 or to calculate required relief rates from individual PRDs. Conventional methods for calculating relief loads are generally conservative and can lead to overly sized relief and flare system designs. Dynamic simulation provides an alternative method to better define the relief load and improves the understanding of what happens during relief.

Dynamic simulation can be applied wherever conventional methods can be applied. Relief load calculations using dynamic simulation shall follow the same rules set elsewhere in this standard for performing relief load calculations using conventional calculation methods. It can be necessary to perform sensitivity analyses with respect to control response in order to identify appropriate control response. In general, no allowance should be taken for automatic control action unless it tends to increase the relief load. The user is cautioned that the inputs, dynamic calculation methods, and results of the dynamic model need to be validated.

Dynamic simulation can be used in systems such as columns, compression, multiple loop refrigeration systems, and heat exchanger tube ruptures (see 4.4.14.2.2) and in analyzing existing flare systems (see 5.3.4.2). A dynamic simulation of a column is discussed below.

If dynamic simulation is used for column-relief system design, it is necessary to ensure that the model is conservative with respect to calculating the maximum relief load. If the physical phenomena are not well understood, the dynamic simulation model shall include conservative assumptions. These assumptions shall be checked by sensitivity analyses to assess their impact on the column-relief load. For example, several simulation runs can be required to determine the effect of different froth correlations on the tray liquid holdup and resulting relief load. Additionally, it can be advisable to have operating personnel review the appropriate aspects of the model. The user of the dynamic simulation program should be aware of the underlying assumptions that are built into the dynamic simulation software code and how they affect the results. The user should not use a dynamic simulation developed for another purpose (i.e. operator training) and assume that this model gives accurate relief loads without a detailed review of the modeling assumptions.

At steady-state conditions, the dynamic model shall closely match the steady-state model. The model shall reflect current or expected operation. An adequate level of detail is required to assure accurate predictions of peak relief loads. The model shall incorporate physical features of the system (e.g. liquid inventories of vessels and piping). Sensitivity analyses should be performed for the full range of operating conditions (e.g. variable compositions and turndown rates). For example, the differences in tray-draining mechanisms between valve and sieve trays can have

a significant impact on the calculated relief load for some columns. An accurate estimate of tray inventories can also be important where column light-ends inventory can impact the peak relief load.

If dynamic simulation is used, sensitivity analyses shall be performed to assess factors such as the effect of PRDs with excess capacity, the action of automatic controls, controller tuning, heat integration with other columns, and operator intervention. These factors increase the number of run permutations that shall be performed.

4.4 Individual Overpressure Causes and Their Relieving Rates

4.4.1 General

Table 1 lists some common occurrences that can require overpressure protection. This table is not intended to be all-inclusive or complete in suggesting maximum required relieving rates; it is merely recommended as a guide. A more descriptive analysis is provided in the remainder of 4.4.

4.4.2 Closed Outlets

4.4.2.1 Description of a Closed Outlet

The inadvertent closure of a valve on the outlet of pressure equipment while the equipment is on stream can expose the equipment to a pressure that exceeds the MAWP. Every valve (i.e. manual, control, or remotely operated) should be considered as being subject to inadvertent operation. If closure of an outlet valve can result in pressure in excess of that allowed by the design code, a PRD is required. If the equipment is designed to the maximum source pressure, then closure of an outlet valve will not result in overpressure, so a PRD is not required for the closed outlet scenario.

In the case of a manual valve, administrative controls can be used to prevent the closed outlet scenario, subject to 4.2.1 and 4.2.2. Note that the entire system including all of the auxiliary devices (e.g. gasketed joints, instrumentation) should be considered for the overpressure during the failure of administrative controls.

In general, the elimination of block valves and the omission of block valves in between vessels in a series can reduce the number of PRDs.

4.4.2.2 Additional Considerations Involving Pumps

The system does not require relief protection for the closed outlet (i.e. shut-in) scenario if the pump, piping, and other equipment downstream of a centrifugal pump (which can be exposed to the shut-in pressure of the pump, e.g. closed block valve in the discharge system) are designed to withstand the maximum shut-in pressure of the pump or other pressure source such as a start-up or warm-up line.

For positive displacement pumps, pressure-relief protection is usually required to protect the pump itself and downstream equipment against shut-in conditions.

4.4.2.3 Additional Considerations Involving Reciprocating Compressors

Even where there are no block valves between stages, interstage PRD(s) might be required to prevent overpressure in the event of loss of interstage cooling. A remote—but possible—contingency, however, would be internal failure of one of the downstream stages. Loss of downstream stage compression would essentially shift all of the gas compression work to the previous stage. A PRD sized for the blocked outlet flow conditions at the interstage design pressure is required to prevent overpressure.

Table 1—Guidance for Required Relieving Rates Under Selected Conditions

Item No.	Condition	Section	Liquid-relief Guidance ^a	Vapor-relief Guidance ^a
1	Closed outlets	4.4.2	Maximum liquid pump-in rate	Total incoming steam and vapor plus that generated therein at relieving conditions
2	Cooling-water failure to condenser	4.4.3	—	Total vapor to condenser at relieving conditions
3	Top-tower reflux failure	4.4.3	—	Total incoming steam and vapor plus that generated therein at relieving conditions less vapor condensed by sidestream reflux
4	Sidestream reflux failure	4.4.3	—	Difference between vapor entering and leaving equipment at relieving conditions
5	Lean-oil failure to absorber	4.4.4	—	None, normally
6	Accumulation of noncondensables	4.4.5	—	Same effect in towers as found for Item 2; in other vessels, same effect as found for Item 1
7	Entrance of highly volatile material	4.4.6	—	Use alternative means of protection to avoid scenario; see Item 15 for heat exchanger tube rupture guidance
	a) Water into hot oil	4.4.6.1	—	
	b) Light hydrocarbons into hot oil	4.4.6.2	—	
8	Overfilling	4.4.7	Maximum liquid pump-in rate	—
9	Failure of automatic controls	4.4.8	—	Analyze on a case-by-case basis
	a) Inlet control devices and bypasses	4.4.8.3		
	b) Outlet control devices	4.4.8.4		
	c) Fail-stationary valves	4.4.8.5		
	d) Choke valves	4.4.8.6		
10	Abnormal process heat or vapor input	4.4.9	—	
	a) Abnormal process heat input	4.4.9.1		Estimated maximum vapor generation including noncondensables from overheating
	b) Inadvertent valve opening	4.4.9.2		
	c) Check valve failure	4.4.9.3		
11	Internal explosions or transient pressure surges (e.g. water, steam, or condensate hammer)	4.4.10	Not controlled by conventional PRDs but by avoidance of circumstances	Not controlled by conventional PRDs but by avoidance of circumstances
12	Chemical reaction	4.4.11	—	Estimated gas/vapor generation from both normal and uncontrolled conditions; consider two-phase effects
13	Hydraulic expansion	4.4.12		
	a) Cold-fluid shut-in	4.4.12	See 4.4.12	—
	b) Lines outside process area shut-in	4.4.12	See 4.4.12	—
14	Exterior fire	4.4.13		Estimated by the methods given in 4.4.13.2 or Annex A
15	Heat transfer equipment failure	4.4.14		
	a) Heat exchanger tube rupture	4.4.14.2	Liquid flowing across a rupture equal to twice the cross-sectional area of one tube	Steam or vapor flowing across a rupture equal to twice the cross-sectional area of one tube
	b) Double pipe	4.4.14.3		
	c) Plate and frame	4.4.14.4		

Table 1—Guidance for Required Relieving Rates Under Selected Conditions (*continued*)

Item No.	Condition	Section	Liquid-relief Guidance ^a	Vapor-relief Guidance ^a
16	Power failure (steam, electric, or other)	4.4.15	—	Study the installation to determine the effect of power failure; size the relief valve for the worst condition that can occur
	a) Fractionators		—	Loss of all pumps, with the result that reflux and cooling water would fail
	b) Reactors		—	Consider failure of agitation or stirring, quench or retarding stream; size the valves for vapor generation from a runaway reaction
	c) Air-cooled heat exchangers		—	Fan failure; size valves for the difference between normal and emergency duty
	d) Surge vessels		—	Maximum liquid inlet rate
17	Maintenance	4.4.16	—	—
^a Consideration can be given to the reduction of the relief rate as the result of the relieving pressure being above operating pressure.				

4.4.2.4 Relieving Rate Estimation for a Closed Outlet

For determining relief loads, it may be assumed that manual or remotely operated valves that are normally open and functioning at the time of inadvertent closure or failure and that are not affected by the primary cause of failure remain in operation at their normal operating positions. A check of possible common mode failures that can affect multiple valves simultaneously (e.g. control systems, electrical equipment, etc.) should be made to assure that the valves are independent and would not be affected by the primary failure.

The quantity of material to be relieved should be determined at conditions that correspond to relieving conditions instead of at normal operating conditions. The required relieving rate is often reduced appreciably when this difference in conditions is considered. The effect of frictional pressure drop in the connecting line between the source of overpressure and the system being protected should also be considered in determining the required relieving rate.

4.4.3 Cooling or Reflux Failure

4.4.3.1 Description of the Cooling or Reflux Failure

The failure of electrical or mechanical equipment that provides cooling or condensation in process streams can cause overpressure in process vessels. The loss of reflux as a result of pump or instrument failure can cause overpressure in a column because of condenser flooding or loss of coolant in the fractionating process. Fans on air-cooled heat exchangers or cooling towers occasionally become inoperative because of a loss of power or a mechanical breakdown. On cooling towers and air-cooled heat exchangers where independent operation of the louvers can be maintained, credit for the cooling effect may be obtained by convection and radiation.

4.4.3.2 Relieving Rate Estimation for Loss of Cooling or Reflux

4.4.3.2.1 General

The required relieving rate is determined by a heat and material balance on the system at the relieving pressure. In a distillation system, the rate can require calculation with or without reflux. Credit is normally not taken for the effect of residual coolant after the cooling stream fails because this effect is time limited and depends on the physical configuration of the piping. However, if the process piping system is unusually large and bare, the effect of heat loss to the surroundings may be considered.

Because of the difficulty in calculating detailed heat and material balances, the simplified bases described in 4.4.3.2.2 through 4.4.3.2.9 have generally been accepted for determining relieving rates.

