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

Industries du pétrole, de la pétrochimie et du gaz naturel — Systèmes de dépressurisation et de protection contre les surpressions

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Foreword

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The main task of technical committees is to prepare International Standards. Draft International Standards adopted by the technical committees are circulated to the member bodies for voting. Publication as an International Standard requires approval by at least 75 % of the member bodies casting a vote.

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ISO 23251 was prepared by Technical Committee ISO/TC 67, *Materials, equipment and offshore structures for petroleum, petrochemical and natural gas industries*, Subcommittee SC 6, *Processing equipment and systems*.

This corrected version of ISO 23251:2006 incorporates corrections to Table 4, column 2, second row under the header, and the five rows of data in column 3.

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.

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.

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.

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)

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

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

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

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.

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

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.

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 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:

In SI units:

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

In USC units:

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

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

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.

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.

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 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.

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,

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;

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,

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.

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;

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.

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.

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.

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.

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;

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 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.

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.

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.

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 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.

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.

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]):

$$p_2 = p_1 + \frac{(T_2 - T_1)(\alpha_\mathrm{v} - 3\alpha_\mathrm{l}) - \left(\dfrac{q_\mathrm{ll} \cdot t}{V}\right)}{\chi + \left(\dfrac{d}{2E \cdot \delta_\mathrm{w}}\right)(2{,}5 - 2\mu)} \qquad (3)$$

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);

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

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);

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.

Table 5 — Effects of fire on the wetted surfaces of a vessel

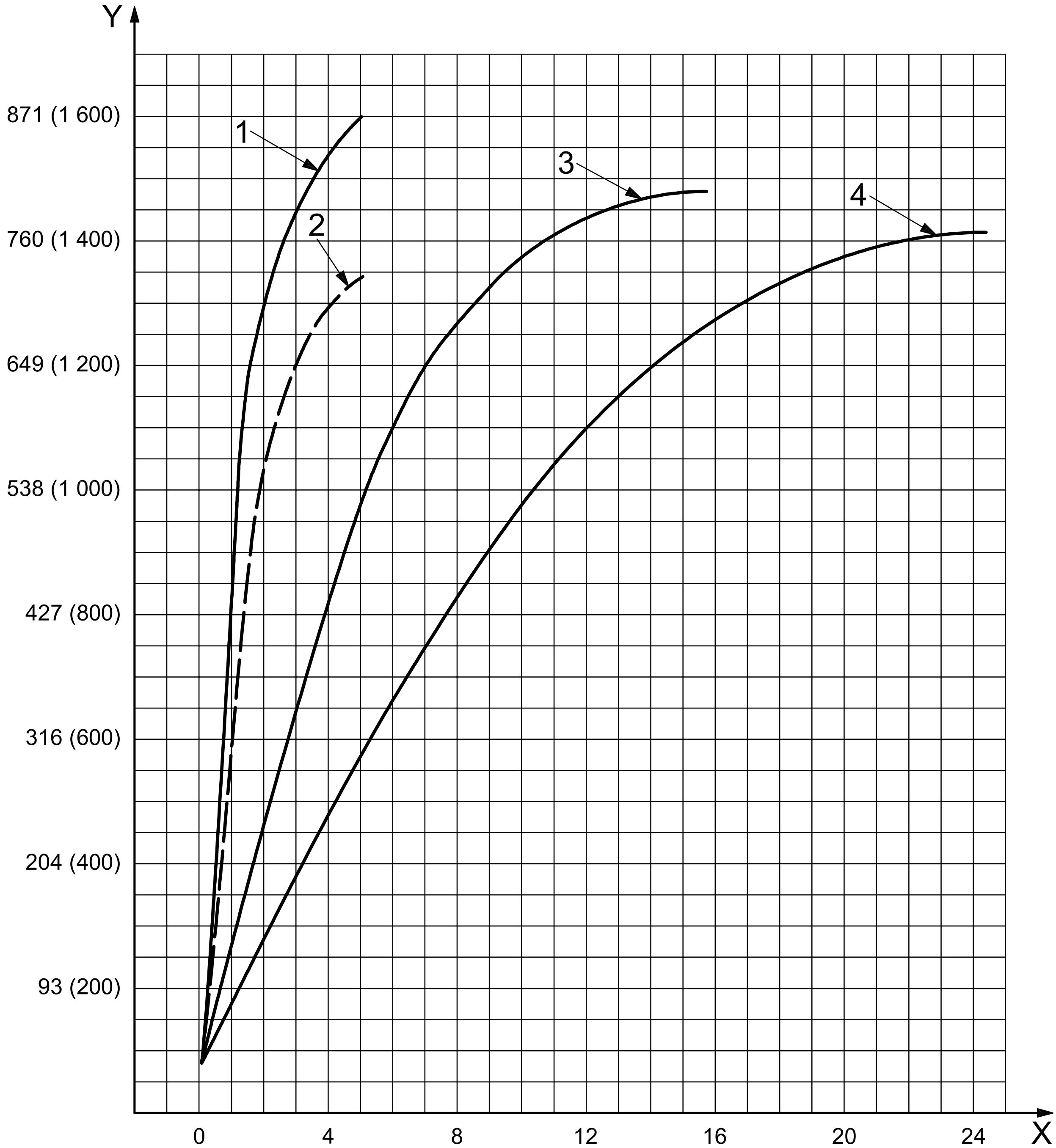
Class of vessel	Portion of liquid inventory	Remarks
Liquid-full, such as treaters	All up to the height of 7,6 m (25 ft).	—
Surge drums, knockout drums, process vessels	Normal operating level up to the height of 7,6 m (25 ft).	—
Fractionating columns	Normal level in bottom plus liquid hold-up from all 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).



