BRITISH STANDARD

BS EN ISO 23251:2007 +A1:2008

Incorporating corrigendum August 2007

Petroleum, petrochemical and natural gas industries — Pressure-relieving and depressuring systems

ICS 75.180.20

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British Standards

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National foreword

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The start and finish of text introduced or altered by amendment is indicated in the text by tags. Tags indicating changes to ISO text carry the number of the ISO amendment. For example, text altered by ISO amendment 1 is indicated by [A1⟩ ⟨A1].

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A list of organizations represented on this subcommittee can be obtained on request to its secretary.

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EUROPEAN STANDARD

NORME EUROPÉENNE

EUROPÄISCHE NORM

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English Version

Petroleum, petrochemical and natural gas industries - Pressure-relieving and depressuring systems (ISO 23251:2006)

Industries du pétrole, de la pétrochimie et du gaz naturel - Systèmes de dépressurisation et de protection contre les surpressions (ISO 23251:2006)

Erdöl-, petrochemische und Erdgasindustrie - Druckentlastungs- und Druckausgleichssysteme (ISO 23251:2006)

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Foreword

The text of ISO 23251:2006 has been prepared by Technical Committee ISO/TC 67 "Materials, equipment and offshore structures for petroleum and natural gas industries" of the International Organization for Standardization (ISO) and has been taken over as EN ISO 23251:2007 by Technical Committee CEN/TC 12 "Materials, equipment and offshore structures for petroleum, petrochemical and natural gas industries", the secretariat of which is held by AFNOR.

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The text of ISO 23251:2006 has been approved by CEN as EN ISO 23251:2007 without any modifications.

Foreword to amendment A1

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Introduction

This International Standard is based on the draft 5th edition of API RP 521, with the intent that the 6th edition of API RP 521 will be identical to this International Standard.

The portions of this International Standard dealing with flares and flare systems are an adjunct to API Std 537 [10], which addresses mechanical 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 data sheets, which can be found in of API Std 537, Appendix A. The use of these data sheets is both recommended and encouraged as a concise, uniform means of recording and communicating design information.

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Petroleum, petrochemical and natural gas industries — Pressure-relieving and depressuring systems

1 Scope

This International Standard is applicable to pressure-relieving and vapour-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 International Standard is intended to supplement the practices set forth in ISO 4126 or API RP 520-I for establishing a basis of design.

This International Standard specifies requirements and gives guidelines for examining the principal causes of overpressure; and determining individual relieving rates; and selecting and designing disposal systems, including such component parts as piping, vessels, flares, and vent stacks. This International Standard does not apply to direct-fired steam boilers.

Piping information pertinent to pressure-relieving systems is presented in 7.3.1.

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.

ISO 4126 (all parts), *Safety devices for protection against excessive pressure*

API RP 520-I:2000, *Sizing, Selection and Installation of Pressure-Relieving Devices in Refineries — Part I: Sizing and Selection* [1)]

3 Terms and definitions

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

3.1
accumulation
pressure increase over the maximum allowable working pressure of the vessel allowed 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.

1) American Petroleum Institute, 1220 L Street, N.W., Washington, D.C., 20005-4070, USA.

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3.2
administrative controls
procedures intended to ensure that personnel actions do not compromise the overpressure protection of the equipment

3.3
assist gas
combustible 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

3.4
atmospheric discharge
release of vapours and gases from pressure-relieving and depressuring devices to the atmosphere

3.5
back pressure
pressure that exists at the outlet of a pressure-relief device as a result of the pressure in the discharge system

NOTE The back pressure is the sum of the superimposed and built-up back pressures.

3.6
balanced pressure-relief valve
spring-loaded pressure-relief valve that incorporates a bellows or other means for minimizing the effect of back pressure on the operational characteristics of the valve

3.7
blowdown
depressurization of a plant or part of a plant, and equipment

NOTE Not to be confused with the difference between the set pressure and the closing pressure of a pressure-relief valve.

3.8
blow-off
loss of a stable flame where the flame is lifted above the burner, occurring if the fuel velocity exceeds the flame velocity

3.9
breaking-pin device
pressure-relief device actuated by static differential or static inlet pressure and designed to function by the breakage of a load-carrying section of a pin that supports a pressure-containing member

3.10
buckling pin device
pressure-relief device actuated by static differential or static inlet pressure and designed to function by the buckling of an axially-loaded compressive pin that supports a pressure-containing member

3.11
built-up back pressure
increase in pressure at the outlet of a pressure-relief device that develops as a result of flow after the pressure-relief device opens

3.12
buoyancy seal
dry vapour seal that minimizes the amount of purge gas needed to protect against air infiltration

NOTE The buoyancy seal functions by trapping a volume of light gas in an internal inverted compartment; this prevents air from displacing buoyant light gas in the flare.

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3.13
burnback
internal burning within the flare tip

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

3.14
burning velocity
flame velocity
speed at which a flame front travels into an unburned combustible mixture

3.15
burn-pit flare
open excavation, normally equipped with a horizontal flare burner that can handle liquid as well as vapour hydrocarbons

3.16
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.

3.17
closed disposal system
disposal system capable of containing pressures that are different from atmospheric pressure

3.18
cold differential test pressure
CDTP
pressure at which a pressure-relief valve is adjusted to open on the test stand

NOTE The cold differential test pressure includes corrections for the service conditions of back pressure or temperature or both.

3.19
combustion air
air required to combust the flare gases

3.20
conventional pressure-relief valve
spring-loaded pressure-relief valve whose operational characteristics are directly affected by changes in the back pressure

3.21
corrected hydrotest pressure
hydrostatic test pressure multiplied by the ratio of stress value at design temperature to the stress value at test temperature

NOTE See 4.3.2.

3.22
deflagration
explosion in which the flame-front of a combustible medium is advancing at less than the speed of sound

cf. **detonation** (3.25)

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3.23
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 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).

3.24
destruction efficiency
mass fraction of the fluid vapour that can be oxidized or partially oxidized

NOTE For a hydrocarbon, this is the mass fraction of carbon in the fluid vapour that oxidizes to CO or CO_2.

3.25
detonation
explosion in which the flame-front of a combustible medium is advancing at or above the speed of sound

cf. **deflagration** (3.22)

3.26
dispersion
dilution of a vent stream or products of combustion as the fluids move through the atmosphere

3.27
elevated flare
flare where the burner is raised high above ground level to reduce radiation intensity and to aid in dispersion

3.28
enclosed flare
enclosure with one or more burners arranged in such a manner that the flame is not directly visible

3.29
enrichment
process of adding assist gas to the relief gas

3.30
flame-retention device
device used to prevent flame blow off from a flare burner

3.31
flare
device or system used to safely dispose of relief gases in an environmentally compliant manner through the use of combustion

3.32
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

3.33
flare header
piping system that collects and delivers the relief gases to the flare

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3.34
flashback
phenomenon occurring in a flammable mixture of air and gas when the local velocity of the combustible mixture becomes less than the flame velocity, causing the flame to travel back to the point of mixture

3.35
ground flare
non-elevated flare

NOTE A ground flare is normally an enclosed flare but can also be a ground multi-burner flare or a burnpit.

3.36
heat release
total heat liberated by combustion of the relief gases based on the lower heating value

3.37
huddling chamber
annular chamber located downstream of the seat of a pressure-relief valve, which assists the valve to lift

3.38
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

3.39
jet fire
fire created when a leak from a pressurized system ignites and forms a burning jet

NOTE A jet fire can impinge on other equipment, causing damage.

3.40
knockout drum
vessel in the effluent handling system designed to remove and store liquids

3.41
lateral
section of pipe from outlet flange(s) of single-source relief device(s) downstream of a header connection where relief devices from 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.

3.42
lift
actual travel of the disc from the closed position when a valve is relieving

3.43
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 back pressure for the flare header

3.44
Mach number
ratio of a fluid's velocity, measured relative to some obstacle or geometric figure, divided by the speed at which sound waves propagate through the fluid

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3.45
manifold
piping system for the collection and/or distribution of a fluid to or from multiple flow paths

3.46
marked burst pressure
rated burst pressure
⟨rupture disk⟩ burst pressure, established by tests for the specified temperature and marked on the disk tag by the manufacturer

NOTE The marked burst pressure can be any pressure within the manufacturing design range unless otherwise specified by the customer. The marked burst pressure is applied to all of the rupture disks of the same lot.

3.47
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

cf. **design pressure** (3.23)

NOTE The MAWP is the least of the values for the internal or external pressure as determined by the vessel design rules for each element of the vessel using actual nominal thickness, exclusive of additional metal thickness allowed for corrosion and loadings other than pressure. The MAWP is the basis for the pressure setting of the pressure-relief devices that protect the vessel.