4.4.3.2.2 Total Condensing

The required relieving rate is the total incoming vapor rate to the condenser, recalculated at a temperature that corresponds to the new vapor composition at relieving conditions, and the heat input prevailing at the time of relief. The surge capacity of the overhead accumulator at the normal liquid level is generally limited to less than 10 min. If cooling failure exceeds this time, reflux is lost and the overhead composition, temperature, and vapor rate can change significantly.

4.4.3.2.3 Partial Condensing

The required relieving rate is the difference between the incoming and outgoing vapor rate at relieving conditions. The incoming vapor rate should be calculated on the same basis used in 4.4.3.2.2. If the composition or rate of the reflux is changed, the incoming vapor rate to the condenser should be determined for the new conditions.

4.4.3.2.4 Air Cooler Fan Failure

Because of natural convection effects, credit for a partial condensing capacity of 20 % to 30 % of normal air cooler duty is often used, unless the effects at relieving conditions are determined to be significantly different. The required relieving rate is then based on the remaining 70 % to 80 %, depending on the service (see 4.4.3.2.2 and 4.4.3.2.3). However, the actual duty available by natural convection is usually a function of the air-cooled heat exchanger design. Some designs can allow significantly more credits if a supporting engineering analysis is performed. In addition, reduction in cooling capabilities can also occur if variable-pitch fans are used and a failure of the pitch mechanism occurs.

4.4.3.2.5 Louver Closure

Louver closure on air coolers is considered to result in total loss of cooling. Louver closure can result from automatic control failure, mechanical linkage failure, or destructive vibration on a manually positioned louver.

4.4.3.2.6 Overhead Circuit

In many cases, failure of the reflux that results, for example, from pump shutdown or valve closure, causes flooding of the overhead condenser, which is equivalent to total loss of cooling. Compositional changes caused by loss of reflux can produce different vapor properties that affect the required relieving rate. A PRD sized for total failure of the coolant is usually adequate for this condition, but each case shall be examined in relation to the particular components and system involved.

4.4.3.2.7 Pump-around Circuit

The required relieving rate is the vaporization rate caused by an amount of heat equal to that removed in the pump-around circuit. The latent heat of vaporization corresponds to the latent heat under the relieving conditions of temperature and pressure at the point of relief.

4.4.3.2.8 Overhead Circuit Plus Pump-around

An overhead circuit plus pump-around is usually arranged so that simultaneous failure of the pump-around and the overhead condenser do not occur; however, partial failure of one with complete failure of the other is quite possible. The required relieving rate is discussed in 4.4.3.2.6 and 4.4.3.2.7.

4.4.3.2.9 Sidestream Reflux Failure

Principles similar to those described in 4.4.3.2.6 and 4.4.3.2.7 apply in overhead condenser flooding (if a condenser is in the system) or changes in vapor properties resulting from changes in composition. The required relieving rate should be large enough to relieve the vaporization rate caused by the amount of heat normally removed from the system.

4.4.4 Absorbent Flow Failure

For lean-oil absorption of hydrocarbons, generally no relief requirement results from lean-oil failure. However, in an acid gas removal unit in which large quantities (25 % or more) of the inlet vapor can be removed in the absorber, loss of absorbent can cause a pressure rise to relief pressure, since the downstream system might not be adequate to handle the increased vapor flow.

Each individual case shall be studied for its process and instrumentation characteristics. The study should include the effect on downstream process units in addition to the reaction in piping and instrumentation immediately downstream of the absorber.

4.4.5 Accumulation of Noncondensables

Noncondensables do not accumulate under normal conditions because they are released with the process streams. If the noncondensables accumulate to the point of blanketing (vapor-locking) the overhead condenser, the effect may be equal to a total loss of cooling. In addition, some relief scenarios may introduce noncondensables that may have the same effect.

4.4.6 Entrance of Volatile Material into the System

4.4.6.1 General

Violent explosions have occurred in the refining, petrochemical, LNG, and other industries due to mixing water or a light hydrocarbon with a significantly hotter fluid or direct contact of the volatile fluid with a hot surface. These physical explosions are termed “superheat-limit explosions,” “vapor explosions,” “steam explosions,” or “rapid phase transitions” (RPTs) [104]. The commonality of these explosions is that cold, volatile liquid is superheated well above its normal boiling temperature at a given pressure. The consensus of published research on this phenomenon is that in order for such explosions to occur, the hot liquid or surface temperature must exceed the superheat limit temperature (SLT) of the cooler volatile liquid. At constant pressure, the SLT is defined as the highest temperature below thermodynamic critical temperature that a liquid can attain without undergoing RPT to vapor. If the SLT is reached or exceeded, the liquid will flash into vapor, in some cases within microseconds. This timeframe is analogous to a detonation. Similar to a chemical detonation, a superheat limit explosion can produce shock waves that generate significantly more damage than that generated by the volume expansion accompanying the conventional vaporization of a liquid. Because of the shock wave potential, PRDs do not provide any mitigation against a superheat limit explosion.

Examples where superheat limit explosions are possible include:

- a) discharge hot heavies/oil into knockout drum containing a layer of water or light hydrocarbon or vice versa;
- b) introduction of water in the feed to a hot feed surge drum;
- c) start-up of hot oil systems containing residual water from maintenance;
- d) hot oil heat exchanger tube failure;
- e) other inadvertent mixing scenarios due to instrumentation, mechanical, or operator error.

4.4.6.2 Superheat Limit Temperature Data

SLTs are experimentally determined. In the case of water, an SLT of 536 °F (280 °C) at atmospheric pressure has been determined [28] [31]. See Reference 31 for experimentally determined SLTs of other pure components. (Note that the values for T_c for methane, ethane, and propane in Table 1 of Reference [28] should be 190K, 305K, and 369K, respectively).

The SLT for pure components can also be estimated using Equation (1) if the critical temperature and pressure are known [141]:

$$T_{sl} = T_{crit} \left[\left(0.11 \frac{P}{P_{crit}} \right) + 0.89 \right] \quad (1)$$

where

T_{sl} is the superheat limit temperature, expressed in K (°R);

T_{crit} is the thermodynamic critical temperature, expressed in K (°R);

P is the system pressure (e.g. design pressure), expressed in kPa (psia);

P_{crit} is the critical pressure, expressed in kPa (psia).

If the hot fluid or surface temperature cannot exceed the SLT, then PRDs can be considered (see 4.4.6.3). If the hot fluid or surface temperature equals or exceeds the SLT, then PRDs do not provide mitigation (see 4.4.6.4).

4.4.6.3 Pressure-relief Device Design Considerations When Below the Superheat Limit Temperature

PRDs can be considered if the temperature of the hotter fluid does not exceed the SLT of the cooler volatile fluid. However, the user is cautioned that as the temperature approaches the SLT, violent boiling with pressure spikes along with slug flow will likely occur with potential for equipment damage depending upon the equipment configuration. Assuming steady-state boiling under these dynamic conditions can lead to underestimating the peak relief load. Further assessments should be considered to determine whether equipment failure is possible under these circumstances.

4.4.6.4 Design Considerations When Above the Superheat Limit Temperature

The user is cautioned not to analyze the potential for superheating liquid to the SLT given the inherent complexities with the heat transfer and mixing processes within equipment. Incidents demonstrate that only small amounts of the cool, volatile liquid are necessary to cause significant damage from a superheat-limit explosion. Proper design, commissioning, and operation of the process system are essential in attempts to eliminate an explosion possibility when above the SLT.

The following are some precautions that can be taken to minimize, though not eliminate, the potential for superheat-limit explosions:

- a) maintaining minimum circulation of hot oil through equipment on standby in order to minimize collection of water;
- b) avoiding water-collecting pockets;
- c) installing proper steam condensate traps;
- d) installing heat tracing to eliminate condensation;
- e) installing double block and bleed valves on water connections to hot process lines;
- f) installing interlocks to trip sources of heat in the event of water-contaminated feedstock;
- g) commissioning and start-up procedures to allow for gradual temperature increase.

Note that incidents have shown that explosions can occur if the hot fluid is added to cooler volatile fluid or vice versa. Hence, designing the water side of a heat exchanger to be at a lower operating pressure than the hot oil side is not an effective mitigation.

4.4.7 Overfilling

4.4.7.1 General

Many process or surge vessels, including columns and towers, have a liquid level present during normal, start-up, or shutdown conditions. Experience has shown that this equipment can be overfilled under certain conditions. If the source pressure of a liquid feed or supply line can exceed the relief device set pressure and/or the design pressure of the equipment, then overfilling shall be evaluated. System design options to deal with liquid overfill include but are not limited to:

- a) increasing the system design pressure and/or PRD set pressure within pressure design code allowances;
- b) designing a pressure-relief system that can safely accommodate the overfill (including the effects of operator intervention response as discussed in 4.2.5);
- c) installing a safety instrumented system (SIS) to prevent the liquid overfill (see Annex E for additional information on SISs).

For all design options, all phases of operation shall be evaluated. Particular attention shall be given to start-up and other nonroutine operations where process conditions (e.g. flow rates, temperature, and density) can be different from normal and where conditions that lead to overfilling can be more likely to occur when compared with normal operations.

The user shall validate the design (foundation, piping, etc.) consistent with the liquid overfill scenario. In addition, the specified design pressures should have a suitable margin between the maximum pressure that would result during overfill and the PRD set pressure where the consequences of the liquid-relief discharge would be significant. If there is not a suitable margin, then the disposal system should be designed to accommodate the liquid release.

If operator intervention is used as part of the system design, then the risk of failure of the operator to properly intervene shall be addressed. The PRD may discharge back into a disposal system such as a lower-pressure section of the process, a flare, or into another disposal system. If it is a basis for the disposal system design, then the user shall consider liquid overfill when designing the knockout drum, collection headers, etc. The design of the disposal system shall prevent the discharge of liquids above their flash point directly to atmosphere (into the air or on the ground) if there is an unacceptable risk of a vapor cloud explosion or other hazardous condition. See 5.8.2.4 for cautions regarding the atmospheric discharge of liquids. See 5.2 regarding options for the disposal of liquids.