Key

X time after start of the fire, expressed in minutes

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

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

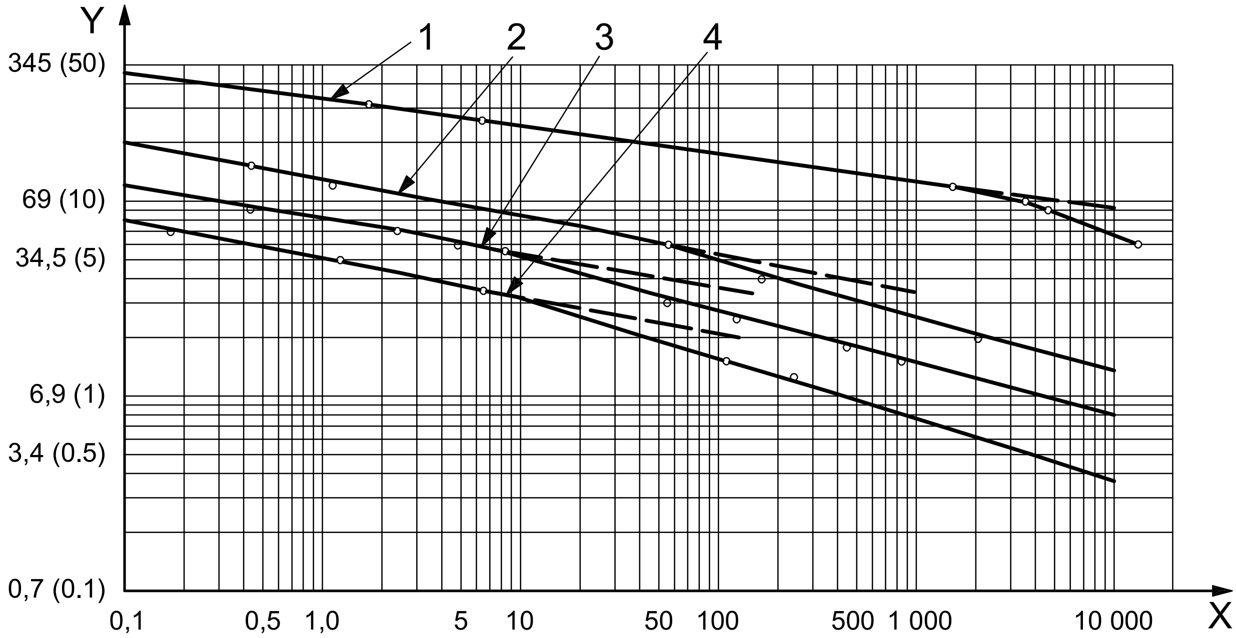
2 plate 3,2 mm (1/8 in) thick, observed

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

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 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.

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

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, $A_{\mathrm{ws}}{}^{0,82}$, 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]:

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

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.

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.

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

where

A is the effective discharge area of the valve, expressed in SI units (square inches);

A' is the exposed surface area of the vessel, expressed in SI units (square feet);

p_1 is the upstream relieving absolute pressure, expressed in SI units (psi);

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).

T_w is the recommended maximum wall temperature of vessel material, expressed in SI units [°R[2]];

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

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;

g is the gravitational constant, expressed in SI units (ft-lb/lbf-s^2).

2) °R is a deprecated unit.

$$T_1 = \frac{p_1}{p_n} \cdot T_n \tag{11}$$

where

p_n is the normal operating gas absolute pressure, expressed in SI units (psi);

T_n is the normal operating gas absolute temperature, expressed in SI units (°R).

The recommended maximum vessel wall temperature, T_w, for the usual carbon steel plate materials is 593 °C (1 100 °F). If vessels are fabricated from alloy materials, the value for T_w should be changed to a more appropriate recommended maximum. See 5.15.4 for guidance on the potential for vessel failure from over-temperature due to fire exposure.

The relief load, $q_{m,\text{ relief}}$, expressed in pounds per hour, can be calculated directly by rearranging the critical vapour equation and substituting Equations (8) and (9), which results in Equation (12):

$$q_{m,\text{relief}} = 0{,}1406\sqrt{M \cdot p_1}\left[\frac{A'(T_w - T_1)^{1,25}}{T_1^{\,1,1506}}\right] \tag{12}$$

where M is the relative molecular mass of the gas.

NOTE Z and K_b in API RP 520-I:2000, Equation 3.2, have each been assumed to have a value of 1.

The derivations of Equations (8), (9) and (12) [50] are based on the physical properties of air and the perfect gas laws. The derivations assume that the vessel is uninsulated and has no mass, that the vessel wall temperature do not reach rupture-stress temperature, and that there is no change in fluid temperature. These assumptions should be reviewed to ensure that they are appropriate for any particular situation. Insulation that meets the fireproofing criteria outlined in 5.15.5 offers a mitigating benefit when gas-filled vessels are exposed to a fire by decreasing the rate at which the metal wall temperature rises. However, no credit for fireproofing is recommended when determining fire relief requirements of gas-filled vessels because, in most cases, a relatively small relief device is required even without a fireproofing credit. Finally, the relationship is empirical and hence there is no engineering basis for providing an environmental factor for this equation.