3.48
non-condensable gas
gas or vapour that remains in the gaseous state at the temperature and pressure expected

3.49
operating pressure
pressure the process system experiences during normal operation, including normal variations

3.50
overpressure
⟨general⟩ condition where the MAWP, or other specified pressure, is exceeded

⟨relieving device⟩ pressure increase over the set pressure of a relieving device

NOTE In the latter context, overpressure is the same as **accumulation** (3.1) only when the relieving device is set to open at the MAWP of the vessel.

3.51
pilot burner
small, continuously operating burner that provides ignition energy to light the flared gases

3.52
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)

3.53
pin device
non-reclosing 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 instantly moves to the fully open position

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3.54
pool fire
burning pool of liquid

3.55
pressure-design code
standard to which the equipment is designed and constructed

EXAMPLE ASME Section VIII, Division 1 [20].

3.56
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 In ISO 4126-1, this is termed a safety valve.

3.57
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

cf. **storage tank** (3.74)

NOTE 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.

3.58
purge gas
fuel gas or non-condensable inert gas added to the flare header to mitigate air ingress and burnback

3.59
quenching
cooling of a fluid by mixing it with another fluid of a lower temperature

3.60
radiation intensity
local radiant heat transfer rate from the flare flame, usually considered at grade level

3.61
rated relieving capacity
relieving capacity used as the basis for the application 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.

3.62
relief gas
flared gas
waste gas
waste vapour
gas or vapour vented or relieved into a flare header for conveyance to a flare

3.63
relief valve
spring-loaded pressure-relief valve actuated by the static pressure upstream of the valve, due to which the valve normally opens in proportion to the pressure increase over the opening pressure

NOTE A relief valve is normally used with incompressible fluids.

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3.64
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.

3.65
rupture-disk device
non-reclosing 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 1 A rupture disk device includes a rupture disk and a rupture disk holder.

NOTE 2 In ISO 4126-2, this is termed a bursting-disc safety device.

3.66
safety instrumented system
SIS
emergency shutdown system
ESD, ESS
high-integrity protection system
HIPS
high-integrity pressure-protection system
HIPPS
safety-shutdown system
SSD
safety-interlock system
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 violated

3.67
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.

3.68
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

3.69
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 1 A safety valve is normally used with compressible fluids.

NOTE 2 This definition is different than that in ISO 4126-1; see 3.56.

3.70
set pressure
inlet gauge pressure at which a pressure-relief device is set to open under service conditions

3.71
shear pin device
non-reclosing pressure-relief device actuated by static differential or static inlet pressure and designed to function by the shearing of a load-carrying member that supports a pressure-containing member

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3.72
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

3.73
stoichiometric air
chemically correct ratio of fuel to air capable of perfect combustion with no unused fuel or air

3.74
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

cf. **process tank** (3.57)

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

3.75
superimposed back pressure
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.

3.76
vapour depressuring system
protective arrangement of valves and piping intended to provide for rapid reduction of pressure in equipment by releasing vapours

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

3.77
velocity seal
dry vapour seal that minimizes the required purge gas needed to protect against air infiltration into the flare burner exit

3.78
vent header
piping system that collects and delivers the relief gases to the vent stack

3.79
vent stack
elevated vertical termination of a disposal system that discharges vapours into the atmosphere without combustion or conversion of the relieved fluid

3.80
vessel
container or structural envelope in which materials are processed, treated or stored

EXAMPLES Pressure vessels, reactor vessels and storage vessels (tanks).

3.81
windshield
device used to protect the outside of a flare burner from direct flame impingement

NOTE The windshield is so named because external flame impingement occurs on the downwind side of an elevated flare burner.

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4 Causes of overpressure

4.1 General

Clause 4 discusses the principal causes of overpressure and offers guidance in plant design to minimize the effects of these causes. 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 application of the principles outlined in Clause 4 are 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 International 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. Pressure-relieving devices 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 Clause 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 Double jeopardy

The causes of overpressure are considered to be unrelated 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 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.

This International 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. Since 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.

4.2.2 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 favourably 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.

4.2.3 Operator error

Operator error is considered a potential source of overpressure.

4.2.4 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 pressure-relieving devices as protection against single-jeopardy overpressure scenarios. There can be circumstances, however, where the use of pressure-relief devices is impractical and

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reliance on instrumented safeguards is needed. Where this is the case, if permitted by local regulations, a pressure-relieving device might not be required.

NOTE See ASME Code Case 2211 [129].

The design shall comply with the local regulations and the owner's risk tolerance criteria, whichever is more restrictive. If these risk tolerance criteria are not available, then, as a minimum, the overall system performance including instrumented safeguards should provide safety-integrity-level 3 (SIL-3) performance. Guidance on the application of safety instrumented systems is given in Annex E.

Although favourable 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 blowdown header, flare, and flare tip, favourable 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 favourable response of instrument systems should consider the number and reliability of applicable instrument systems. See 7.1 for more details on sizing disposal systems.

4.3 Potentials for overpressure

4.3.1 General

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

a) the normal operating pressure at operating temperatures;

b) the effect of any combination of process upsets that are likely to occur;

c) the differential between the operating, and set pressures of the pressure-relieving device;

d) the effect of any combination of supplemental loadings such as earthquake and wind.

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. The principal causes of overpressure listed in 4.3.2 through 4.3.15 are guides to generally accepted practices. Annex B provides guidance on the use of a common relief device to protect multiple pieces of equipment from overpressure.

4.3.2 Closed outlets on vessels

The inadvertent closure of a manual block valve on the outlet of a pressure vessel while the equipment is on stream can expose the vessel to a pressure that exceeds the maximum allowable working pressure. If closure of an outlet-block valve can result in overpressure, a pressure-relief device is required unless administrative controls are in place. Every valve should be considered as being subject to inadvertent operation. In general, the omission of block valves interposed in vessels in a series can simplify pressure-relieving requirements. If the pressure resulting from the failure of administrative controls can exceed the corrected hydrotest pressure (see 3.21), reliance on administrative controls as the sole means to prevent overpressure might not be appropriate. 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.

For example, an ASTM A 515 [22] Grade 70 carbon steel vessel with a design gauge pressure of 517 kPa (75 psi) and design temperature of 343 °C (650 °F) has an allowable stress of 130 MPa (18 800 psi) at these design conditions. Because the hydrostatic test is often performed at a temperature less than design temperature, the hydrostatic test pressure should be specified to account for the allowable stress differences at the two temperatures by multiplying the design pressure by the ratio of stress at test temperature to the stress at design temperature. At ambient temperature, the allowable stress of ASTM A 515 Grade 70 carbon steel is 138 MPa (20 000 psi). If the pressure-design code requires the hydrostatic test be performed at 130 % of the design pressure, then the hydrostatic test pressure is as follows:

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In SI units:

$517 \times (138/130) \times 1{,}3 = 713$ kPa (gauge)

In USC units:

$75 \times (20/18{,}8) \times 1{,}3 = 103{,}7$ psig

The uncorrected hydrotest gauge pressure is $517 \times 1{,}3 = 672$ kPa ($75 \times 1{,}3 = 97{,}5$ psi). In this example, reliance on administrative controls as the sole means of overpressure protection might not be appropriate if the gauge pressure caused by closure of the outlet valve exceeds 672 kPa (97,5 psi). This assumes the overpressure occurs while the vessel is at design temperature. Within stage 1 and stage 2 creep, short-duration pressure exceedances up to 1,5 times the design pressure at design temperature should not result in damage, provided that there is no significant coincident temperature increase. This is based on allowable values in the creep regime being based on 100 000 h design.

Similarly, the inadvertent closure of a remotely operated valve on the outlet of a pressure vessel while the equipment is on stream can expose the vessel to a pressure that exceeds the maximum allowable working pressure. If closure of a remotely operated outlet valve can result in overpressure, a pressure-relief device is required. Every control valve should be considered as being subject to inadvertent operation.

For determining relief loads, it may be assumed that manual or remotely operated valves that are normally open and functioning at the time of failure and that are not affected by the primary cause of failure remain in operation at their normal operating positions. See 5.10.4 for additional information.

4.3.3 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. This action can require pressure-relieving capacity unless administrative controls, as defined in 3.2, are in place to prevent inadvertent valve opening.

4.3.4 Check-valve leakage or failure

4.3.4.1 Causes of overpressure due to check-valve leakage or failure

A single check valve is not always an effective means for preventing overpressure by reverse flow from a high-pressure source. For example, if a fluid is pumped into a system that contains vapour at significantly higher pressure than the design rating of the equipment upstream of the pump, loss of pumped flow with leakage or latent failure of a check valve in the discharge line results in a reversal of the liquid's flow. When high-pressure fluid enters the low-pressure system, overpressure can result.

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 destroy mechanical equipment, causing loss of containment. If this hazard is of concern, additional means of backflow prevention should be provided.