4.4.7.2 Mitigation Measures

When designing the system to mitigate liquid overfill, the following can affect the design and shall be evaluated:

- a) risk of failure of the operator to respond properly;
- b) operator training and operating procedures that include the expected response of instrumentation;

EXAMPLE A differential pressure or displacer-level measurement will read low compared to actual level if the fluid-specific gravity is less than the design-specific gravity. This can mean that the indicated level cannot reach 100 % even if the actual level is well above the measured range.
- c) availability of instrumentation that is required for adequate operator intervention;
- d) availability of instrumentation that is required for SIS response;
- e) potential consequence associated with relief stream disposition (e.g. discharge back into the process, discharge to atmosphere, discharge to a treatment system such as flare, etc.);
- f) the pressure design code requirements.

If operator intervention and/or SIS options are selected, then a risk analysis method shall determine the adequacy of the protection.

4.4.7.3 Level Instruments

Some criteria to consider when evaluating level instrumentation and alarms to demonstrate availability and independence from the basic process control system include the following:

- a) whether the level instruments used for safeguards against overfilling are on separate process taps from the process control system;
- b) whether level instruments used for safeguards against overfilling are susceptible to the same common mode device failures as those used for the basic process control system; diverse instrumentation can minimize the potential for common-mode device failures (e.g. differential pressure and radar, displacer and float, etc.);
- c) whether the programmable level transmitter is set to show a low or zero level when the level exceeds the instrument range;
- d) whether an open system differential pressure cell is used that can show a false low level due to accumulation of liquid in the upper impulse line if it is overfilled or collects condensed liquid;
- e) whether density changes due to temperature effects or composition changes impact readings from differential pressure transmitters;
- f) whether the instruments are proven in use for the specific process applications;
- g) whether the range of at least one of the level measurement(s) can indicate a valid level reading over the full range between the high critical alarm point and any shutdown or interlock point;
- h) whether operating characteristics of the level measurement during off-design, start-up, and shutdown operations are considered in display of level, physical properties, setting alarms, trip points, operator training, and operating procedures;
- i) maintenance and testing frequency required for the instrumentation.

4.4.8 Failure of Automatic Controls

4.4.8.1 General

Automatic control devices directly actuated from the process or indirectly actuated from a process variable (e.g. pressure, flow, liquid level, or temperature) are used at inlets and outlets of vessels or systems. When the transmission signal or operating medium to a final control element (such as a valve operator) fails, the control devices should assume either a fully open or fully closed position according to their basic design. Final control elements that fail in a stationary position should be assumed to fail fully open or fully closed (see 4.4.8.5). The failure of a process-measuring element in a transmitter or controller without coincidental failure of the operating power to the final controlled element should be reviewed to determine the effect on the final controlled element. Operation of the manual bypass valve is discussed in 4.4.8.3. Possible failure of the control device while the manual bypass valve is fully or partially open deserves to be considered; however, this factor is not intended to cover the condition of an undersized control valve.

In evaluating relief considerations, the designer should assume proper sizing of the control valve and unit operation at or near design unless a specific condition exists to the contrary. The designer should be alert to temporary start-up or upset conditions when unit operators are using the control valve's bypass valve. Since these are upset and off-control conditions, the probability that relieving requirements will arise is usually greater than when the unit is running normally under control with all bypasses closed.

4.4.8.2 Capacity Credit

In evaluating relieving requirements due to any cause, any automatic control valves that are not under consideration as causing a relieving requirement and that would tend to relieve the system should be assumed to remain in the position required for minimum normal processing flow. In other words, no credit should be taken for any favorable instrument response. Minimum normal valve position is the expected position of the valve prior to the upset incident, that is, the position of the valve when at minimum design flow rates (unit turndown conditions). Therefore, unless the condition of flow through the control valves changes (see 4.4.8.7), credit can be taken for the normal minimum flow of these valves, corrected to relieving conditions, provided that the downstream system is capable of handling any increased flow. Although controllers actuated by variables other than the system pressure can try to open their valves fully, credit can be taken for such control valves only to the extent permitted by their operating position at normal minimum flow regardless of the valve's initial condition.

4.4.8.3 Inlet Control Devices and Bypass Valves

There can be single or multiple inlet lines fitted with control devices. The scenario to consider is that one inlet valve is in a fully opened position regardless of the control valve failure position. Opening of this control valve can be caused by instrument failure or misoperation. If the system has multiple inlets, the position of any control device in those remaining lines shall be assumed to remain in its normal operating position. Therefore, the required relieving rate is the difference between the maximum expected inlet flow and the normal outlet flow, adjusted for relieving conditions and considering unit turndown, assuming that the other valves in the system are still in operating position at normal flow (i.e. normally open, normally closed, or throttling). If one or more of the outlet valves are closed, or more inlet valves are opened by the same failure that caused the first inlet valve to open, the required relieving rate is the difference between the maximum expected inlet flow and the normal flow from the outlet valves that remain open. All flows should be calculated at relieving conditions. An important consideration is the effect of having a manual bypass on the inlet control valve(s) at least partially open. If, during operation, the bypass valve is opened to provide additional flow, then this total flow (control valve wide open and bypass valve normal position) shall be considered in the relieving scenario.

The potential for the bypass valve to be inadvertently opened (e.g. during normal operations, control valve maintenance, start-up, shutdown, or special operations) while the control valve is operating (both bypass and control valve wide open) should also be considered. Administrative controls can be used to prevent inadvertent opening of the bypass valve, subject to 4.2.1 and 4.2.2. The user is cautioned that some systems can have unacceptable risk due to failure of administrative controls and resulting consequences due to loss of containment. In these cases, limiting the overpressure to the normally allowable overpressure can be more appropriate. Note that the entire system, including all of the auxiliary devices (e.g. gasketed joints, instrumentation), should be considered for the overpressure during the failure of administrative controls.

Other situations can arise where problems involved in evaluating relief requirements after the failure of an inlet control device are more complex and of special concern (e.g. a pressure vessel operating at a high pressure where liquid bottoms are on level control and discharge into a lower-pressure system). Usually, when the liquid is let down from the high-pressure vessel into the low-pressure system, only the flashing effect is of concern in the event that the low-pressure system has a closed outlet. However, the designer should also consider that vapors flow into the low-pressure system if loss of liquid level occurs in the vessel at higher pressure. In this case, if the volume of the source of incoming vapors is large compared with the volume of the low-pressure system or if the source of vapor is unlimited, serious overpressure can rapidly develop. When this occurs, it can be necessary to size relief devices on the low-pressure system to handle the full vapor flow through the liquid control valve. The loss of liquid level followed by high-pressure vapor flow is commonly referred to as "vapor breakthrough" or "gas blowby." Annex G provides additional considerations to evaluate the impact of the "vapor breakthrough" or "gas blowby" on low-pressure system and overpressure relief requirements.

In circumstances where process systems involve significant differences in pressure level and the volume of vapor contained by the high-pressure equipment is less than the volume of the low-pressure system, the additional pressure can, in some cases, be absorbed without overpressure.

In the event of loss of liquid level, the vapor flow into the low-pressure system depends on what the interconnecting system, which usually consists of wide-open valves and piping, passes with a differential pressure based on the normal operating pressure upstream and the relieving pressure on equipment downstream. The user is cautioned

that a higher upstream pressure should be used if it occurs coincident with loss of liquid level. This pressure drop at initial conditions frequently results in critical flow (choking across a control valve) and can cause the rate to be several times higher than the normal rate of vapor inflow to the high-pressure system. Unless makeup equals outflow, this condition is of short duration as the upstream reservoir is depleted. Nonetheless, the relief facilities that protect the low-pressure system shall be sized to handle the peak flow. If the low-pressure side has a large vapor volume, it can prove worthwhile to take credit for the following: The transfer of vapors from the high-pressure system needed to raise the pressure on the downstream side from operating pressure to relieving pressure (normally 110 % of design pressure or MAWP) lowers the upstream pressure. This decrease produces a corresponding reduction in the flow that establishes the relieving requirement. Where such credit is taken, an allowance shall be made for the normal makeup of vapor to the high-pressure system that tends to maintain upstream pressure. The vapor breakthrough flow rate can also be reduced by, for example, the use of a smaller valve, reduced trim, mechanical stop, or an orifice restriction in the flow path.

See 4.4.17 for additional guidance concerning piping design for cases with potential vapor breakthrough.

4.4.8.4 Outlet Control Devices

Each outlet control valve should be considered in both the fully opened and the fully closed positions for the purposes of relief load determination. This is regardless of the control valve failure position because failure can be caused by instrument system failure or misoperation. If one or more of the inlet valves are opened by the same failure that caused the outlet valve to close, PRDs can be required to prevent overpressure. The required relieving rate is the difference between the maximum inlet and maximum outlet flows. All flows should be calculated at relieving conditions. Also, one should consider the effects of inadvertent closure of control devices by operator action.

For applications involving single outlets with control devices that fail in the closed position, PRDs can be required to prevent overpressure. The required relieving rate is equal to the maximum expected inlet flow at relieving conditions and should be determined as outlined in 4.4.2.4.

For applications involving more than one outlet and a control device that fails in the closed position on an individual outlet, the required relieving rate is the difference between the maximum expected inlet flow and the design flow (adjusted for relieving conditions and considering unit turndown) through the remaining outlets, assuming that the other valves in the system remain in their normal operating position.

For applications involving more than one outlet, each with control devices that fail in the closed position because of the same failure, the required relieving rate is equal to the maximum expected inlet flow at relieving conditions.

4.4.8.5 Fail-stationary Valves

Even though some control devices are designed to remain stationary in the last controlled position, one cannot predict the position of the valve at time of failure. Therefore, the designer should always consider that such devices could be either open or closed: no reduction in required relieving rate should be considered when such devices are used.

4.4.8.6 Choke Valve Failure

Experience has shown that, under certain conditions and in certain services, choke valves may be subject to a sudden and complete failure of the valve trim, leading to a dramatic increase in the valve's flow capacity. Such a failure may lead to a sustained flow that is substantially larger than normal flow. The design flow should include the additional flow arising from a single failed choke as the worst-case flow for overpressure calculations and system response requirements. A dynamic analysis may be required to determine these effects (see 4.3.3). The user should determine whether choke valve collapse is a basis for design or not credible.