The surface area potentially exposed to a fire should be used when determining the fire-relief requirements of gas-filled vessels. Note that either the vapour thermal-expansion relief load or the boiling-liquid vaporization relief load, but not both, should be used when sizing the relief device for fire exposure. There are no known experimental studies where separate contributions of vapour thermal expansion versus boiling liquid vaporization have been determined.

When sizing the pressure-relief device for fire exposure, the contribution of vaporizing liquid compared with vapour expansion is governing unless only the wetted surface is fireproofed in accordance with 5.15.5 (and the unwetted surfaces are not) or for high-boiling-point liquids.

5.15.2.3 More rigorous calculations

If the user considers that the preceding assumptions in 5.15.2.2 are not appropriate, more rigorous methods of calculations may be specified. In such cases, it can be necessary to obtain the required physical properties of the containing fluid from actual data or estimated from equations of state. It might be necessary to consider the effects of vessel mass and insulation. The pressure-relieving rate is based on an unsteady state. As the fire continues, the vessel-wall temperature and the contained-gas temperature and pressure increase with time. The pressure-relief valve opens at the set pressure. With the loss of fluid on relief, the temperatures further increases at the relief pressure. If the fire is of sufficient duration, the temperature increases until vessel rupture occurs. Procedures are available for estimating the changes in average vessel-wall and contained-fluid temperatures that occur with time and the maximum relieving rate at the set pressure [51], [52]. These procedures require successive iteration. For fire-insulated segments exposed to fire, it is recommended to assume the fire temperature outside the insulation layer and that the heat input to the fluid is calculated by conduction through the insulation layer and the vessel wall. The heat-transfer resistance from the wall to the fluid is very low compared to the insulation layer's resistance and can be (is usually) neglected. A more rigorous method is described in Reference [141].

There are temperature differences between the liquid and gas phases. Tools are becoming available to perform non-equilibrium temperature calculations; for further information, see Reference [30].

5.15.3 Fluids to be relieved

5.15.3.1 General

A vessel can contain liquids or vapours or fluids of both phases. The liquid phase can be subcritical at operating temperature and pressure and can pass into the critical or supercritical range during the duration of a fire as the temperature and pressure in the vessel increase.

The quantity and composition of the fluid to be relieved during a fire depend on the total heat-input rate to the vessel under this contingency and on the duration of the fire.

The total heat input rate to the vessel may be computed by means of one of the formulas in 5.15.2 using the appropriate values for wetted or exposed surfaces and for the environment factor.

Once the total heat-input rate to the vessel is known, the quantity and composition of the fluid to be relieved can be calculated, providing that enough information is available on the composition of the fluid contained in the vessel.

If the fluid contained in the vessel is not completely specified, assumptions should be made to obtain a realistic relief flow rate for the relief device. These assumptions may include the following:

a) estimation of the latent heat of the boiling liquid and the appropriate relative molecular mass of the fraction vaporized;

b) estimation of the thermal-expansion coefficient if the relieving fluid is a liquid, a gas or a supercritical fluid where a phase change does not occur.

5.15.3.2 Vapour

For pressure and temperature conditions below the critical point, the rate of vapour formation (a measure of the rate of vapour relief required) is equal to the total rate of heat absorption divided by the latent heat of vaporization. The vapour to be relieved is the vapour that is in equilibrium with the liquid under conditions that exist when the pressure-relief device is relieving at its accumulated pressure.

The latent heat and relative molecular mass values used in calculating the rate of vaporization should pertain to the conditions that are capable of generating the maximum vapour rate.

The vapour and liquid composition can change as vapours are released from the system. As a result, temperature and latent-heat values can change, affecting the required size of the pressure-relief device. On occasion, a multicomponent liquid can be heated at a pressure and temperature that exceed the critical temperature or pressure for one or more of the individual components. For example, vapours that are physically or chemically bound in solution can be liberated from the liquid upon heating. This is not a standard latent-heating effect but is more properly termed degassing or dissolution. Vapour generation is determined by the rate of change in equilibrium caused by increasing temperature.

For these and other multicomponent mixtures that have a wide boiling range, it might be necessary to develop a time-dependent model where the total heat input to the vessel not only causes vaporization but also raises the temperature of the remaining liquid, keeping it at its boiling point.

Reference [52] gives an example of a time-dependent model used to calculate relief requirements for a vessel that is exposed to fire and that contains fluids near or above the critical range.

The recommended practice of finding a relief vapour flow rate from the heat input to the vessel and from the latent heat of liquid contained in the vessel becomes invalid near the critical point of the fluid, where the latent heat approaches zero and the sensible heat dominates.

If no accurate latent heat value is available for these hydrocarbons near the critical point, a minimum value of 115 kJ/kg (50 Btu/lb) is sometimes acceptable as an approximation.

For fire contingencies with regard to vessels containing heavy ends (e.g. vacuum-column bottoms), the vaporization temperature can be significantly above the temperature at which the vessel fails. Hence, sizing should not be based on liquid vaporization. In this case, the pressure-relief device may be sized for the products of thermal cracking at a temperature at which the decomposition occurs.

If pressure-relieving conditions are above the critical point, the rate of vapour discharge depends only on the rate at which the fluid expands as a result of the heat input because a phase change does not occur.