4.3.4.2 Pressure consideration for single check-valve latent failure

Overpressure protection shall be provided for single check-valve latent failure (e.g. stuck open or broken flapper) where the maximum normal operating pressure of the high-pressure system is greater than the design pressure or MAWP of vessels, equipment and piping in the upstream low-pressure system and there is enough stored energy in the high-pressure system to cause an overpressure in the low-pressure system (e.g. large vapour cap in the high-pressure system).

When sizing a pressure-relief device to prevent exceeding the allowable accumulation of the protected equipment for the latent check-valve failure, the reverse flow rate through a single check valve may be determined using the normal flow characteristics (i.e., forward-flow Cv) of the check valve. If the check valve

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Cv is unavailable, one may conservatively assume that the check valve is not there by taking no credit for its flow resistance.

If the single check valve is inspected and maintained to ensure its reliability and capability to limit reverse flow, the user may determine that the check-valve latent failure is unlikely. In this case, overpressure protection should be provided where the maximum normal operating pressure of the high-pressure system is greater than the upstream equipment's corrected hydrotest pressure (see 3.21 and 4.3.2). The user is cautioned that some systems can have unacceptable risk due to latent failure of the check valve 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 latent failure of the check valve.

4.3.4.3 Pressure consideration for single check-valve leakage

Even properly inspected and maintained check valves might not completely eliminate check-valve seat leakage. Consequently, the user should be aware that isolation of the low-pressure system upstream of the check valves can still result in overpressure. If operator action (e.g. manual isolation from the high-pressure system) is being considered to prevent overpressure due to check-valve leakage, then a layers-of-protection analysis (LOPA) or other process hazard analysis can be useful to understand possible risk and sensitivities of key assumptions. A detailed analysis can then show that automatic isolation, a pressure-relief device sized for leakage, or an alternative means of protection can be preferred. It is necessary for the user to define the appropriate leakage rate for his/her specific system.

4.3.4.4 Pressure consideration for series back-flow prevention

Experience has shown that when inspected and maintained to ensure reliability and capability to limit reverse flow, two back-flow-prevention devices in series are sufficient to eliminate significant reverse flow. As the differential pressure increases, the use of additional safeguards should be considered to reduce the risk of check-valve latent failures resulting in mechanical equipment damage causing loss of containment. The user might want to consider diverse back-flow prevention devices.

If reliability of the series back-flow prevention cannot be assured, then it can be necessary to estimate the reverse flow. The quantity of back-flow leakage through check valves in series depends on the types of check valves, the fouling nature of the fluid and other system considerations. Therefore, it is the responsibility of the user to determine an appropriate technique for estimating the reverse flow through check valves in series. The maximum expected differential pressure across the check valve at the time of demand should be used as the basis for relief rate calculations.

Where no specific experience or company guidelines exist, one may estimate the reverse flow through series check valves as the flow through a single orifice with a diameter equal to one-tenth of the largest check valve's nominal flow diameter. A lower value may be used if a condition-monitoring system for the check valves (e.g. pressure indicators with appropriate volumes between the check valves) is installed to monitor the condition to ensure that the leakage rate is below the capacity of the low-pressure side relief device.

4.3.5 Utility failure

The consequences that can develop from the loss of any utility service, whether plant-wide or local, shall be carefully evaluated. Cases of both the complete loss of a utility and the partial loss of a utility shall be considered. In some cases, a partial utility failure can cause a higher relief load than a total failure because some equipment that contributes to the relief load would remain in operation. Table 1 gives the normal utility services that can fail and a partial listing of affected equipment that can cause overpressure.

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Table 1 — Possible utility failures and equipment affected

Utility failure	Equipment affected
Electric	Pumps for circulating cooling water/medium, boiler feed, quench, or reflux
	Fans for air-cooled exchangers, cooling towers, or combustion air
	Compressors for process vapour, instrument air, vacuum or refrigeration
	Instrumentation
	Motor-operated valves
Cooling water/medium	Condensers for process or utility service
	Coolers for process fluids, lubricating oil or seal oil
	Jackets on rotating or reciprocating equipment
Instrument air	Transmitters and controllers
	Process regulating valves
	Alarm and shutdown systems
Steam	Turbine drivers for pumps, compressors, blowers, combustion air fans, or electric generators
	Reciprocating pumps
	Equipment that uses direct steam injection
	Eductors
Steam/heating medium	Heat exchangers (e.g. reboilers)
Fuel (oil, gas, etc.)	Boilers
	Reheaters (reboilers)
	Engine drivers for pumps or electric generators
	Compressors
	Gas turbines
Inert gas	Seals
	Catalytic reactors
	Purge for instruments and equipment

An evaluation of the effect of overpressure that is attributable to the loss of a particular utility service should include the chain of developments that can occur and the reaction time involved. In situations in which the equipment fails but operates in parallel with equipment that has a different energy source, operating credit may be taken for the unaffected and functioning equipment to the extent that service is maintained.

EXAMPLE 1 An example is a cooling-water circulating system that consists of two pumps in parallel service and continuous operation whose drivers have unrelated energy sources. If one of the two energy sources fails, partial credit can be taken for the other power source that continues to function (see 4.2). The quantity of excess vapour generated because of the energy failure then depends solely on the quantity and available head of the remaining cooling-water pump.

EXAMPLE 2 Another example is two cooling-water pumps in parallel service, with one pump providing the full flow of cooling water and the second being in standby service. The second pump has a separate energy source and is equipped with controls for automatic start-up if the first pump fails. No protective credit is taken for the standby pump because it is not considered totally reliable for the design of individual process equipment relief.

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After detailed study, full or partial protective credit may be taken for parallel, normally operating instrument air compressors and electric generators that have two unrelated sources of energy to the drivers. Manual cut-in of auxiliaries is operator- and time-dependent and shall be carefully analysed before it is used as insurance against overpressure.

4.3.6 Electrical or mechanical failure

The failure of electrical or mechanical equipment that provides cooling or condensation in process streams can cause overpressure in process vessels.

4.3.7 Loss of fans

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 exchangers where independent operation of the louvers can be maintained, credit for the cooling effect may be obtained by convection and radiation.

4.3.8 Loss of heat

4.3.8.1 Loss of heat in series fractionation systems

In series fractionation (that is, where the bottoms from one column feed into another column), the loss of heat input to an upstream column can overpressure the downstream column. Loss of heat results in some of the light ends mixing with the bottoms and being transferred to the next column as feed. Under this circumstance, the overhead load of the downstream column can consist of its normal vapour load plus the light ends from the first column. If the downstream column does not have the condensing capacity for the additional vapour load, excessive pressure could occur.

4.3.8.2 Loss of heat in other equipment

In cases where loss of heat can cause carryover of lighter ends, the potential for downstream equipment overpressure should be considered.

4.3.9 Loss of instrument air or electric instrument power

The loss of instrument air drives all air-operated valves to their specified fail position. This action of many valves can result in overpressure if the specified failure positions of the valves are not selected to prevent overpressure. Likewise, failure of electric instrument power can drive control systems and electrically operated valves to their specified failure positions. Consideration should be given to the effect on flare- or blowdown-system loading of valves failing open or closed due to instrument-air failure or power failure.

4.3.10 Reflux failure

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.

4.3.11 Abnormal heat input from reboilers

Reboilers 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, vapour generation can exceed the process system's ability to condense or otherwise absorb the build-up of pressure, which may include non-condensables caused by overheating.

4.3.12 Heat exchanger tube failure

In shell-and-tube heat exchangers, the tubes are subject to failure from a number of causes, including thermal shock, vibration and corrosion. Whatever the cause, the result is the possibility that the high-pressure stream

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overpressures equipment on the low-pressure side of the exchanger. The ability of the low-pressure system to absorb this release should be determined. The possible pressure rise shall be ascertained to determine whether additional pressure relief is required if flow from the tube rupture discharges into the lower-pressure stream. See 5.19 for additional details.

4.3.13 Transient pressure surges

4.3.13.1 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 pressure-relief valves, since the response time of the valves can be too slow. The oscillating peak pressures, 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 or special bladder-type surge valves should be considered, contingent on proper analysis.

4.3.13.2 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. Pressure-relief valves 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.

4.3.13.3 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).

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

4.3.14 Plant fires

Fire as a cause of overpressure in plant equipment is discussed in 5.15. A provision for initiating a controlled shutdown or installation of a depressuring system for the units can minimize overpressure that results from exposure to external fire.

To limit vapour generation and the possible spread of fire, facilities should also allow for the removal of liquids from the systems. Normally operating product withdrawal systems are considered superior and more effective for removing liquids from a unit, compared with separate liquid pulldown systems. Liquid hold-up required for normal plant operations, including refrigerants or solvents, can be effective in keeping the vessel wall cool and does not necessarily require systems for its removal. Provisions may be made either to insulate the vessel's vapour space and apply external water for cooling or to depressure the vessel using a vapour depressuring system.