Production chokes [e.g. subsea choke, topside (riser) choke, wellhead choke] may be subject to a collapse and subsequent sudden increase in flow capacity. Among the possible causes are:

- a) impact by solids from the formation (e.g. during a well cleanup or well start-up);

- b) impact by fragments of tools/devices from the well;
- c) erosion/corrosion;
- d) mechanical fatigue;
- e) brittle fracture.

Measures to consider for mitigating the consequences of a choke failure include:

- a) pressure-relief device;
- b) replacing a large valve with multiple smaller valves, as it is necessary to consider only one choke valve failure at a time;
- c) the user may define choke failure as not credible based on selection of high-integrity choke designs; such a conclusion should be substantiated by a careful analysis of the nature and characteristics of mechanical loads and/or a program for testing and qualification;
- d) use of dedicated choke valves or change of choke trim for special operations in which solids are expected, such as well cleanup;
- e) restrictions (e.g. perforated plate) in the flow path that will serve to limit the maximum mechanical load caused by impacts.

The user should avoid rapid opening of the choke valve and carefully evaluate the need for regular inspection and maintenance programs for the choke valves.

4.4.8.7 Special Capacity Considerations

Although control devices, such as diaphragm-operated control valves, are specified and sized for normal design operating conditions, they are also expected to operate during upset conditions, including periods when PRDs are relieving. Valve design and valve operator capability should be selected to position the valve plug properly in accordance with control signals during abnormal conditions. Because the control valve capacities at pressure-relieving conditions are not the same as those at normal conditions, the control valve capacities should be calculated for the relieving conditions of temperature and pressure in determining the required relieving rates. In extreme cases, the state of the controlled fluid can change (e.g. from liquid to gas or from gas to liquid). The wide-open capacity of a control valve selected to handle a liquid can, for example, differ greatly when it handles a gas. This becomes a matter of particular concern where loss of liquid level can occur, causing the valve to pass high-pressure gas to a system sized to handle only the vapor flashed from the normal liquid entry. This is the vapor breakthrough scenario described in 4.4.8.3.

4.4.9 Abnormal Process Heat or Vapor Input

4.4.9.1 Abnormal Process Heat Input

Reboilers and other process heating equipment are designed with a specified heat input. When they are new or recently cleaned, additional heat input above the normal design can occur. In the event of a failure of temperature control, vapor generation can exceed the process system's ability to condense or otherwise absorb the buildup of pressure, which may include noncondensables caused by overheating.

The required relieving rate is the maximum rate of vapor generation at relieving conditions (including any noncondensables produced from overheating) less the rate of normal condensation or vapor outflow. In every case, the designer should consider the potential behavior of the system and each of its components. For example, the fuel or heat-medium control valve or the tube heat flux can be the limiting consideration. To be consistent with the practice used for other causes of overpressure, design values should be used for an item such as control valve size.

However, built-in overcapacity, which is applicable to the common practice of specifying burners capable of 125 % of heater design heat input, shall be considered.

If limit stops are installed on control valves, the wide-open capacity, rather than the capacity at the stop setting, should normally be used. However, if a mechanical stop is installed and is adequately documented, use of the limited capacity can be appropriate. In shell-and-tube heat exchange equipment, heat input should be calculated on the basis of clean, rather than fouled, conditions.

4.4.9.2 Inadvertent Valve Opening

The inadvertent opening of any valve from a source of higher pressure, such as high-pressure steam or process fluids, should be considered. Administrative controls can be used to prevent inadvertent valve opening, subject to 4.2.1 and 4.2.2. The relief load should be determined using the maximum operating pressure upstream of the valve and the relieving pressure on equipment downstream of the valve. If the pressure source is a pipeline or a production well, the pressure upstream of the valve may reach the maximum shut-in pressure of the source after a shutdown. The user should determine whether inadvertent valve opening combined with maximum shut-in pressure in upstream system is a credible relief case.

The following applies when a manual or actuated valve is inadvertently opened, causing pressure buildup in a vessel. The vessel should have a PRD large enough to pass a rate equal to the flow through the open valve; credit may be taken for the flow capacity of vessel outlets that can reasonably be expected to remain open. The manual or actuated valve should be considered as passing its capacity at a full-open position with the pressure in the vessel at relieving conditions. Volumetric or heat-content equivalents may be used if the manual or actuated valve admits a liquid that flashes or a fluid that causes vaporizing of the vessel contents. It is typical to consider only one inadvertently opened manual or actuated valve at a time, although simultaneous inadvertent opening of multiple valves shall be considered if a common cause is identified (e.g. sequential valve operations). Automatic control failures are discussed separately in 4.4.8.

See 4.4.17 for additional guidance concerning piping design for cases with inadvertent valve opening.

4.4.9.3 Check Valve Leakage or Failure

4.4.9.3.1 General

Check valves are intended to limit reverse flow but these are not an effective means for preventing overpressure by reverse flow because check valves can leak or can fail. For example, if a fluid is pumped into a system that contains vapor at significantly higher pressure than the design rating of the equipment upstream of the pump, loss of pumped flow with leakage or failure of a check valve in the discharge line results in a reversal of the liquid's flow. This flow reversal could result in overpressure. This standard describes the following three reverse flow conditions and when these should be considered.

- a) Complete check valve failure—Potential for gross reverse flow because the check valve does not function at all (e.g. it is stuck wide open or the internals are gone).
- b) Severe check valve leakage—Potential for significant reverse flow because of check valve seat damage or obstruction.
- c) Normal check valve leakage—Potential for minor reverse flow due to normal check valve wear.

4.4.9.3.2 Overpressure Due to Reverse Flow Considerations

4.4.9.3.2.1 Normal Check Valve Leakage

The potential to overpressure piping or equipment due to a blocked- in condition and normal check valve leakage should be considered. Potential overpressure can be avoided by specifying higher design pressures, may be managed using administrative controls, subject to 4.2.1 and 4.2.2, or may be mitigated by providing pressure relief. If the design includes a PRD, it is up to the user to determine how to size this device.

Administrative controls may be effective in some scenarios but not effective in others, even for the same equipment. For example, while lining up a spare pump, administrative controls may be effective in preventing overpressure caused by reverse flow across the check valve caused by opening the discharge valve before the pump is started. However, administrative controls would not be effective at preventing reverse flow across the check valve if the same pump stops during normal operation.

4.4.9.3.2.2 Severe Check Valve Leakage

In addition to the normal check valve leakage case, consideration should be given to the more severe case where reverse flow may occur through one or more check valves in series. Where the maximum operating pressure of the high-pressure system is greater than the low-pressure equipment's corrected hydrotest pressure (see 3.1.19 and 4.2.2), consider additional risk reduction beyond administrative controls. Note that the entire low-pressure system, including vessels, piping and auxiliary devices (e.g. gasketed joints, instrumentation), should be considered for the overpressure scenario.

For certain rotating equipment such as reciprocating compressors and positive displacement pumps, the user may determine that significant reverse flow through the machinery will not occur. Overpressure due to normal check valve leakage should still be considered. If a recycle line is installed downstream of such equipment, then significant reverse flow is a potential due to opening of the recycle valve.

In most cases, focus should be on prevention of reverse flow. It is important to note that, in addition to overpressure of the upstream system, reverse flow through machinery can damage rotating equipment, potentially causing loss of containment. Additional means of reverse flow prevention should be provided.

Experience has shown that two reverse flow prevention devices in series are typically sufficient to eliminate significant reverse flow when these are tested or internally inspected and maintained to demonstrate reliability. As the differential pressure increases, the use of additional safeguards should be considered to reduce the risk of check valve failures that could lead to mechanical equipment damage and/or loss of containment. The user might want to consider dissimilar reverse flow prevention devices (e.g. swing check and a piston check valve). For two or more check valves in series that are internally inspected and maintained to demonstrate reliability, only normal check valve leakage may be assumed.

4.4.9.3.3 Calculation of Reverse Flow Through Failed Check Valves

Complete check valve failure is assumed for all check valves in series that are not inspected and maintained and for a single check valve regardless if it is inspected and maintained. The following assumptions may be made to estimate the reverse flow through a failed check valve (select one):

- a) the failed check valve has no flow resistance in the reverse flow direction;
- b) the failed check valve has the same flow resistance in the reverse flow direction as in the forward flow direction;
- c) the failed check valve is equivalent to an orifice equal to the check valve flow area without internals.

For check valves that are internally inspected and maintained but reliability does not meet the user's risk criteria, it will be necessary to estimate the reverse flow. The severity of reverse flow through these check valves depends on the following:

- a) the rigor and experience with the user's check valve inspection/maintenance program;
- b) the types of check valves;
- c) the fouling nature of the fluid;
- d) the materials of construction;
- e) whether the service is pulsating;
- f) other system considerations.

Therefore, it is the responsibility of the user to determine an appropriate technique for estimating the reverse flow. Where no specific experience or company guidelines exist, the following approach may be used.

- a) Where multiple check valves in series are inspected and maintained, assume the smallest check valve has completely failed and all other inspected and maintained check valves have severe leakage. Because of potential common mode failures the user is cautioned against taking a larger credit for more than two check valves in series that are inspected and maintained.

There are widely differing practices to estimate reverse flow with an inspected and maintained check valve that has severe leakage. It is up to the user to determine how to estimate. Two practices are presented below.

- 1) Treat the check valve as an orifice with a diameter equal to 10 % of the check valve's nominal flow diameter (i.e. orifice area equal to 1 % of the nominal flow area). The reverse flow is calculated using the orifice area, the relief fluid properties, the maximum operating pressure on the side normally downstream of the check valve, and the allowable accumulation on the side normally upstream of the check valve.
 - 2) Treat the check valve as an orifice that is sized to pass 10 % of the normal forward flow. The orifice area is calculated using the normal fluid properties, maximum operating pressure on the upstream side of the check valve, and the maximum operating pressure on the downstream side of the check valve. To determine the reverse flow rate, this orifice is then rated using the relief fluid properties, the maximum operating pressure on the side normally downstream of the check valve, and the allowable accumulation on the side normally upstream of the check valve.
- b) The pressures used on each side of the "orifice" can reflect credit for pressure drop in the piping, valves, and equipment. The hydraulics profile should be based on the high-pressure side at maximum operating pressure and the low-pressure side at allowable accumulation. The hydraulic calculations should take into account the effect of fluid phase change during the reverse flow when applicable (e.g. reverse vapor breakthrough).