5.15.3.3 Liquid

The hydraulic-expansion equations given in 5.14.3 may be used to calculate the initial liquid relieving rate in a liquid-filled system when the liquid is still below its boiling point. However, this rate is valid for a very limited time, after which vapour generation becomes the determining contributor in the sizing of the pressure-relief device.

There is an interim time period between the liquid-expansion and the boiling-vapour relief during which it is necessary to relieve the mixtures of both phases simultaneously, either as flashing, bubble, slug, froth or mist flow until sufficient vapour space is available inside the vessel for phase separation. With the exception of foamy fluids, reactive systems and narrow-flow passages (such as vessel jackets), this mixed-phase condition is usually neglected during sizing and selecting of the pressure-relief device. The aforementioned exceptions are discussed further in 5.15.3.4. Experience as well as recent work in this area [53], [54], [55], [56] has shown that the time required to heat a typical system from the relief-device set pressure to the relieving conditions allows for the relief of any two-phase flow prior to reaching the relieving conditions. As such, full disengagement of the vapour is realized at the relieving conditions and the assumption of vapour-only venting is appropriate for relief device sizing.

Experience has shown there is minimal impact on the discharge system for the two-phase transition period. However, the user may consider the impact of transient two-phase flow on the design of the downstream systems.

If a pressure-relief device is located below the liquid level of a vessel exposed to fire conditions, the pressure-relief device should be able to pass a volume of fluid equivalent to the volume of vapour generated by the fire.

Determination of the appropriate state of the fluid can be complicated. A typical conservative assumption is to use bubble point liquid.

5.15.3.4 Mixed phase

Two-phase relief-device sizing is not normally required for the fire case, except for unusually foamy materials or reactive chemicals [53], [54], [55], [56], [57].

In non-reactive systems subjected to an external fire, boiling occurs at or near the walls of the vessel, commonly referred to as wall-heating. On the other hand, reactive systems in which an external fire can result in an exothermic reaction are subject to boiling throughout the volume of the vessel due to heat evolved from the reaction. This is commonly referred to as volumetric heating, which results in more liquid-swell than wall-heating and, thus, increases the potential for longer-duration two-phase relief. Furthermore, significantly higher heat-generation rates associated with runaway reactions result in higher vapour velocities and further potential for long-duration two-phase flow. The Design Institute of Emergency Relief Systems concluded an intensive research programme to develop methods for the design of emergency relief systems to handle runaway reactions. The interested reader can obtain more information on this subject from References [39] and [51].

5.15.4 Protective measures excluding insulation

5.15.4.1 General

The determination can be made that a pressure-relief device does not provide sufficient protection from vessel rupture for an unwetted-wall vessel or a vessel containing high-boiling-point liquid. Where a pressure-relief valve alone is not adequate, additional protective measures should be considered, such as water sprays (see 5.15.4.2), depressuring (see 5.15.4.3), fireproofing (see 5.15.5), earth-covered storage and diversion walls.

Where local jurisdiction permits, it can be appropriate to utilize these protective measures as an alternative to relief devices sized for the fire case under the following circumstances.

a) Vessel contains vapour only or a high-boiling-point liquid.

b) An engineering analysis indicates that additional protection provided by the relief device serves little value in reducing the likelihood of vessel rupture.

If calculations indicate that vessel rupture can occur, a rupture disk device will burst at a lower pressure if it is heated (e.g. by the fire).

The design should allow sufficient time for operator reaction and initiation of firefighting procedures to avoid vessel rupture. Operator action may include depressuring, using water sprays, employing firewater monitors and isolating the source of fuel.

5.15.4.2 Cooling the surface of a vessel with water

Under ideal conditions, water films covering the metal surface can absorb most incident radiation from pool-fire-flame impingement. The reliability of water application depends on many factors. Freezing weather, high winds, clogged systems, unreliable water supply and vessel surface conditions can prevent uniform water coverage. Because of these uncertainties, no reduction in environment factor (see Table 6) is recommended; however, as stated previously, properly applied water can be very effective. NFPA 15 [33] and API Publ 2030 [16] provide design guidance for fixed water spray systems.

The effect of water is twofold: cooling of surface and reduction of fire heat flux.

5.15.4.3 Depressuring systems

Controlled depressuring of the vessel reduces internal pressure and stress in the vessel walls. It also guards against the potential of adding fuel to the fire should the vessel rupture. Depressurization of the leak source also reduces the fire duration.

The design of depressuring systems should recognize the following factors.

a) Manual controls near the vessel may be inaccessible during a fire.

b) Failure position (i.e., open, closed or last position) of the depressuring valve is selected in order to

  — avoid event escalation,

  — prevent exceeding the flare capacity in the event of an instrument air failure,

  — avoid environmental excursions.

c) Early initiation of depressuring is desirable to limit vessel stress to acceptable levels commensurate with the vessel wall temperature that can result from a fire.

d) Safe disposal of vented streams should be provided.

e) Since depressuring systems/procedures can fail, no credit for the depressuring system is recommended when pressure-relief devices are being sized for fire exposure.

Further information on depressuring is provided in 5.15.6, 5.20, and 7.1.3.

5.15.4.4 Earth-covered storage

Covering a pressure vessel with earth is another effective method of limiting heat input.

5.15.4.5 Limiting fire areas with diversion walls

Diversion walls can be provided to deflect vessel spills from other vessels.