Area design should include adequate surface drainage facilities and a means for preventing the spread of flammable liquids from one operating area to another. Easy access to each area and to the process equipment shall be provided for firefighting personnel and their equipment. Fire hydrants, firefighting equipment and fire monitors should be placed in readily accessible locations.

Relief load reduction credit for insulation can be taken provided that the requirements of 5.15.5 are met.

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4.3.15 Process changes/chemical reactions

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 maximum allowable working pressure (MAWP). Where normal pressure-relieving devices cannot protect against these situations, controls are necessary to warn of changes outside the intended temperature/pressure limits to provide corrective action (see 5.9, 5.10 and 5.13).

The potential for a chemical reaction in conjunction with the other overpressure scenarios in 4.3 should be considered when appropriate.

4.4 Recommended minimum relief system design content

The lettered items are categories of the minimum recommended information necessary to provide a complete pressure-relief system. The numbered items are information that might or might not be appropriate for each category depending on the specific installation and company practices.

a) Relief system information:

   1) name,

   2) location,

   3) relief device identification numbers (multiple or combination),

   4) revision number and date,

   5) approvals, if required,

   6) device summary (for multiple devices);

b) Description of protected components:

   1) list of equipment with design conditions,

   2) piping with design conditions,

   3) drawings and equipment files [e.g. P&ID, mechanical drawings, process-flow diagrams (PFDs)];

c) Design codes/standards:

   1) pressure-design code,

   2) system maximum allowable accumulated pressure;

d) Analysis of causes of system overpressure:

   1) required rate and/or area for each cause,

   2) supporting calculations and assumptions,

   3) schematic of system,

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4) heat and material balances,

5) consider all causes of overpressure,

6) credit for safety instrumented systems/HIPS, including safety integrity level and associated reliability calculations;

e) System operating conditions:

1) fluid composition,

2) pressure,

3) temperature,

4) level,

5) phase,

6) hazards (e.g. air present);

f) Systems relieving conditions:

1) fluid composition,

2) relieving pressure,

3) relief temperature,

4) phase,

5) properties,

6) hazards (e.g. air present);

g) Relief device selection/configuration:

1) rupture disk device,

2) spring-loaded conventional pressure-relief valve,

3) spring-loaded balanced bellows pressure-relief valve,

4) pilot-operated pressure-relief valve,

5) buckling-pin device,

6) combinations,

7) other;

h) Pressure-relief valve/rupture disk combination capacity factor (e.g. ASME K_c);

i) Relief system required area:

1) pressure-relief valve,

2) rupture-disk coefficient of discharge (e.g. ASME K_d) method;

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j) Relief-system capacity for rupture disk:

1) superimposed back pressure,

2) system resistance (e.g. ASME K_r),

3) calculation of uncertainty factor involving a derating multiplier in accordance with the pressure-design code (e.g. 0,9 or less for ASME);

k) Pressure-relief valve rated capacity:

1) relief device set pressure;

l) Spring-loaded pressure-relief valve cold differential test pressure;

m) Pressure-relief valve capacity correction for maximum back pressure:

1) built-up back pressure,

2) maximum and minimum superimposed back pressure,

3) back pressure capacity correction factors (e.g. API RP 520-I:2000, K_b or K_w);

n) Rupture-disk specified burst pressure and manufacturing design range selection;

o) Rupture-disk specified disk temperature;

p) Relieving-fluid disposal requirements (closed or atmospheric):

1) flare,

2) flame impingement,

3) thermal radiation,

4) dispersion (toxic or flammable vapour),

5) vapour-cloud explosion,

6) other (personnel exposure, noise, housekeeping, etc.),

7) vapour/liquid separation,

8) environmental considerations,

9) bellows vent to safe location,

10) smokeless burning;

q) Relief device physical installation:

1) pipe-stress analysis,

2) free drainage of inlet and outlet line piping,

3) heat tracing,

4) maintenance considerations,

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5) reaction forces,

6) drains and bleeds,

7) bellows-valve bonnet vent and pilot vent to safe location;

r) Pressure-relief valve inlet-line pressure drop;

s) Relief-device specification sheet(s);

t) Criteria for vacuum protection (see 4.6).

4.5 List of items required in flare-header calculation documentation

The following are the documentation requirements for flare-header analyses:

a) For each flare-header design scenario, a description of the initiating event and of the intermediate consequences that lead to relief flow. For example, for an electric power failure, this description would include the primary element assumed to fail, a list of all power users that would consequently be de-energized, and the consequences of the loss of each user.

b) Documentation of the basis used to define flare-system configuration for the network-flow simulation model. For the base case, this documentation generally consists of a list of piping drawings with revision numbers. For alternate piping configurations, changes from the base case may be marked on the schematic diagram of the system or described in narrative form.

c) Schematic diagram of the flare system showing a pressure profile for each case analysed. The pressure profile shall show calculated back pressure at each relief source.

d) Electronic copies of input files used for the network-flow simulation. At a minimum, files for existing piping network shall be provided.

e) Relief-valve size-selection data sheets showing valve manufacturer (for existing valves), type of valve, set pressure, size and inlet and outlet flange ratings.

f) List of disposal-system loads (e.g. loads from relief devices, depressuring valves and control valves) including source name, temperature, relative molecular mass (“molecular weight”) or composition, and flow rate.

NOTE In this International Standard, the SI term “relative molecular mass” is used rather than “molecular weight”.

g) List of all credits taken to reduce or eliminate disposal-system peak loads, including instrumentation (see 7.1.4 for details).

h) List of instrumentation assumed not to work for each relieving scenario and basis for selection of failure combination.

i) Back pressure limit for each source and basis for limit (e.g. downstream piping design pressure, API Std 526 [9], ISO 4126, manufacturer, critical flow, or derated valve capacity).

j) Acceptance criteria for flare-system capacity, including assumptions and design basis for blowdown drum, knockout drum, flare stack.

4.6 Guidance on vacuum relief

In 4.3 are described the fundamental heat- and-material balance factors that can lead to an increase in operating pressure. Under different circumstances, the same factors can lead to a fall in operating pressure to the extent that vacuum relief is required to prevent the equipment from failing under vacuum.

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Large vessels and equipment designed for low positive pressures are more susceptible to vacuum, as smaller vessels often have an inherent vacuum design capability due to other design requirements. Although there is substantial evidence of damage to storage tanks as a result of vacuum, serious damage has also occurred on process vessels [138].

If equipment contains internal partitions, the potential for a vacuum condition existing in one of the compartments should be considered. This also includes shell-and-tube heat exchangers designed on the basis of a limiting pressure differential across the tubesheet. If the pressure on one side falls to a vacuum condition, the differential pressure limit can be breached.

A vacuum can potentially result when the following occur, singly or in combination.

— The volumetric outflow of material from the protected system exceeds the inflow.

— The energy outflow from the protected system exceeds the energy inflow; or a phase change occurs resulting in material changing from a phase with a higher specific volume to a phase with a lower specific volume.

Examples of possible causes of vacuum include the following:

— removal of liquid from a vessel due to pump-out, siphoning or gravity drainage;

— removal of vapour from a vessel by attachment to pumps/compressors, or by attachment to equipment capable of pulling a vacuum (whether by design, such as vacuum pumps and ejectors, or inadvertently such as vent-collection systems where the flow of material along a header can induce a vacuum in other equipment);

— ambient temperature changes, resulting in contraction of the vapour space; this is normally only a significant issue for storage tanks; see API Std 2000 [14] or prEN 14015-1 [6];

— condensation of vapour, whether by ongoing heat transfer through a condenser (e.g. following loss of reboil to a distillation or regeneration column), by gradual heat transfer (e.g. cooling of equipment following steam-out or plant shutdown) or by injection of cold material into the vapour space (e.g. following a loss of pre-heat);

— physical absorption or adsorption, e.g. the ongoing absorption of a soluble vapour into a liquid (e.g. ammonia into water) following a shutdown or through the inadvertent entry of a suitable absorbent into the process;

— chemical absorption, e.g. the ongoing absorption of sour gas or carbon dioxide into a scrubbing solution;

— other chemical reactions that remove gas or vapour from the vapour space (e.g. the removal of oxygen to form rust in an isolated item of equipment);

— inadvertent blockage of vents designed to allow inflow of gas or vapour to prevent a damaging vacuum from forming. This is principally an issue for storage tanks (see API Std 2000 or prEN 14015). Typical problems arise from covering or blocking vents during maintenance and not reinstating the vent afterwards, using a lute or seal pot to minimize emissions from a vent but thereby preventing gas or vapour from flowing through the vent to mitigate a vacuum and not draining bunded areas so that liquid can cover vent lines that have been brought to grade.

Protection against vacuum typically takes one or more of the following forms:

a) operating procedures;

b) mechanical design, the most robust form of protection;

c) relief system design, often only practicable with low-pressure equipment and relying on suitable designs of pressure-relief valve or atmospheric vent;

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d) instrumented protective system, designed to admit a gas into the protected system to prevent the vacuum from exceeding a specified level.