4.4.9.4 Reciprocating Compressor Rod Packing Failure

Reciprocating compressors should be protected from rod packing failures in the distance piece by an adequately sized vent line or PRD. Options to size the PRD or vent line can include the following:

- a) determine the annular gap between the compressor rod and packing gland and calculate flow using an equivalent square edge orifice area, or
- b) provide a PRD with inlet size equal to the nominal pipe size (NPS) of the vent piping.

An alarm may be provided to indicate to operators that there is abnormally high packing leak off rate (e.g. a high-temperature alarm or a high-pressure alarm).

4.4.10 Internal Explosions or Transient Pressure Surges

4.4.10.1 Internal Explosion (Excluding Detonation)

If overpressure protection is to be provided against internal explosions caused by ignition of vapor-air mixtures where the flame speed is subsonic (i.e. deflagration but not detonation), rupture disks or explosion vent panels, not relief valves, should be used. These devices respond in milliseconds. In contrast, relief valves react too slowly to protect the vessel against the extremely rapid pressure buildup caused by internal flame propagation. The vent area required is a function of a number of factors including the following:

- a) initial conditions (pressure, temperature, composition);
- b) flame propagation properties of the specific vapors or gases;
- c) volume of the vessel;

- d) pressure at which the vent device activates;
- e) maximum pressure that can be tolerated during a vented explosion incident.

It should also be noted that the peak pressure reached during a vented explosion is usually higher, sometimes much higher, than the pressure at which the vent device activates.

Design of explosion-relief systems should follow recognized guidelines such as those contained in NFPA 68 ^[125]. Simplified rules-of-thumb should not be used as these can lead to inadequate designs. If the operating conditions of the vessel to be protected are outside the range over which the design procedure applies, explosion vent designs should be based on specific test data, or an alternate means of explosion protection should be used.

Some alternate means of explosion protection are described in NFPA 69 ^[126], including explosion containment, explosion suppression, oxidant-concentration reduction, and so forth.

Explosion-relief systems, explosion containment, and explosion suppression should not be used for cases where detonation is considered a credible risk. In such cases, the explosion hazard should be mitigated by preventing the formation of mixtures that could detonate.

Explosion prevention measures, such as inert gas purging, in conjunction with suitable administrative controls can be considered in place of explosion-relief systems for equipment in which internal explosions are possible only as a result of air contamination during start-up or shutdown activities.

4.4.10.2 Transient Pressure Surges

4.4.10.2.1 Water Hammer

The probability of hydraulic shock waves, known as water hammer, occurring in any liquid-filled system should be carefully evaluated. Water hammer is a type of overpressure that cannot be controlled by typical PRVs, since the response time of the valves can be too slow. The oscillating pressure peaks, measured in milliseconds, can raise the normal operating pressure by many times. These pressure waves damage the pressure vessels and piping where proper safeguards have not been incorporated. Water hammer is frequently avoided by limiting the speed at which valves can be closed in long pipelines. Where water hammer can occur, the use of pulsation dampeners, special bladder-type accumulators, or surge valves should be considered, contingent on proper analysis.

4.4.10.2.2 Steam Hammer

An oscillating peak-pressure surge, called steam hammer, can occur in piping that contains compressible fluids. The most common occurrence is generally initiated by rapid valve closure. This oscillating pressure surge occurs in milliseconds, with a possible pressure rise in the normal operating pressure by many times, resulting in vibration and violent movement of piping and possible rupture of equipment. PRVs cannot effectively be used as a protective device because of their slow response time. Avoiding the use of quick-closing valves can prevent steam hammer.

4.4.10.2.3 Condensate-induced Hammer

Isolation of a steam bubble by cold condensate can lead to the eventual rapid collapse of the bubble and catastrophic damage to steam pipework. Proper design and operation of the process system are essential in attempts to eliminate this possibility [e.g. by the use of drains, steam traps, appropriate pipe slope, training, and careful management of change (MOC)].

The hazard is particularly acute during turnaround and maintenance activities where dead-legs that trap a steam bubble can be inadvertently created. PRDs cannot effectively be used as a protective device.

4.4.11 Chemical Reaction

4.4.11.1 General

In some reactions and processes, loss of process control can result in a significant change in temperature and/or pressure. The result can exceed the intended limits of the materials selected. Thus, where cryogenic fluids are being processed, a reduction in pressure could lower the temperature of the fluids to a level below the minimum allowable design temperature of the equipment, with the attendant risk of a low-temperature brittle failure. For exothermic reactions (e.g. decompositions, acid dilutions, polymerizations), excessive temperatures and/or pressures associated with runaway reactions can reduce the allowable stress levels below the design point, or increase the pressure above the MAWP. Where normal PRDs cannot protect against these situations, controls are necessary to warn of changes outside the intended temperature/pressure limits to provide corrective action (see 4.4.8 and 4.4.9). The potential for a chemical reaction in conjunction with the other overpressure scenarios in 4.4 should be considered when appropriate.

The methodology for determining the appropriate size of an emergency vent system for chemical reactions has been established by the Design Institute for Emergency Relief Systems (DIERS) [67, 69, 75, 106, 107] and API 520, Part 1.

The DIERS methodology is based on the following:

- a) defining the design basis upset conditions for the reaction system;
- b) characterizing the systems through bench-scale tests simulating the design basis upset conditions;
- c) using vent-sizing equation that accounts for two-phase gas/liquid vent flow.

4.4.11.2 Scenarios

The design basis upset conditions are process specific but generally include one or more of the following:

- a) external fire;
- b) loss of mixing;
- c) loss of cooling;
- d) incorrect charge of reagents.

4.4.11.3 Reaction Rates/Sizing

Reaction rates are rarely known; therefore, bench-scale tests simulating the design basis upset condition are usually required. There are a number of test apparatus available for this purpose. With the information obtained from the bench-scale tests, the system can be characterized by one of the following terms.

- **Tempered:** Tempered systems are those in which the unwanted reaction produces condensable products and the rate of temperature rise is tempered by liquid boiling at system pressure. Typically, tempered systems are liquid-phase reactions in which a reactant (or solvent) is a major portion of the reactor contents.
- **Gassy:** Gassy systems are those in which the unwanted reaction produces noncondensable products and the rate of temperature rise is not tempered by boiling liquid. Gassy systems can be either liquid-phase decompositions or vapor-phase reactions.
- **Hybrid:** Hybrid systems are those in which the rate of temperature rise due to an unwanted reaction can be tempered by liquid boiling at system pressure but can also give rise to the generation of noncondensable gas.

Following characterization of the system, the appropriate vent-sizing equation can be selected. An excellent discussion of these procedures is contained in Grolmes et al. [75]. However, the reader should be cautioned that this is an area with rapidly changing technology and the most current technology should be used, if available.

If the bench-scale simulations indicate the potential for an explosion, the considerations in 4.4.10 should be applied. It can also be prudent to consider housing the reactor in a specially constructed bay to handle potentially explosive reactions or to increase the equipment-design conditions to contain maximum expected temperature and pressure.

Where feasible, a PRD should be used to control overpressure. Where this is infeasible, other design strategies can be employed to control equipment overstressing. These strategies can include using safety systems such as automatic shutdown systems, inhibitor injection, quench, de-inventorying, alternative power supplies, and depressuring. When this approach is taken, the reliability of the protective system(s) should be addressed in a formal risk analysis. This analysis is outside the scope of this standard.

Other forms of reactions that generate heat (dilution of strong acids) should also be evaluated.

4.4.12 Hydraulic Expansion

4.4.12.1 Causes

Hydraulic expansion is the increase in liquid volume caused by an increase in temperature (see Table 2). It can result from several causes, the most common of which are the following.

- a) Piping or vessels are blocked in while they are filled with cold liquid and are subsequently heated by heat tracing, coils, ambient heat gain, or fire.
- b) A heat exchanger is blocked in on the cold side with flow in the hot side.

Caution—Block valves have the potential to leak, thereby admitting either cold or hot fluid into a heat exchanger that is intended to be blocked in, resulting in a potential overpressure.

- c) Piping or vessels are blocked in while they are filled with liquid at near-ambient temperatures and are heated by direct solar radiation.

Table 2—Typical Values of Cubic Expansion Coefficient for Hydrocarbon Liquids and Water

Gravity of Liquid °API	Cubic Expansion Coefficient ^a 1/°C (1/°F)
3 to 34.9	0.00072 (0.0004)
35 to 50.9	0.0009 (0.0005)
51 to 63.9	0.00108 (0.0006)
64 to 78.9	0.00126 (0.0007)
79 to 88.9	0.00144 (0.0008)
89 to 93.9	0.00153 (0.00085)
94 and lighter	0.00162 (0.0009)
Water	0.00018 (0.0001)
^a At 15.6 °C (60 °F); for other temperatures, Equation (5) can be used to estimate the cubical expansion coefficient.	

In certain installations, such as cooling water circuits, the processing scheme, equipment arrangements and methods, and operation procedures make the elimination of the hydraulic expansion as a cause of overpressure feasible. Typical of such conditions are multiple-shell heat exchangers with at least one cold-fluid block valve of the locked-open design on each shell and a single-shell unit in a given service where the shell can reasonably be expected to remain in service, except on shutdown. In this instance, closing the cold-fluid block valves on the heat exchanger unit should be controlled by administrative procedures and possibly the addition of signs stipulating the proper venting and draining procedures when shutting down and blocking in. The designer is cautioned to review each installation before deciding that administrative controls can be used to eliminate hydraulic expansion as a cause of overpressure. For example, an isolation valve can leak on an isolated heat exchanger or piping containing cold fluid can be blocked-in by a control or shutdown system. If overpressure by thermal hydraulic expansion is deemed credible, a PRD should be designed and installed per the guidelines provided in the following sections.

Caution—When hydraulic expansion overpressure is caused by expansion of sub-cooled liquid, the relief rates are typically relatively small, requiring relatively small PRDs. If the trapped liquid can be heated above its bubble point temperature at the relief pressure, vaporization can occur when the fluid is still contained causing potential loss of containment with possible boiling liquid expanding vapor explosion (BLEVE) unless a significantly larger PRD is installed to protect the equipment from overpressure. See 4.4.13.2.5.3 for guidance.