5.15.5 External insulation

5.15.5.1 General

Credit for thermal insulation is typically not taken because it usually does not meet the fire-protection insulation requirements given in 5.15.5.2 through 5.15.5.4. If these requirements are met, a reduction in fire-heat input can be obtained by using the environment factor, F, (Table 6) and Equation (13) or by calculating the actual heat flow through insulation taking the conductivity and thickness into consideration. Where credit is taken for the reduction of heat input as a result of fire protection insulation, this should be documented in the relief system design basis information, see 4.4.

5.15.5.2 Installation considerations for external insulation systems

The designer should be certain that any system of insulating materials permits the basic insulating material to function effectively at temperatures up to 900 °C (approx.1 660 °F) during a fire. This period of exposure can be for up to 2 h, depending on the adequacy of firefighting provisions, the accessibility of equipment, and the degree of skill and training of the firefighting group. This consideration is especially pertinent to newer installations using foamed or cellular plastic materials that have excellent properties at operating conditions but that (unless they were specially treated and pre-tested) have melted, vaporized or otherwise been destroyed at temperatures as low as 260 °C (500 °F). Although jacketing and coatings can burn off or disintegrate, the insulation system should retain its shape, most of its integrity in covering the vessel and its insulating value. Corrosion under insulation should be considered when installing insulation. See API Publ 2218 [18] for further guidance.

The finished installation should ensure that fire protection insulation is not dislodged when it is subjected to the high-pressure water streams used for fire fighting, such as streams from hand lines or monitor nozzles, if installed. Some criteria that should be considered include the ability of the protected system to withstand direct-flame impingement. Fire insulation, or insulation that is part of a composite system, should be capable of withstanding an exposure temperature of 900 °C (approx. 1 660 °F) for up to 2 h. Insulation system materials selection should consider equipment metallurgy while providing required jacket integrity at fire-water pressures and fire temperatures. Stainless-steel jacketing and banding have demonstrated satisfactory performance in fire situations. On the other hand, jacketing systems that use aluminium exclusively have not demonstrated satisfactory performance. Insulation materials that may decompose during fires should be avoided or suitably protected with layered composite systems.

5.15.5.3 Physical properties of insulation systems

The value of thermal conductivity used in calculating the environmental-factor credit for insulation should be the thermal conductivity of the insulation at the mean temperature between 904 °C (1 660 °F) and the process temperature expected at relieving conditions (see 5.15.5.4). If reasonably possible, the variation in conductivity due to service and maintenance practices from known laboratory values should be taken into account. Where multiple-layer insulating systems consist of different materials, the physical characteristics of each material under the expected temperature conditions should be examined. Typical values of thermal conductivity for various insulating materials appear in Table 7.

Table 7 — Thermal conductivity values for typical thermal insulations

Average temperature of insulation °C (°F)	Thermal conductivity for selected material W/mK (Btu·in/h·ft²·°F)						
	Calcium silicate type I [23]	**Calcium silicate type II** [23]	**Mineral fiber mesh blanket/block** [a] [25], [26], [28]	**Cellular glass type I Gr 2** [24]	**Molded expanded perlite block** [27]	**Lightweight cementitious** [b] [58]	**Dense cementitious** [b] [58]
−18 (0)	—	—	—	0,045 (0,31)	—	0,519 (3,6)	1,760 (12,2)
38 (100)	—	—	0,039 (0,27)	0,053 (0,37)	—	0,519 (3,6)	1,731 (12,0)
93 (200)	0,065 (0,45)	0,078 (0,54)	0,049 (0,34)	0,063 (0,44)	0,079 (0,55)	0,519 (3,6)	1,702 (11,8)
149 (300)	0,072 (0,50)	0,084 (0,58)	0,063 (0,44)	0,075 (0,52)	0,087 (0,60)	0,519 (3,6)	1,673 (11,6)
204 (400)	0,079 (0,55)	0,088 (0,61)	0,079 (0,55)	0,091 (0,63)	0,095 (0,66)	0,519 (3,6)	1,659 (11,5)
260 (500)	0,087 (0,60)	0,092 (0,64)	0,101 (0,70)	—	0,107 (0,74)	0,519 (3,6)	1,630 (11,3)
315 (600)	0,095 (0,66)	0,97 (0,67)	0,128 (0,89)	—	0,115 (0,80)	0,519 (3,6)	1,615 (11,2)
371 (700)	0,102 (0,71)	0,101 (0,70)	0,163 (1,13)	—	0,127 (0,88)	0,519 (3,6)	1,587 (11,0)
427 (800)	—	0,105 (0,73)	—	—	—	0,519 (3,6)	1,572 (10,9)
482 (900)	—	0,108 (0,75)	—	—	—	0,519 (3,6)	1,543 (10,7)
538 (1 000)	—	0,111 (0,77)	—	—	—	0,519 (3,6)	1,514 (10,5)
593 (1 100)	—	—	—	—	—	0,519 (3,6)	1,486 (10,3)
649 (1 200)	—	—	—	—	—	0,519 (3,6)	1,471 (10,2)
	Maximum temperature for use as insulation °C (°F)						
	649 (1 200)	927 (1 700)	649 (1 200)	[c]	[c]	approx. 870 (1 600)	approx. 1 090 (2 000)

[a] “Mineral fiber blanket/block” comprises rock, slag, or glass processed from the molten state into fibrous form. The thermal conductivities shown in the table are the highest values for the various forms of the insulation suitable for the maximum use temperature indicated.

[b] Thermal conductivities for lightweight and dense cementitious materials are approximate.

[c] Maximum use temperature not given in ASTM C552 [24] and ASTM C610 [27].