The simplest solution to a potential vacuum is to design the equipment to withstand full vacuum. This option should always be considered for new equipment. Although smaller equipment can come with a degree of inherent mechanical strength against vacuum, provision of vacuum design for larger equipment can require additional metal thickness or the provision of vacuum support rings. The latter introduce potential corrosion problems unless suitable provision is made for liquid to drain from the rings and the drain holes are kept clear.

Operating procedures are sometimes relied upon for activities associated with preparation for maintenance (e.g. vessel draining and steam-out) and activities such as hydrotesting. However, they are not a foolproof form of protection and designers and operators should agree where procedures can safely be relied upon.

If a vacuum-relief device is used, it shall be rated at a pressure comparable to that of the protected equipment.

An instrumented system or regulator (repressuring system) can be used to add gas to prevent the vacuum from exceeding a specified level. The choice of gas should be considered carefully: air, nitrogen and fuel gas are commonly used, but each introduces different factors of cost, potential product contamination and flammability.

Instrumented protective systems (whether through process control or by a HIPS) are often used to protect pressurized equipment such as distillation columns against vacuum. The philosophy is essentially the same as for the use of relief devices: to admit a suitable gas into the protected equipment to prevent the vacuum from exceeding a defined level. The same considerations about the selection of gas apply. It is unlikely that a simple control scheme has sufficient reliability to be regarded a suitable form of protection; a HIPS (see Annex E) is likely to be required with the attendant requirements of regular proof testing.

If admission of a suitable gas is relied upon for controlling the level of vacuum in the process, calculation of the required rate can prove problematic, especially as the calculation is for a transient condition rather than a steady-state condition. Suitably conservative assumptions about heat transfer rates and compositions should be made and the calculations can prove very complex.

If mechanical design for a pressure other than full vacuum is used, the level of vacuum should be justified by appropriate calculations. It might be possible to show that a vacuum cannot be generated (e.g. because the vapour pressure of the liquid remains above atmospheric).

5 Determination of individual relieving rates

5.1 Principal sources of overpressure

The basis for determining individual relieving rates that result from various causes of overpressure is presented in Clause 5 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.

Table 2 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 Clause 5.

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Table 2 — Guidance for required relieving rates under selected conditions

Item No.	Condition	Liquid-relief guidance [a]	Vapour-relief guidance [a]
1	Closed outlets on vessels	Maximum liquid pump-in rate	Total incoming steam and vapour plus that generated therein at relieving conditions
2	Cooling-water failure to condenser	—	Total vapour to condenser at relieving conditions
3	Top-tower reflux failure	—	Total incoming steam and vapour plus that generated therein at relieving conditions less vapour condensed by sidestream reflux
4	Sidestream reflux failure	—	Difference between vapour entering and leaving equipment at relieving conditions
5	Lean-oil failure to absorber	—	None, normally
6	Accumulation of non-condensables	—	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	—	—
	Water into hot oil	—	For towers, usually not predictable
	Light hydrocarbons into hot oil	—	For heat exchangers, assume an area twice the internal cross-sectional area of one tube to provide for the vapour generated by the entrance of the volatile fluid due to tube rupture
8	Overfilling storage or surge vessel	Maximum liquid pump-in rate	—
9	Failure of automatic controls	—	Analyse on a case-by-case basis
10	Abnormal heat or vapour input	—	Estimated maximum vapour generation including non-condensables from overheating
11	Split exchanger tube	Liquid entering from twice the cross-sectional area of one tube	Steam or vapour entering from twice the cross-sectional area of one tube; also same effects found in Item 7 for exchangers
12	Internal explosions	—	Not controlled by conventional relief devices but by avoidance of circumstances
13	Chemical reaction	—	Estimated vapour generation from both normal and uncontrolled conditions; consider two-phase effects
14	Hydraulic expansion:		
	Cold-fluid shut in	See 5.14	—
	Lines outside process area shut in	See 5.14	—
15	Exterior fire [b]	See 5.15.3.3	Estimated by the methods given in 5.15.2.2 or 5.15.3.2
16	Power failure (steam, electric, or other)	—	Study the installation to determine the effect of power failure; size the relief valve for the worst condition that can occur
	Fractionators	—	Loss of all pumps, with the result that reflux and cooling water would fail
	Reactors	—	Consider failure of agitation or stirring, quench or retarding stream; size the valves for vapour generation from a runaway reaction
	Air-cooled exchangers	—	Fan failure; size valves for the difference between normal and emergency duty
	Surge vessels	—	Maximum liquid inlet rate

[a] Consideration can be given to the reduction of the relief rate as the result of the relieving pressure being above operating pressure.

[b] Guidance on fire relief is given in Annex A.

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5.2 Sources of overpressure

The liquid or vapour 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.

5.3 Effects of pressure, temperature, and composition

Pressure and temperature should be considered to determine individual relieving rates, since they affect the volumetric and compositional behaviour of liquids and vapours. Vapour is generated when heat is added to a liquid. The rate at which vapour 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 vapour that is rich in low-boiling components. As heat input is continued, successively heavier components are generated in the vapour. Finally, if the heat input is sufficient, the heaviest components are vaporized.

During pressure-relieving, the changes in vapour rates and relative molecular masses at various time intervals should be investigated to determine the peak relieving rate and the composition of the vapour. 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 (8) (see 5.15.2.2.2) oversizes the pressure-relief valve. This equation should be used only when physical properties for the fluid are not available.

5.4 Effect of operator response

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.

5.5 Closed outlets

To protect a vessel or system from overpressure when all outlets on the vessel or system are blocked, the capacity of the relief device shall be at least as great as the capacity of the sources of pressure. If all outlets are not blocked, the capacity of the unblocked outlets may properly be considered. The sources of overpressure include pumps, compressors, high-pressure supply headers, stripped gases from rich absorbent and process heat. In the case of heat exchangers, a closed outlet can cause overpressure due to either thermal expansion (see 5.14) or vapour generation.

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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.

5.6 Cooling or reflux failure

5.6.1 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 5.6.2 through 5.6.9 have generally been accepted for determining relieving rates.

5.6.2 Total condensing

The required relieving rate is the total incoming vapour rate to the condenser, recalculated at a temperature that corresponds to the new vapour 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 vapour rate can change significantly.

5.6.3 Partial condensing

The required relieving rate is the difference between the incoming and outgoing vapour rate at relieving conditions. The incoming vapour rate should be calculated on the same basis used in 5.6.2. If the composition or rate of the reflux is changed, the incoming vapour rate to the condenser should be determined for the new conditions.

5.6.4 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 5.6.2 and 5.6.3). 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.

5.6.5 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.

5.6.6 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 vapour properties that affect the required relieving rate. A pressure-relief device 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.

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5.6.7 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.

5.6.8 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 5.6.6 and 5.6.7.

5.6.9 Sidestream reflux failure

Principles similar to those described in 5.6.6 and 5.6.7 apply in overhead-condenser flooding (if a condenser is in the system) or changes in vapour 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.

5.7 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 vapour 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 vapour flow. The case of a synthesis-gas carbon-dioxide removal unit in which the downstream gas goes to a methanator is more complicated to analyse. Any quantity of carbon dioxide above design capability that enters the methanator, as occurs on even partial absorbent failure, produces a rapid temperature rise that usually closes a methanator feed shutoff valve and opens a vent to the atmosphere. If the vent to the atmosphere fails to open, the possibility of overpressure arises.

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.

5.8 Accumulation of non-condensables

Non-condensables do not accumulate under normal conditions, because they are released with the process streams. However, with certain piping configurations, non-condensables can accumulate to the point that the overhead condenser is blocked. This effect is equal to a total loss of cooling.

5.9 Entrance of volatile material into the system

5.9.1 Water into hot oil

Although the entrance of water into hot oil remains a source of potential overpressure, no generally recognized method for calculating the relieving requirements is available. In a limited sense, if the quantity of water present and the heat available in the process stream are known, the size of the pressure-relief device can be calculated like that of a steam valve. Unfortunately, the quantity of water is almost never known, even within broad limits. Also, since the expansion in volume from liquid to vapour is so great (approximately 1:1 400 at atmospheric pressure) and the speed of vapour generation is essentially instantaneous, it is questionable whether the pressure-relief device could open fast enough to be of value. Normally, a pressure-relieving device is not provided for this contingency. Proper design and operation of the process system are essential in attempts to eliminate this possibility. The following are some precautions that can be taken:

a) designing the water side to be at a lower operating pressure than the hot oil side;

b) maintaining minimum circulation of hot oil through equipment on stand-by in order to minimize collection of water;

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c) avoiding water-collecting pockets;

d) installing proper steam condensate traps;

e) installing heat tracing to eliminate condensation;

f) installing double-block and bleed valves on water connections to hot process lines;

g) installing interlocks to trip sources of heat in the event of water-contaminated feedstock.