Caution—In some cases, the contained fluid can be heated above its SLT. In such cases, experience has shown that equipment failure due to thermal hydraulic expansion can result in a BLEVE instead of a minor flange release. See 4.4.6 for a discussion of the SLT.

4.4.12.2 Sizing and Set Pressure

The required relieving rate is not easy to determine. Since every application is for a relieving liquid, the required relieving rate is small; specifying an oversized device is, therefore, reasonable. A nominal diameter (DN) 20 × DN 25 (NPS ¾ × NPS 1) relief valve is commonly used. If there is reason to believe that this size is not adequate, the procedure in 4.4.12.3 can be applied. If the liquid being relieved is expected to flash or form solids while it passes through the relieving device, the procedure in 4.9.2 is recommended.

Proper selection of the set pressure for these relieving devices should include a study of the design rating of all items included in the blocked-in system. The thermal-relief pressure setting should never be above the maximum pressure permitted by the weakest component in the system being protected. However, the PRD should be set high enough to open only under hydraulic expansion conditions. If thermal-relief valves discharge into a closed system, the effects of backpressure should be considered.

4.4.12.3 Special Cases

Two general applications for which thermal-relieving devices larger than a DN 20 × DN 25 (NPS 3/4 × NPS 1) valve can be required are long pipelines of large diameter in uninsulated, aboveground installations and large vessels or heat exchangers operating liquid-full. Long pipelines can be blocked in at or below ambient temperature; the effect of solar radiation raises the temperature at a calculable rate. If the total heat transfer rate and thermal expansion coefficient for the fluid are known, a required relieving rate can be calculated. See Parry^[135] for additional information on thermal relief.

If the fluid properties vary significantly with temperature, the worst-case temperature should be used. Alternatively, more sophisticated calculation methods that include temperature-dependent fluid properties can be used to optimize the size of the relief device.

For liquid-full systems, expansion rates for the sizing of relief devices that protect against thermal expansion of the trapped liquids can be approximated using Equation (2) in International System (SI) of units or Equation (3) in US customary (USC) units:

$$q = \frac{\alpha_v \times \phi}{1000d \times c} \quad (2)$$

where

q is the volume flow rate at the relieving conditions, expressed in m³/s;

α_v is the cubic expansion coefficient for the liquid at the relieving conditions, expressed in 1/°C;

NOTE This information is best obtained from the process design data; however, Table 2 shows typical values for hydrocarbon liquids and water at 15.6 °C.

ϕ is the total heat transfer rate, expressed in watts;

NOTE For heat exchangers, this can be taken as the maximum heat exchanger duty during operation.

d is the relative density referred to water ($d = 1.00$ at 15.6 °C), dimensionless;

NOTE Compressibility of the liquid is usually ignored.

c is the specific heat capacity of the trapped fluid, expressed in J/kg·K;

1000 is the density of water at 15.6 °C, expressed in kg/m³.

$$q = \frac{\alpha_v \times \phi}{500d \times c} \quad (3)$$

where

q is the volume flow rate at the relieving conditions, expressed in US gallons per minute;

α_v is the cubic expansion coefficient for the liquid at the relieving conditions, expressed in $1/^\circ\text{F}$;

NOTE This information is best obtained from the process design data; however, Table 2 shows typical values for hydrocarbon liquids and water at 60°F .

ϕ is the total heat transfer rate, expressed in Btu/h;

NOTE For heat exchangers, this can be taken as the maximum heat exchanger duty during operation.

d is the relative density referred to water ($d = 1.00$ at 60°F), dimensionless;

NOTE Compressibility of the liquid is usually ignored.

c is the specific heat capacity of the trapped fluid, expressed in $\text{Btu/lb}\cdot^\circ\text{F}$;

$500 = 0.13368 \text{ ft}^3/\text{gallon} \times 60 \text{ min/h} \times 62.366 \text{ lb/ft}^3$ (water density at 60°F).

This calculation method provides only short-term protection in some cases. If the blocked-in liquid has a vapor pressure higher than the relief design pressure, then the PRD should be capable of handling the vapor-generation rate. If discovery and correction before liquid boiling is expected, then it is not necessary to account for vaporization in sizing the PRD.

4.4.12.4 Piping

4.4.12.4.1 Theoretical Background for Rigorous Calculations

Where the system under consideration for thermal relief consists of piping only (does not contain pressure vessels or heat exchangers), a PRD might not be required to protect piping from thermal expansion if any of the following apply:

- a) the piping always contains a pocket of noncondensing vapor, such that it can never become liquid-full; or

Caution—Small vapor or gas pockets can disappear upon heating due to compression and/or solubilization. In contrast, multicomponent mixtures with a wide boiling range can always have sufficient vapor present to preclude becoming completely liquid-full. The liquid-volume change upon solar heating, heat tracing, heating to ambient temperature, or heat from another source should be estimated to determine if the volume of the vapor pocket is sufficient for liquid expansion.

- b) the piping is in continuous use (i.e. not batch or semicontinuous use) and drained after being blocked in using well supervised procedures or permits; or
- c) the fluid temperature is greater than the maximum temperature expected from solar heating [usually approximately 60°C to 70°C (approximately 140°F to 160°F)] and there are no other heat sources such as heat tracing (note that fire is generally not considered when evaluating pressure-relief requirements for piping); or
- d) the estimated pressure rise from thermal expansion is within the design limits of the equipment or piping.

The pressure rise due to simultaneous heating of the pipe and blocked-in liquid can be calculated from Equation (4) [Karcher ^[97] and the Center for Chemical Process Safety (CCPS) ^[42]]:

$$p_2 = p_1 + \frac{(T_2 - T_1)(\alpha_v - 3\alpha_l) - \left(\frac{q_{ll} \times t}{V}\right)}{\chi + \left(\frac{d}{2E \times \delta_w}\right)(2.5 - 2\mu)} \quad (4)$$

where

- p_2 is the final gauge pressure of blocked-in, liquid-full equipment, expressed in kPa (psig);
- p_1 is the initial gauge pressure of blocked-in, liquid-full equipment, expressed in kPa (psig);
- T_2 is the final temperature of blocked-in, liquid-full equipment, expressed in °C (°F);
- T_1 is the initial temperature of blocked-in, liquid-full equipment, expressed in °C (°F);
- α_v is the cubic expansion coefficient of the liquid, expressed in 1/°C (1/°F);
- α_l is the linear expansion coefficient of metal wall, expressed in 1/°C (1/°F);
- χ is the isothermal compressibility coefficient of the liquid, expressed in 1/kPa (1/psi);
- d is the internal pipe diameter, expressed in m (in.);
- E is the modulus of elasticity for the metal wall at T_2 , expressed in kPa (psi);
- δ_w is the metal wall thickness, expressed in m (in.);
- μ is Poisson's ratio, usually 0.3;
- q_{ll} is the liquid leakage rate across the block valve seat (usually taken as 0), expressed in m³/s (in.³/s);
- t is the elapsed time for leakage, expressed in seconds;
- V is the pipe volume, expressed in m³ (in.³).

Selected data for α_l and E are given in Table 3. See *Perry's Chemical Engineers' Handbook* ^[136] for data on other materials.

Table 3—Values of Linear Expansion Coefficient, α_l , and Modulus of Elasticity, E

Metal	α_l 1/°C (1/°F)	E kPa (psi)
Carbon steel (1020)	1.21×10^{-5} (6.7×10^{-6})	207×10^6 (30×10^6)
304 stainless steel	1.73×10^{-5} (9.6×10^{-6})	193×10^6 (28×10^6)
316 stainless steel	1.60×10^{-5} (8.9×10^{-6})	193×10^6 (28×10^6)
Alloy 600	1.1×10^{-5} to 1.66×10^{-5} (6.1×10^{-6} to 9.2×10^{-6})	172×10^6 to 221×10^6 (25×10^6 to 32×10^6)
Nickel-copper alloy	1.01×10^{-5} to 1.42×10^{-5} (5.6×10^{-6} to 7.9×10^{-6})	169×10^6 to 213×10^6 (24.5×10^6 to 30.9×10^6)

Where data are unavailable, Equation (5) and Equation (6) can be used to estimate, respectively, the isothermal compressibility coefficient, χ (see *Lange's Handbook of Chemistry*, Twelfth Edition [51], pp. 10–122) and the cubic expansion coefficient, α_v (see *Perry's Chemical Engineers' Handbook* [136], Fifth Edition, pp. 3–227):

$$\alpha_v = \frac{\rho_1^2 - \rho_2^2}{2(T_2 - T_1)\rho_1 \times \rho_2} \quad (5)$$

where

- α_v is the cubic expansion coefficient, expressed in $1/^\circ\text{C}$ ($1/^\circ\text{F}$);
- ρ_1 is the density of liquid at temperature T_1 , expressed in kg/m^3 (lb/ft^3);
- ρ_2 is the density of liquid at temperature T_2 , expressed in kg/m^3 (lb/ft^3);
- T_1 is the temperature at the beginning of the interval, expressed in $^\circ\text{C}$ ($^\circ\text{F}$);
- T_2 is the temperature at the end of the interval, expressed in $^\circ\text{C}$ ($^\circ\text{F}$).

$$\chi = \frac{1}{v_1} \frac{(v_1 - v_2)}{(p_2 - p_1)} \quad (6)$$

where

- χ is the isothermal compressibility coefficient, expressed in $1/\text{kPa}$ ($1/\text{psi}$);
- v_1 is the specific volume of liquid at pressure p_1 , expressed in m^3/kg (ft^3/lb);
- v_2 is the specific volume of liquid at pressure p_2 , expressed in m^3/kg (ft^3/lb);
- p_1 is the absolute pressure at the beginning of the interval, expressed in kPa (psia);
- p_2 is the absolute pressure at the end of the interval, expressed in kPa (psia).

4.4.12.4.2 Other Aspects

For the thermal expansion scenario, no credit should be taken for reverse flow back through a check valve (i.e. assume the check valve holds) or a closed block valve. Alternatives are to drill a small [e.g. 6 mm ($1/4$ in.)] hole in the block valve gate, install a small bypass around the block valve with appropriate administrative controls, or install a three-way valve to ensure that the piping system cannot be completely blocked in.

If the above criteria cannot be met for a piping system, then the following factors should be evaluated for the fluid and the piping system, when determining if a thermal-relief valve is warranted to protect the system.