5.15.5.4 Calculation of environmental factor for external insulation

Limiting the heat input from fires by external insulation reduces both the rise of the vessel-wall temperature and the generation of vapour inside the vessel. Insulation can also reduce the problem of disposing of the vapours and the expense of providing an exceptionally large relieving system to conduct the effluent to a point of disposal.

If an external insulation system is designed to limit fire heat input, it should conform to the insulation considerations of 5.15.5.2.

If insulation or fireproofing is applied, the heat absorption can be computed by assuming that the outside temperature of the insulation jacket or other outer covering has reached an equilibrium temperature of 904 °C (1 660 °F). With this temperature and the operating temperature for the inside of the vessel, together with the thickness and conductivity of the fire-protection coating, the average heat transfer rate to the contents can be computed. It should be kept in mind that the thermal conductivity of the insulation increases with the temperature and that a mean value should be used.

For insulated vessels, the environment factor (see Table 6) for insulation is given by Equation (13):

In SI units:

$$F = \frac{k(904 - T_\mathrm{f})}{66\ 570\delta_\mathrm{ins}} \tag{13}$$

In USC units:

$$F = \frac{k(1\ 660 - T_\mathrm{f})}{21\ 000\delta_\mathrm{ins}}$$

where

k is the thermal conductivity of insulation at mean temperature, expressed in W/m·K (Btu·in/h·ft^2·°F);

δ_ins is the thickness of insulation, expressed in metres (inches);

T_f is the temperature of vessel contents at relieving conditions, expressed in °C (°F).

If pressure-relief facilities are designed taking credit for insulation ($F < 1$) and the insulation is removed at a later time, the pressure-relief device sizing should be rechecked using $F = 1{,}0$ to assure the relief device is adequate for the new condition.

5.15.6 Vapour depressuring

Before the relief requirement is calculated for conditions caused by fire, 4.3.14 and 7.1.3 should be reviewed. In connection with fire protection, particularly in higher-pressure services, the designer should consider vapour depressuring facilities (see 5.20). Unless special provisions are made, a pressure-relief valve cannot provide depressuring; it merely limits the pressure rise to a given value under emergency conditions. In evaluating a vapour-depressuring system for fire load, it is particularly worthwhile to consider the possibility of a fire occurring around a vessel that contains both liquid and vapour. The unwetted portion of the vessel will probably reach a temperature at which the strength of the material is reduced. In this instance, the pressure-relief valve does not protect against rupture; whereas, a vapour-depressuring system can reduce the pressure to a safe level.

5.15.7 Air-cooled exchangers

5.15.7.1 General

The problem of heat input to air-cooled coolers and condensers on fire exposure shall be taken into account. Although the material in 5.15.7.1 through 5.15.7.4 is offered as a guide, the individual circumstances involved in each situation should be considered.

Air-cooled exchangers are unique, because, unlike shell-and-tube units, their heat-transfer surface is exposed directly to the fire. They are designed for ambient inlet-air conditions and they rapidly lose all cooling and condensing ability when they are exposed to fire-heated air. Assuming that the exchangers are treated as vessels (see 5.15.2), the relieving load can be calculated using the wetted bare-tube area of the tube bundle exposed to radiation from the fire as a basis for establishing the area term. The bare-tube area is used instead of the finned-tube area because most types of fins are destroyed within the first few minutes of exposure to fire. Heat input due to convective heat transfer may be neglected.

The calculation of the wetted bare-tube area exposed to radiation from the fire depends on the location of the exchanger relative to the potential fire and the exchanger service. As a general rule, it is necessary to consider only that portion of the bare surface on air-cooled exchangers located within the fire-risk area being evaluated in the calculation of fire loads. This would normally exclude all air-cooled exchangers located directly above pipe racks, since the area under pipe racks is not normally included within the boundaries of fire-risk areas. Guidelines for specific services are provided in 5.15.7.2 through 5.15.7.4.

5.15.7.2 Gas cooling service

It is not necessary to consider the bare area of air-cooled exchangers in gas-cooling service in the calculation of fire loads, since there is no associated vapour generation and the tubes are likely to fail due to overheating.

5.15.7.3 Condensing service

It is not necessary to consider the bare area for air-cooled condensers, whether partial or total condensing, as long as both of the following conditions are satisfied.

a) The tubes are self-draining.

b) There is no control valve or pump connected directly to the condenser liquid outlet that would prevent liquids from draining during the fire.

The reason for this is that, in the event of a fire, condensation stops and any residual condensate drains freely to the downstream receiver.

If the conditions specified above are not met, the condenser shall be treated as a liquid cooler for the purposes of estimating fire loads (see 5.15.7.4).

5.15.7.4 Liquid cooling service

For liquid coolers and for condensers not covered by 5.15.7.3, the wetted area shall be the bare area of the tubes located within the fire-risk area and within 7,6 m (25 ft) of grade (or any other surface at which a major fire could be sustained, such as a solid platform). For tubes located higher than 7,6 m (25 ft) above grade (or other surface at which a major fire could be sustained), the wetted area shall be taken as zero for forced draft units (the tubes are shielded from radiant-heat exposure by the fan hood) and as the projected area (length times width) of the tube bundle for induced draft units. In calculating the heat absorption due to fire exposure, use Equations (6) and (7), applying an exponent of 1,0 to the wetted area term.