5.9.2 Light hydrocarbons into hot oil

The information in 5.9.1 applies to the entrance of light hydrocarbons into hot oil even though the ratio of liquid volume to vapour volume can be considerably less than 1:1 400.

5.10 Failure of process stream automatic controls

5.10.1 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 5.10.5). 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 5.10.3. 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 he knows a specific condition that 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.

5.10.2 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 favourable 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 5.10.6), 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.

5.10.3 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

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system are still in operating position at normal flow (that is, 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. If the bypass is used only during maintenance to permit the control valve to be blocked in and removed from service, then the maximum flow of either the control valve or bypass valve needs to be considered.

The potential for the bypass valve to be inadvertently opened while the control valve is operating should also be considered unless administrative controls are in place. If the pressure resulting from the opening of the bypass valve can exceed the corrected hydrotest pressure (see 3.21 and 4.3.2), reliance on administrative controls as the sole means to prevent overpressure might not be appropriate. 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 vapours 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 vapours is large compared with the volume of the low-pressure system or if the source of vapour 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 vapour flow through the liquid control valve.

In circumstances where process systems involve significant differences in pressure level and the volume of vapour 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 vapour 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. 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 vapour 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 vapour volume, it can prove worthwhile to take credit for the following: The transfer of vapours 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 vapour to the high-pressure system that tends to maintain upstream pressure.

5.10.4 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, pressure-relieving devices 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.

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For applications involving single outlets with control devices that fail in the closed position, pressure-relieving devices 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 5.5.

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.

5.10.5 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.

5.10.6 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 pressure-relieving devices 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 vapour flashed from the normal liquid entry.

5.10.7 Piping design considerations for gas breakthrough

Gas breakthrough across a control valve can result in slug-flow high liquid velocities. The resultant transient loads on the piping shall be taken into account, including the mechanical design and pipe supports.

NOTE Locating the relief device closer to the upstream control valve can reduce the amount of pipe support required and can also reduce the size of the relief device.

5.11 Abnormal process heat input

The required relieving rate is the maximum rate of vapour generation at relieving conditions (including any non-condensables produced from overheating) less the rate of normal condensation or vapour outflow. In every case, the designer should consider the potential behaviour 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.

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5.12 Internal explosion (excluding detonation)

If overpressure protection against internal explosions caused by ignition of vapour-air mixtures is to be provided, rupture disks or explosion-vent panels, not relief valves, should be used. Relief valves react too slowly to protect the vessel against the extremely rapid pressure build-up 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 vapours 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 [34]. 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 [35], 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 lieu 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.

5.13 Chemical reaction

5.13.1 The methodology for determining the appropriate size of an emergency vent system for chemical reactions has been established by DIERS (Design Institute for Emergency Relief Systems) [38], [39], [40], [41], [42].

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 formula that account for two-phase gas/liquid vent flow.

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

— external fire;

— loss of mixing;

— loss of cooling;

— mischarge of reagents.

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5.13.3 Reaction rates are rarely known; therefore, bench scale tests simulating the design basis upset condition are usually required. There are a number of test apparati 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 non-condensable products and the rate of temperature rise is not tempered by boiling liquid. Gassy systems can be either liquid-phase decompositions or vapour-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 non-condensable gas.

Following characterization of the system, the appropriate vent-sizing formula can be selected. An excellent discussion of these procedures is contained in Grolmes *et al* [38]. 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 5.12 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 pressure-relief device should be used to control overpressure. Where this is infeasible, other design strategies can be employed to control equipment over-stressing. 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 International Standard.

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

5.14 Hydraulic expansion

5.14.1 Causes

Hydraulic expansion is the increase in liquid volume caused by an increase in temperature (see Table 3). 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) An exchanger is blocked in on the cold side with flow in the hot side.

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

In certain installations, such as cooling circuits, the processing scheme, equipment arrangements and methods, and operation procedures make feasible the elimination of the hydraulic-expansion relieving device, which is normally required on the cooler, fluid side of a shell-and-tube exchanger. Typical of such conditions are multiple-shell units 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 exchanger unit should be controlled by administrative procedures and possibly the addition of signs stipulating the proper venting and draining

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procedures when shutting down and blocking in. Such cases are acceptable and do not compromise the safety of personnel or equipment, but the designer is cautioned to review each case carefully before deciding that a relieving device based on hydraulic expansion is not warranted.

Table 3 — 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	0,000 72 (0,000 4)
35 to 50,9	0,000 9 (0,000 5)
51 to 63,9	0,001 08 (0,000 6)
64 to 78,9	0,001 26 (0,000 7)
79 to 88,9	0,001 44 (0,000 8)
89 to 93,9	0,001 53 (0,000 85)
94 and lighter	0,001 62 (0,000 9)
water	0,000 18 (0,000 1)

[a] At 15,6 °C (60 °F).

5.14.2 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 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 5.14.3 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 5.21.2 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 pressure-relieving device should be set high enough to open only under hydraulic expansion conditions. If thermal-relief valves discharge into a closed system, the effects of back pressure should be considered.

5.14.3 Special cases

Two general applications for which thermal relieving devices larger than a DN 20 × DN 25 (NPS ¾ × NPS 1) valve can be required are long pipelines of large diameter in uninsulated, aboveground installations and large vessels or 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 [43] 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.

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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 (1), in SI units, or Equation (2) in USC units:

$$q = \frac{\alpha_{\mathrm{v}} \cdot \phi}{1\,000 d \cdot c} \tag{1}$$

where

q is the volume flow rate at the flowing temperature, expressed in cubic metres per second;

α_{v} is the cubic expansion coefficient for the liquid at the expected temperature, expressed in 1/°C;

NOTE This information is best obtained from the process-design data; however, Table 3 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 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.

$$q = \frac{\alpha_{v} \cdot \phi}{500 d \cdot c} \tag{2}$$

where

q is the volume flow rate at the flowing temperature, expressed in U.S. gallons per minute;

α_{v} is the cubic expansion coefficient for the liquid at the expected temperature, expressed in 1/°F;

NOTE This information is best obtained from the process design data; however, Table 3 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 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 Btu/lb·°F.

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

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5.14.4 Piping

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

a) the piping always contains a pocket of non-condensing vapour, such that it can never become liquid-full;

CAUTION — Small vapour or gas pockets can disappear upon heating due to compression and/or solubilization. In contrast, multi-component mixtures with a wide boiling range can always have sufficient vapour 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 vapour pocket is sufficient for liquid expansion.

or

b) the piping is in continuous use (i.e., not batch or semi-continuous 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 (3) (Karcher [44] and CCPS [45]):

$$p_2 = p_1 + \frac{(T_2 - T_1)(\alpha_v - 3\alpha_l) - \left(\frac{q_{ll} \cdot t}{V}\right)}{\chi + \left(\frac{d}{2E \cdot \delta_w}\right)(2{,}5 - 2\mu)} \qquad (3)$$

where

p_2 is the final gauge pressure of blocked-in, liquid-full equipment, expressed in kPa (psi);

p_1 is the initial gauge pressure of blocked-in, liquid-full equipment, expressed in kPa (psi);

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 metres (inches);

E is the modulus of elasticity for the metal wall at T_2, expressed in kPa (psi);

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δ_w is the metal wall thickness, expressed in metres (inches);

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 m^3/s (in^3/s);

t is the elapsed time for leakage, expressed in seconds;

V is the pipe volume, expressed in cubic metres (cubic inches).

Selected data for α_l and E are given in Table 4. See Perry's Handbook [46] for data on other materials.