- a) Length and size of the piping system—The quantity of fluid that can be released is dependent on the length and size of the piping system.
- b) Hazardous and flammable nature of the fluid—For a hazardous or highly flammable fluid, even a small amount of leakage might not be allowable.
- c) Location of the piping system—Leakage into a confined area can be especially hazardous depending on the fluid properties.
- d) Vapor pressure of the fluid at the heated temperature—Fluids above their atmospheric boiling point continue to release material as vapor through a leak until the fluid temperature cools to the boiling point.
- e) Adequacy of procedures and administrative controls to avoid blocking in.

4.4.13 Fires

4.4.13.1 General

Fire exposure on equipment can result in overpressure due to vapor generation (boiling of liquid contents or decomposition reaction) and/or fluid expansion. Fire exposure can also cause overheating of the vessel walls resulting in a reduction in material strength. For the purpose of this standard, fires are characterized as open pool fires (see 4.4.13.2), confined pool fires (see 4.4.13.3), or jet fires (see 4.4.13.4). A pool fire typically results from an ignited liquid spill, whereas a jet fire results from an ignited pressurized leak. The heat flux from jet fires is very high and localized, whereas the heat flux from pool fires is lower and not localized. Confined pool fires are those that occur inside a structure or are confined by embankments thereby causing higher heat fluxes than open pool fires in some cases. Fire heat intensities can vary dramatically depending on fuel, ventilation, release rate, and other factors. Typical ranges of heat intensities are as follows.

- Open pool fire: 50 kW/m² to 150 kW/m² (15,850 Btu/h·ft² to 47,550 Btu/h·ft²).
- Confined pool fire: 100 kW/m² to 250 kW/m² (31,700 Btu/h·ft² to 79,250 Btu/h·ft²).
- Jet fire: 100 kW/m² to 400 kW/m² (31,700 Btu/h·ft² to 126,800 Btu/h·ft²).

Annex A describes fire scenarios in detail and provides guidance in modeling fires. This information may be useful for assessing fires that may have less intensity than that premised in 4.4.13.2 or may be used for purposes other than PRD sizing (e.g. for emergency depressuring). Caution should be taken when using the Annex A method for sizing PRDs because it may underestimate the fire heat input.

4.4.13.2 Open Pool Fires

4.4.13.2.1 General

For the purposes of PRD sizing for equipment within the scope of this standard, the design fire scenario has been and continues to be an open pool fire. The recommended method was empirically derived to size PRDs for open pool fires involving hydrocarbons in a refinery environment that is typical for the facilities within the scope of this standard. The method is supported by full-scale test data (see A.2 and C.6 for details).

Both API 521 and API 2000 ^[11] use open pool fires as the basis for sizing PRDs for the fire case. It is important to apply the appropriate standard when sizing for fire relief because there are differences in assumptions for the fire pool fire intensity, exposed area, and other factors specified in those standards. API 2000 is limited to aboveground liquid-petroleum or petroleum-products storage tanks and aboveground and underground refrigerated storage tanks designed for operation at gauge pressures from vacuum through 103.4 kPa (15 psi).

An open pool fire can affect multiple vessels simultaneously. See 5.3.2 for a discussion.

If the open pool fire involves other types of fuels (e.g. alcohols) that have significantly different radiative fluxes than fuels similar to gasoline, diesel, liquefied petroleum gas (LPG), etc., the method in Annex A can be used with adjusted variables (see Annex A for details).

Relieving temperatures are often above the design temperature of the equipment being protected. If the elevated temperature is likely to cause vessel rupture, additional protective measures should be considered (see 4.4.13.2.6 and 4.4.13.2.7).

4.4.13.2.2 Effect of Fire on the Wetted Surface of a Vessel

To determine vapor generation, it is necessary to recognize only that portion of the vessel that is wetted by its internal liquid and is equal to or less than 7.6 m (25 ft) above the source of flame.

NOTE Hydrocarbon fires can exceed 7.6 m (25 ft) in height; however, experience has shown that it is necessary only to size relief devices on the basis of the averaged heat input up to a height of 7.6 m (25 ft) above the base of a pool fire.

The term “base of a pool fire” usually refers to ground level but could be at any level at which a substantial spill or pool fire could be sustained. Various classes of vessels are operated only partially full. Table 4 gives recommended portions of liquid inventory for use in calculations. Wetted surfaces higher than 7.6 m (25 ft) are normally excluded because pool fire flames are not likely to impinge for long durations above this height. Also, vessel heads protected by support skirts with limited ventilation are normally not included when determining wetted area. The user shall specify whether to include the wetted surface area of connected piping in the wetted-area calculation.

The wetted area for spheres includes all area up to the maximum diameter. Table 4 recommends the wetted surface area for spheres be based on “the maximum horizontal diameter or up to the height of 7.6 m (25 ft), whichever is greater.” Hence, as a minimum, the wetted surface area of the entire bottom hemisphere shall be used even when the sphere “equator” exceeds 7.6 m (25 ft) in height. The criterion is supported by previous incidents and tests that have shown that pool fire flames can follow the underside profile of spheres resulting in the entire bottom hemisphere being exposed to a high fire heat load.

Table 4—Effects of Fire on the Wetted Surfaces of a Vessel

Class of Vessel	Portion of Liquid Inventory	Remarks
Liquid-full, such as treaters	All up to the height of 7.6 m (25 ft)	—
Surge drums, knockout drums, process vessels	Normal operating level up to the height of 7.6 m (25 ft)	—
Fractionating columns	Normal level in bottom plus liquid holdup from all trays dumped to the normal level in the column bottom; total wetted surface up to the height of 7.6 m (25 ft)	Level in reboiler is to be included if the reboiler is an integral part of the column
Working storage	Maximum inventory level up to the height of 7.6 m (25 ft) (portions of the wetted area in contact with foundations or the ground are normally excluded)	For low-pressure [i.e. <103 kPa (15 psig)] storage tanks and process tanks (see API 2000 ^[11])
Spheres and spheroids	Up to the maximum horizontal diameter or up to the height of 7.6 m (25 ft), whichever is greater	—

4.4.13.2.3 Effect of Fire on the Unwetted Surface of a Vessel

Unwetted wall vessels are those that have no liquid in contact with the internal vessel walls (e.g. internal walls are exposed only to a gas, vapor, or supercritical fluid or they are internally insulated regardless of the contained fluids). These include vessels that contain separate liquid and vapor phases under normal conditions but become single phase (above the critical) at relieving conditions.

Vessels can be designed to have internal insulation (e.g. refractory) and such areas may be considered unwetted. If, however, a vessel can become insulated by the deposition of coke or other materials, the vessel wall shall still be considered wetted for fire-relief sizing (without credit for any insulating effects) but additional protection should be considered (see 4.4.13.2.6 and 4.4.13.2.7).

A characteristic of a vessel with an unwetted internal wall is that heat flow from the wall to the contained fluid is low as a result of the heat transfer resistance of the contained fluid or any internal insulating material. Heat input from a fire to the bare outside surface of an unwetted or internally insulated vessel can, in time, be sufficient to heat the vessel wall to a temperature high enough to rupture the vessel. Figure 1 and Figure 2 indicate how quickly an unwetted bare vessel wall can be heated to rupture conditions. Figure 1 illustrates the rise in temperature that occurs with time in the unwetted plates of various thicknesses exposed to open fire. For example, an unwetted steel plate 25 mm (1 in.) thick takes about 12 min to reach 593 °C (1100 °F) and 17 min to reach 704 °C (1300 °F) when the plate is exposed to a typical open fire. Recent calculations indicate that the absorbed heat flux from the fire is in the range of approximately 80 kW/m² (25,200 Btu/h·ft²) for the observed curve to 100 kW/m² (31,500 Btu/h·ft²) for the calculated curves.

Figure 2 shows the effect of overheating ASTM A515 Grade 70 carbon steel, from data published in Reference [164]. The figure indicates that at a stress of 138 MPa (20,000 psi), an unwetted steel vessel ruptures in about 0.1 h at 649 °C (1200 °F). A source for time-dependent rupture stress for different metals is ASTM Data Series DS 11S1 ^[27], which contains stress rupture and other elevated temperature property data for wrought carbon steel.

This work was performed by the Materials Properties Council but is available through ASTM. Another source is *Guidelines for the Protection of Pressurised Systems Exposed to Fire* ^[148].