5.15.7.5 Fire mitigation alternatives

The fire case can result in extremely large relief loads, particularly for liquid-filled air-coolers. Installation of sufficient relief area for the fire case can result in the entire contents of the air-cooler being vented in a few seconds. After venting, the tubes will no longer be wetted resulting in their prompt failure in the fire. Further, the large relief load can significantly impact the design of the discharge system (i.e., knockout drum and flare).

Air-coolers essentially consist of piping with inlet and outlet manifolds. It is the convention not to consider sizing pressure-relief devices for piping when considering the fire scenario. Instead of pressure relief, fire protection, equipment isolation and other means are employed to mitigate the consequences of piping exposed to fire. Similarly, mitigation options can be considered in lieu of a pressure-relief device for air-coolers when considering the fire scenario. The following are guidelines to mitigate the fire case for air-coolers.

a) The air-cooler should not be located above equipment containing or transporting large amounts of flammable liquids. Equipment in this classification includes pumps, heat exchangers, surge drums, reboilers and accumulators, but rack piping can be normally excluded.

b) All grading below air-coolers should be sloped so that a pool fire does not occur below the air-cooler.

c) The air-coolers should be located either at the ends of a process unit or as far distant as possible from other liquid full equipment.

d) If the location criteria cited above cannot be met, an automatic water deluge system should be considered to cool the tubes if a fire should occur. Alternatively, a means to isolate the air-cooler from large inventories of liquid during a fire should be considered. The use of remotely or automatically activated valves is the preferred isolation method. Manual isolation can also be considered, provided the valves are in a location that is accessible during a fire.

e) In some cases, the air-cooler cannot readily be isolated from major inventories of flammable or combustible liquids (e.g. if the air-cooler is located between a tower and reflux drum). In these cases, it is prudent to have the air-cooler free-draining towards the drum or column, thereby minimizing the wetted surface area exposed to a fire.

5.16 Jet fires

Protection from jet-fire exposure is typically addressed through means other than pressure-relief devices because failure often occurs due to localized overheating for which a pressure-relief device is ineffective.

Examples of different jet-fire characteristics are unpredictable flame-impingement points, significantly increased heat loading to the vessel's wetted and unwetted surfaces, mainly due to higher flame temperatures [local instantaneous heat fluxes within jet fires as high as 300 kW/m^2 (94 500 $Btu/ft^2{\cdot}h$) have been reported] and the effect of the jet velocity on fire-water deluge/monitor-vessel coverage.

Jet fires can occur when almost any combustible/flammable fluid under pressure is released to atmosphere. The primary concern with jet-fire impingement is that the equipment can fail due to intense, localized overheating of the metal wall where the jet fire impinges. Failure can occur even without increasing the pressure in the equipment to the set point of the relief device. This is due to the localized nature of heating whereby the bulk fluid temperature might not increase appreciably. Hence, a relief device (i.e., overpressure protection device) might not prevent vessel failure from jet fire impingement.

Instead of a pressure-relief system, protection against jet fires focuses on prevention of leaks through proper maintenance and/or mitigation systems such as fireproofing, depressuring systems, isolation of leaks, equipment and/or flange orientation and minimization and emergency response. Installation of fireproofing provides additional time (an impinging jet fire can cause vessel failure in less than 5 min, depending on the vessel's wall thickness and material) but might not prevent failure as the fireproofing can be eroded by the momentum effects of the jet fire. Depressuring systems are discussed in 5.20. Finally, unlike a pool fire, a jet fire can, in essence, be “turned off” through isolation and depressurization of the jet fire source (i.e., leaking pipe, vessel or other equipment).

5.17 Opening manual valves

The following applies when a manual valve is inadvertently opened, causing pressure build-up in a vessel. The vessel should have a pressure-relief device large enough to pass a rate equal to the flow through the open valve; less credit is given for alternative vessel outlets that can reasonably be expected to be operational. The manual valve should be considered as passing its capacity at a full-open position with the pressure in the vessel at relieving conditions. Volumetric or heat-content equivalents may be used if the manual valve admits a liquid that flashes or a fluid that causes vaporizing of the vessel contents. It is necessary to consider only one inadvertently opened manual valve at a time.

5.18 Electric power failure

5.18.1 General

Determination of relieving requirements resulting from power failures requires a careful plant or system analysis to evaluate what equipment is affected by the power failure and how failure of the equipment affects plant operation. Careful study and consideration should be given to the material presented in 4.3.5 and 4.3.6. Automatic standby is an excellent method for maximizing the unit's on-stream time, minimizing unit upsets and ensuring unit production rates. However, the circuitry, sequences and components involved are not yet considered sufficiently reliable to permit credit for them in establishing individual relieving requirements.

5.18.2 Analysis

Electric power failure should be analysed in the following three ways:

a) as a local power failure in which one piece of equipment is affected;

b) as an intermediate or partial power failure in which one distribution centre, one motor control centre, or one bus is affected;

c) as a total power failure in which all electrically operated equipment is simultaneously affected.

The effects of a local power failure are easily evaluated when individual pieces of equipment, such as pumps, fans and solenoid valves, are affected. Most of these effects are covered in other clauses of this International Standard. Once the upsetting cause is resolved, the relieving requirements can be determined from these clauses. For example, a pump failure can cause a loss of cooling water or a loss of reflux. For the effects of loss of reflux and/or cooling water and relieving requirements, see 5.6. Loss of absorbent is covered in 5.7.