Table 4 — 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)	$1{,}21 \times 10^{-5}$ ($6{,}7 \times 10^{-6}$)	30×10^6 (207×10^6)
304 stainless steel	$9{,}6 \times 10^{-6}$ ($1{,}73 \times 10^{-5}$)	28×10^6 (193×10^6)
316 stainless steel	$1{,}60 \times 10^{-5}$ ($8{,}9 \times 10^{-6}$)	28×10^6 (193×10^6)
Alloy 600	$1{,}1 \times 10^{-5}$ to $1{,}66 \times 10^{-5}$ ($6{,}1 \times 10^{-6}$ to $9{,}2 \times 10^{-6}$)	25×10^6 to 32×10^6 (172×10^6 to 221×10^6)
Nickel-copper alloy	$1{,}01 \times 10^{-5}$ to $1{,}42 \times 10^{-5}$ ($5{,}6 \times 10^{-6}$ to $7{,}9 \times 10^{-6}$)	$24{,}5 \times 10^6$ to $30{,}9 \times 10^6$ (169×10^6 to 213×10^6)

Where data are unavailable, the Equations (4) and (5) can be used to estimate, respectively, the isothermal compressibility coefficient, x (see Lange's Handbook of Chemistry, 12th Edition [47], pages 10 to 122) and the cubic expansion coefficient, α_v (see of Perry's Handbook [46], 5th Edition, pages 3 to 227):

$$\alpha_v = \frac{\rho_1^2 - \rho_2^2}{2(T_2 - T_1)\ \rho_1 \cdot \rho_2} \tag{4}$$

where

α_v is the cubic expansion coefficient, expressed in 1/°C (1°F);

ρ_1 is the density of liquid at the first temperature, expressed in kg/m^3 (lb/ft^3);

ρ_2 is the density of liquid at the second temperature, expressed in kg/m^3 (lb/ft^3);

T_1 is the first temperature, expressed in °C (°F);

T_2 is the second temperature, expressed in °C (°F).

$$x = \frac{1}{v_1} \frac{(v_1 - v_2)}{(p_2 - p_1)} \tag{5}$$

where

x is the isothermal compressibility coefficient, expressed in 1/kPa (1/psi);

v_1 is the specific volume of liquid at the first pressure, expressed in m^3/kg (ft^3/lb);

v_2 is the specific volume of liquid at the second pressure, expressed in m^3/kg (ft^3/lb);

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p_1 is the first absolute pressure, expressed in kPa (psi);

p_2 is the second absolute pressure, expressed in kPa (psi).

5.14.4.2 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 3-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) vapour pressure of the fluid at the heated temperature: Fluids above their atmospheric boiling point continue to release material as vapour through a leak until the fluid temperature cools to the boiling point.

e) adequacy of procedures and administrative controls to avoid blocking in.

5.15 External pool fires

5.15.1 General

5.15.1.1 Effect of fire on the wetted surface of a vessel

To determine vapour 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 40 m (approx. 130 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 5 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.

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 5.15.4). Also, if exposure to fire results in vapour generation from thermal cracking, alternate sizing methods can be appropriate.

The wetted area for spheres normally includes all area up to the maximum diameter. Table 5 recommends the wetted surface area for spheres be based on “the maximum horizontal diameter or up to a 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.

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Table 5 — 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 hold-up 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 storage tanks and process tanks, see API Std 2000 or prEN 14015.
Spheres and spheroids	Up to the maximum horizontal diameter or up to the height of 7,6 m (25 ft), whichever is greater.	—

5.15.1.2 Effect of fire on the unwetted surface of a vessel

5.15.1.2.1 Unwetted wall vessels are those in which the internal walls are exposed to a gas, vapour or super-critical fluid, or are internally insulated regardless of the contained fluids. These include vessels that contain separate liquid and vapour 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 shall be considered (see 5.15.4 and 5.15.5).

5.15.1.2.2 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 resistance of the contained fluid or any internal insulating material. Heat input from an open 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. Figures 1 and 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 (1 100 °F) and 17 min to reach 704 °C (1 300 °F) when the plate is exposed to an open fire. Recent calculations indicate that the heat flux of the fire is in the range of approximately 80 kW/m^2 to 100 kW/m^2 (25 200 Btu/ft^{2}h to 31 500 Btu/ft^{2}h).

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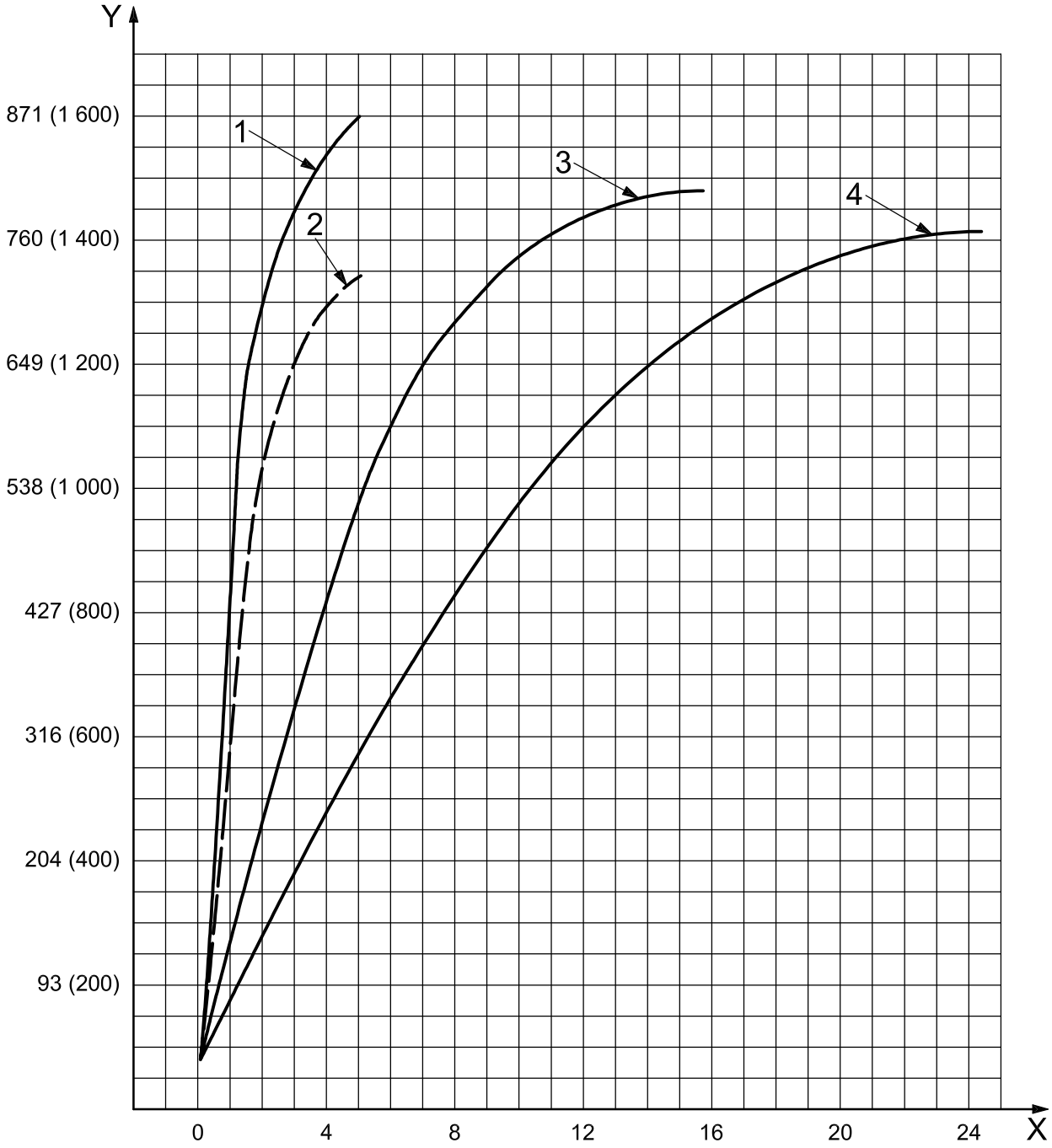

Key

X time after start of the fire, expressed in minutes

[AC1] Y plate temperature, averaged over 2,3 m^2 (24 ft^2), expressed in degrees Celsius (degrees Fahrenheit) [AC1]

1 plate 3,2 mm (1/8 in) thick, as computed

2 plate 3,2 mm (1/8 in) thick, observed

3 plate 12,7 mm (1/2 in) thick, as computed

4 plate 25,4 mm (1 in) thick, as computed

Figure 1 — Average rate of heating steel plates exposed to open gasoline fire on one side