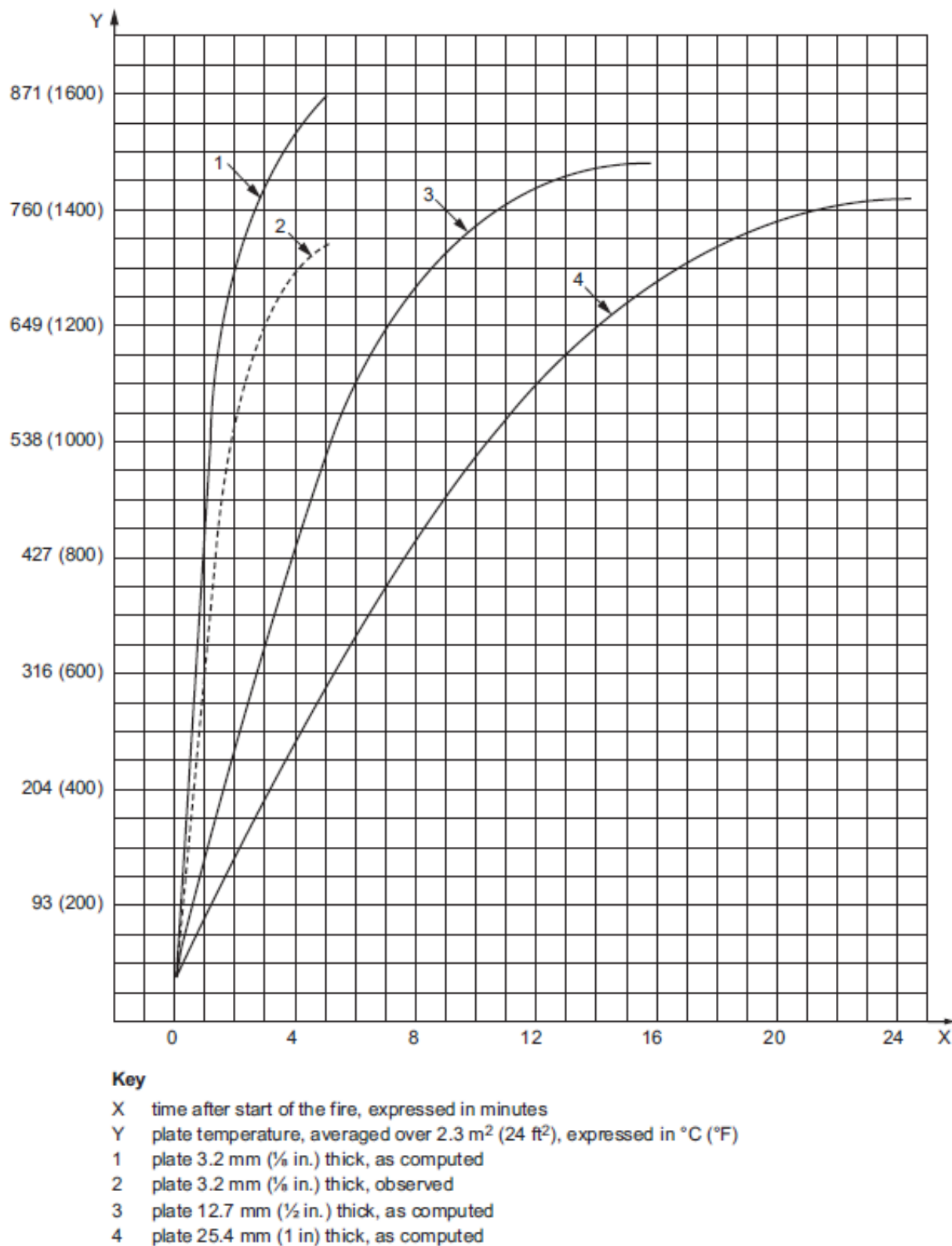


Figure 1—Average Rate of Heating Steel Plates Exposed to Open Gasoline Fire on One Side

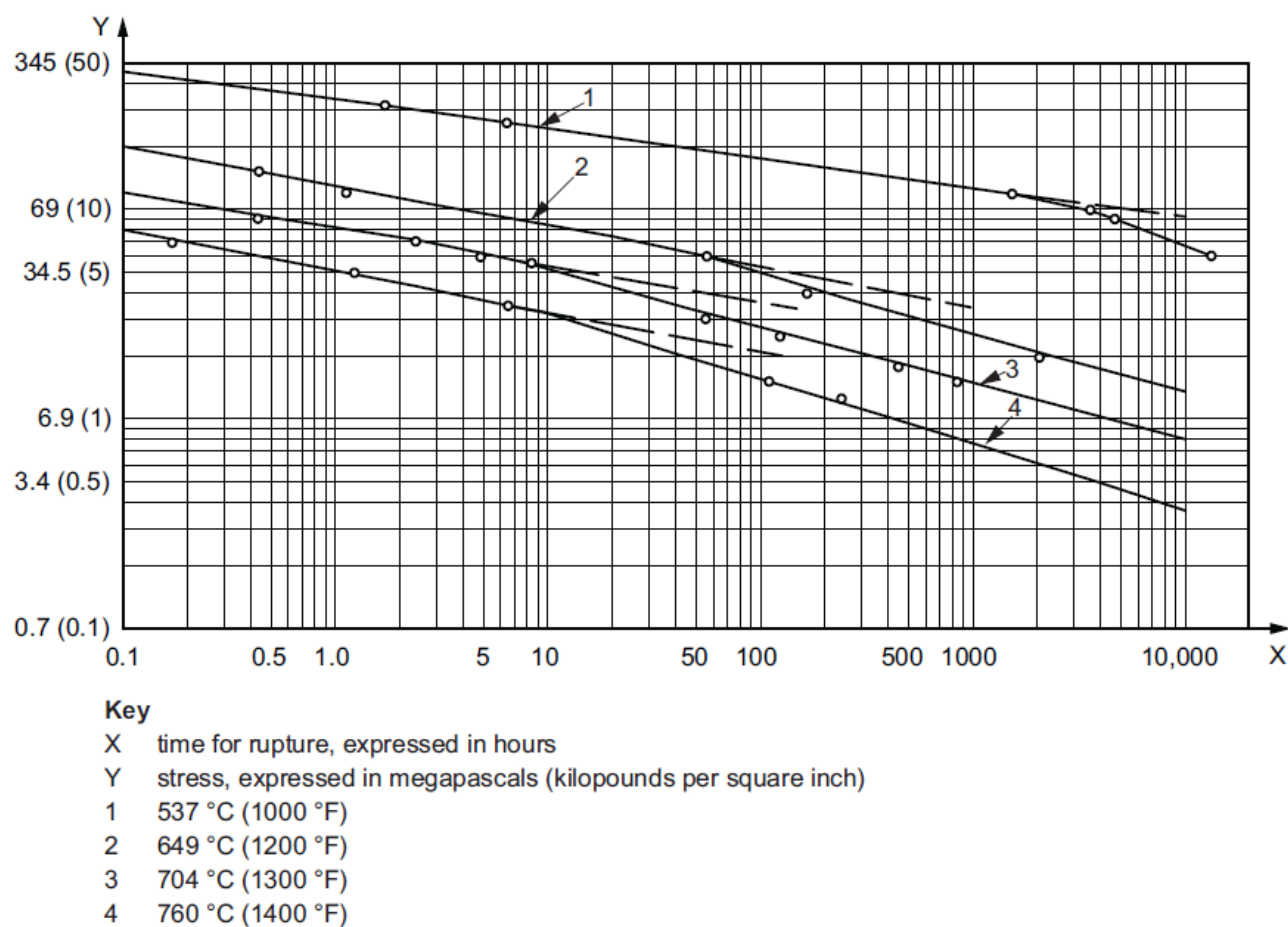


Figure 2—Effect of Overheating Carbon Steel (ASTM A515, Grade 70)

4.4.13.2.4 Fire-relief Loads

4.4.13.2.4.1 General

It is typically assumed that the vessel is isolated during a fire in order to simplify the analysis, although a more detailed analysis can be warranted in certain cases. Crediting for flow paths that remain open during an overpressure event is generally an acceptable practice. However, it should be recognized that operators and/or emergency responders may attempt to isolate certain lines and vessels during a fire condition in order to limit the fire spread and to safely shutdown the unit. There can also be actuated valves that fail in the closed condition when exposed to a fire. It can be difficult to establish with a degree of certainty whether a particular line will indeed remain open under all fire conditions. Further, consideration should be given to the potential that the fire-relief flow in the flow path will overpressure other equipment. Hence, it can be necessary to add the fire-relief load elsewhere. Ultimately, the user shall decide whether a scenario is credible or not.

4.4.13.2.4.2 provides the heat absorption equations for vessels containing liquids, and 4.4.13.2.4.3 provides the equations to determine open-pool-fire-relief requirements for vessels containing only gases/vapors.

Either the vapor thermal-expansion-relief load or the boiling-liquid vaporization-relief load, but not both, should be used. It is a practice that has been used for many years. There are no known experimental studies where separate contributions of vapor thermal expansion versus boiling-liquid vaporization have been determined. When sizing the PRD for fire exposure, the contribution of vaporizing liquid compared with vapor expansion is generally governing unless, for example, the wetted surface has external insulation in accordance with 4.4.13.2.7 and the unwetted surfaces are not insulated.

4.4.13.2.4.2 Heat Absorption Equations for Vessels Containing Liquids

The amount of heat absorbed by a vessel exposed to an open fire is markedly affected by the type of fuel feeding the fire, the degree to which the vessel is enveloped by the flames (a function of vessel size and shape), the environment factor, firefighting, and drainage. Equation (7) is used to evaluate these conditions if there are prompt firefighting efforts and drainage of flammable materials away from the vessels:

$$Q = C_1 \times F \times A_{ws}^{0.82} \quad (7)$$

where

Q is the total heat absorption (input) to the wetted surface, expressed in W (Btu/h);

C_1 is a constant [= 43,200 in SI units (21,000 in USC units)];

F is an environment factor (see Table 5);

A_{ws} is the total wetted surface, expressed in m² (ft²).

NOTE 1 See 4.4.13.2.2 and Table 4.

NOTE 2 The expression, $A_{ws}^{0.82}$, is the area exposure factor or ratio. This ratio recognizes that large vessels are less likely than small ones to be completely exposed to the flame of an open fire.

Table 5—Environment Factor

Type of Equipment		Environment Factor F
Bare vessel		1.0 ^e
Insulated vessel ^{a,b} with insulation conductance values (i.e. insulation thermal conductivity divided by thickness) for fire exposure conditions in W/m ² ·K (Btu/h·ft ² ·°F)	22.71 (4)	0.3
	11.36 (2)	0.15
	5.68 (1)	0.075
	3.80 (0.67)	0.05
	2.84 (0.5)	0.0376
	2.27 (0.4)	0.03
	1.87 (0.33)	0.026
Water application facilities, on bare vessel ^c		1.0 ^e
Depressurizing and emptying facilities ^d		1.0 ^e
Earth-covered storage		0.03
Below-grade storage		0.00

^a These suggested values for the conditions assumed in 4.4.13.2.4. If these conditions do not exist, engineering judgment should be exercised either in selecting a higher factor or in providing means of protecting vessels from fire exposure as suggested in 4.4.13.2.6 and 4.4.13.2.7.

^b Insulation should resist being dislodged by fire hose streams (see 4.4.13.2.7.2). For the examples, a temperature difference of 871 °C (1600 °F) was used. These conductance values are computed from Equation (17) or Equation (18) and are based on insulation having thermal conductivity of 0.58 W/m·K (4 Btu·in./h·ft²·°F) at 538 °C (1000 °F) and correspond to various thicknesses of insulation between 25.4 mm (1 in.) and 304.8 mm (12 in.). See Equation (17) or Equation (18) to determine the environment factor, F .

^c See 4.4.13.2.6.2.

^d See 4.6 and Annex A.

^e The environment factor, F , in Equation (7) and Equation (8) does not apply to uninsulated vessels. The environment factor should be replaced by 1.0 when calculating heat input to uninsulated vessels.