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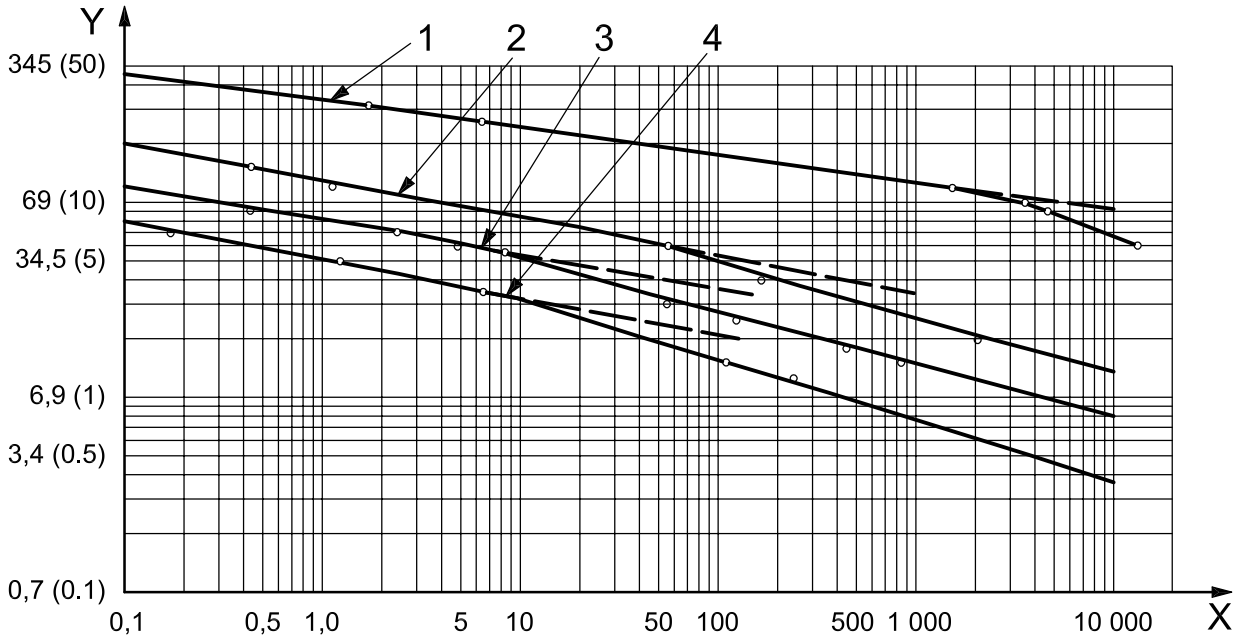
Figure 2 shows the effect of overheating ASTM A515 Grade 70 carbon steel, from data published in Reference [48]. The figure indicates that at a stress of 138 MPa (20 000 psi), an unwetted steel vessel ruptures in about 0,1 h at 649 °C (1 200 °F). A source for time-dependent rupture stress for different metals is ASTM Data Series DS 11S1 [29], 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. A more recent source is Guidelines for the Protection of Pressurised Systems Exposed to Fire [30].



Key

X time for rupture, expressed in hours

Y stress, expressed in megapascals (kilopounds per square inch)

1 537 °C (1 000 °F)

2 649 °C (1 200 °F)

3 744 °C (1 300 °F)

4 760 °C (1 400 °F)

Figure 2 — Effect of overheating carbon steel (ASTM A515, Grade 70)

5.15.2 Fire relief loads

5.15.2.1 General

The appropriate fire sizing equation from the API or ISO International Standard that applies to the equipment being evaluated should be used. For example, the scope of API Std 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). API Std 2000 is used for process and other storage tanks designed in accordance with API Std 620 [12] or API Std 650 [13]. The fire-sizing equations in Clause 5 apply to process vessels and storage vessels, including those designed to the pressure-design code. These equations were re-evaluated by the API Pressure Relief Subcommittee and found to be appropriate for the specific equipment covered by this International Standard. The fire-sizing equations in Clause 5 assume typical in-plant conditions for facilities within the scope of this International Standard but can be understated for vessels in partially enclosed or enclosed areas, such as those in buildings or on-offshore platforms. For further information, see References [67] or [30]; these documents provide an alternative approach based on analytical methods and can be used to model fire-heat input for all

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types and sizes of fire. To use these methods for fire-relief calculations, it is necessary to specify the average fire temperature, rather than the instantaneous peak temperature. For a wetted area of 10 m^2 (approx. 100 ft^2) and an average fire temperature of 750 °C (approx. 1 400 °F), these calculation methods give the same results as Equation (7).

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 alternative relief paths that remain open during an overpressure event is generally an acceptable practice. However, it should be recognized that operators and/or emergency responders 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, unless the line is open to atmosphere, consideration should be given to the potential that the fire-relief flow in the alternative relief 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.

In 5.15.2.2 is described the heat absorption equations for vessels containing liquids and heat absorption equations for vessels containing only gases/vapours.

Either the vapour 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 vapour thermal expansion versus boiling-liquid vaporization have been determined.

5.15.2.2 Heat absorption equations for vessels

5.15.2.2.1 Heat absorption to 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) and fireproofing measures. Equation (6) is used to evaluate these conditions if there are prompt firefighting efforts and drainage of flammable materials away from the vessels.

$$Q = C_1 \cdot F \cdot A_{ws}^{0,82} \tag{6}$$

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 6);

A_{ws} is the total wetted surface, expressed in square metres (square feet).

NOTE 1 See 5.15.1.1 and Table 5.

NOTE 2 The expression, $A_{ws}^{0,82}$, is the area exposure factor or ratio. This ratio recognizes the fact that large vessels are less likely than small ones to be completely exposed to the flame of an open fire.

Where adequate drainage and firefighting equipment do not exist, Equation (7) should be used [49]:

$$Q = C_2 \cdot F \cdot A_{ws}^{0,82} \tag{7}$$

where C_2 is a constant [= 70 900 in SI units (34 500 in USC units)].

Some measures are necessary to control the spread of major spills from one area to another and to control surface drainage and refinery waste water. This can be accomplished by the strategic use of sewers and trenches with adequate capacity and/or by using the natural slope of the land.

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The selection of the appropriate fire heat-flux equation requires determination if there is “adequate drainage”. The determination of what constitutes adequate drainage is subjective and left to the user to decide but it should be designed to carry flammable/combustible liquids away from a vessel. The method of removal (e.g. sewers, open trenches, sloping, etc.) should consider not only the flow of the flammable or combustible liquids causing the pool fire but also the firewater that is applied by emergency responders. Some example drainage criteria are given in API 2510 [19].

Table 6 — Environment factor

Type of equipment		Environment factor [a] F
Bare vessel		1,0 [c]
Insulated vessel [b], with insulation conductance values 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)	0,037 6
	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

NOTE Local instantaneous pool fire heat fluxes as high as 190 kW/m^2 (60 000 Btu/ft^2-h) have been reported. When designing pressure-relief systems, consideration is generally given to the use of time-weighted average fire heat fluxes rather than instantaneous peaks as some time is required for the contents to reach relieving conditions.

[a] These are suggested values for the conditions assumed in 5.15.2. 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 5.15.4 and 5.15.5.

[b] Insulation should resist being dislodged by firehose streams (5.15.5.2). For the examples, a temperature difference of 871 °C (1 600 °F) was used. These conductance values are computed from Equation (13) and are based upon insulation having thermal conductivity of 0,58 W/m·K (4 Btu·in/h·ft^2·°F) at 538 °C (1 000 °F) and correspond to various thicknesses of insulation between 25,4 mm (1 in) and 304,8 mm (12 inches). See Equation (13) to determine the environment factor, F.

[c] See 5.15.4.2.

[d] See 5.15.4.3.

[e] The environment factor, F, in Equations (6) and (7) does not apply to uninsulated vessels. The environment factor should be replaced by 1,0 when calculating heat input to uninsulated vessels.

5.15.2.2.2 Vessels containing only gases, vapours or super-critical fluids

See 5.15.1.2 for a discussion of the effect of fire on the unwetted surface of a vessel.

The discharge areas for pressure-relief devices on vessels containing super-critical fluids, gases or vapours exposed to open fires can be estimated using Equation (8). In certain cases, the normal operating pressure can be below the thermodynamic critical conditions but the relieving pressure is supercritical. In such cases, the guidance in 5.15.2.2.2 can be used to size the relief device. In the use of Equation (8), no credit has been taken for insulation. Credit for insulation may be taken per Table 6.

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$$A = \frac{F' \cdot A'}{\sqrt{p_1}} \tag{8}$$

where

[AC1⟩ A is the effective discharge area of the valve, expressed in square inches;

A' is the exposed surface area of the vessel, expressed in square feet;

p_1 is the upstream relieving absolute pressure, expressed in psi; ⟨AC1]

NOTE p_1 is the set pressure plus the allowable overpressure plus the atmospheric pressure.

F' can be determined using Equation (9). If calculated using Equation (9) and the result is less than 0,01, then use a recommended minimum value of $F' = 0{,}01$. If insufficient information is available to use Equation (9), then use $F' = 0{,}045$.

$$F' = \frac{0{,}1406}{C \cdot K_D}\left[\frac{(T_w - T_1)^{1{,}25}}{T_1^{0{,}6506}}\right] \tag{9}$$

where

K_D is the coefficient of discharge (obtainable from the valve manufacturer);

NOTE A K_D value of 0,975 is typically used for preliminary sizing of pressure-relief valves (see API RP 520-1 or ISO 4126).

[AC1⟩ T_w is the recommended maximum wall temperature of vessel material, expressed in °R;

T_1 is the gas absolute temperature, at the upstream relieving pressure, determined from Equation (11), expressed in °R. ⟨AC1]

The constant, C, is given by Equation (10):

$$C = 520\sqrt{k\left(\frac{2}{k+1}\right)^{\frac{k+1}{k-1}}} \tag{10}$$

where

$520 = 3\,600\sqrt{\frac{g}{R}}$;

k is the specific heat ratio (C_p/C_v) of gas or vapour at relieving conditions;

[AC1⟩ g is the gravitational constant, expressed in ft-lb/lbf-s^2. ⟨AC1]

[AC1⟩ *Text deleted* ⟨AC1]

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