Intermediate or partial power failure can cause more serious effects than either of the other two types of failure. Depending on the method of dividing various pumps and drivers among the electrical feeders, it is possible to lose all the fans at an air-cooler at the same time that the reflux pumps are lost. This can, then, flood the condenser and can void any credit normally taken for the effect of natural convection of the air-cooled condenser.

Total power failure requires additional study to analyze and evaluate the combined effects of multiple equipment failures. Special consideration should be given to the effect of the simultaneous opening of relief valves in several services, particularly if the relief valves discharge into a closed header system.

5.19 Heat-transfer equipment failure

5.19.1 Requirements

Heat exchangers and similar vessels should be protected with a relieving device of sufficient capacity to avoid overpressure in case of an internal failure. This statement defines a broad problem but also presents the following specific problems:

a) type and extent of internal failure that can be anticipated;

b) determination of the required relieving rate if overpressure of the low-pressure side of the exchanger and/or connected equipment occurs as a result of the postulated failure;

c) selection of a relieving device that reacts fast enough to prevent the overpressure;

d) selection of the proper location for the device so that it senses the overpressure in time to react to it.

Provision of overpressure protection for the heat exchanger and associated pipework does not remove the need for a process hazard analysis to consider the wider process implications of any inter-stream leakage.

These tube-rupture guidelines were established without considering a chemical reaction in the event that the high-pressure fluid mixes with the low-pressure fluid. If the heat exchanger contains reactive chemicals, then a careful evaluation shall be performed to ensure that the reactive situation does not result in the pressure exceeding the low-pressure side's corrected hydrotest pressure (see 3.21 and 4.3.2).

5.19.2 Pressure considerations

Complete tube rupture, in which a large quantity of high-pressure fluid flows to the lower-pressure exchanger side, is a remote but possible contingency. Minor leakage can seldom overpressure an exchanger during operation, however such leakage occurring where the low-pressure side is closed-in can result in overpressure. Loss of containment of the low-pressure side to atmosphere is unlikely to result from a tube rupture where the pressure in the low-pressure side (including upstream and downstream systems) during the tube rupture does not exceed the corrected hydrotest pressure (see 3.21 and 4.3.2). The user may choose a pressure other than the corrected hydrotest pressure, given that a proper detailed mechanical analysis is performed showing that a loss of containment is unlikely. The use of maximum possible system pressure instead of design pressure may be considered as the pressure of the high-pressure side on a case-by-case basis where there is a substantial difference in the design and operating pressures for the high-pressure side of the exchanger.

Pressure relief for tube rupture is not required where the low-pressure exchanger side (including upstream and downstream systems) does not exceed the criteria noted above. The tube rupture scenario can be mitigated by increasing the design pressure of the low-pressure exchanger side (including upstream and downstream systems), and/or assuring that an open flow path can pass the tube rupture flow without exceeding the stipulated pressure, and/or providing pressure relief.

The user may perform a detailed analysis and/or appropriately design the heat exchanger to determine the design basis other than a full-bore tube rupture. However, each exchanger type should be evaluated for a small tube leak. The detailed analysis should consider

a) tube vibration,

b) tube material,

c) tube wall thickness,

d) tube erosion,

e) brittle fracture potential,

f) fatigue or creep,

g) corrosion or degradation of tubes and tubesheets,

h) tube inspection programme,

i) tube to baffle chafing.

The basis for the analysis should be documented and maintained with the relief-system design information, see 4.4.

5.19.3 Determining the required relief flow rate

In practice, an internal failure can vary from a pinhole leak to a complete tube rupture. For the purpose of determining the required relieving flow rate for the steady-state approach, the following basis should be used.

a) The tube failure is a sharp break in one tube.

b) The tube failure is assumed to occur at the back side of the tubesheet.

c) The high-pressure fluid is assumed to flow both through the tube stub remaining in the tubesheet and through the other longer section of tube.

A simplifying assumption of two orifices may also be used in lieu of the above method, since this produces a larger relief flow rate than the above approach of a long open tube and tube stub.

The dynamic approach requires a detailed analysis to determine if a design basis smaller than a full-bore tube rupture is adequate.

In determining the relief rate, allowance should be made for any liquid that flashes to vapour either as a result of the pressure reduction or, in the case of volatile fluids being heated, because of the combined effects of pressure reduction and vaporization, as the fluid is intimately contacted by the hotter material on the low-pressure side.

For liquids that do not flash when they pass through the opening, the discharge rate through the failure should be computed using incompressible-flow equations. For vapour passing through the ruptured tube opening, compressible-flow theories apply. Typical steady-state equations for evaluating the flow rate through an orifice or an open tube end, for gas or non-flashing liquid service, are presented in Crane Technical Paper No. 410 [59] or other fluid flow references.

A two-phase flow method should be used in determining the flow rate through the failure for flashing liquids or two-phase fluids. The flow models developed by DIERS and others can be adapted for this purpose. Additional information concerning these models is available in References [60] and [61]. In cases where the fluid flashes at the low-pressure side of the heat exchanger, two-phase flow methods based on the homogenous equilibrium model (HEM), such as those proposed by DIERS, may be used for the flow through the tube to the break, which is assumed to be at the tubesheet. For the flow across the tubesheet to the break, the thickness of the tubesheet should be considered when determining whether single- or two-phase methods should be used. Guidelines on the minimum horizontal flow path length required for homogeneous two-phase methods to be applicable are presented in the literature on this subject. Two literature resources are available for clarification: Reference [62], specifically the section entitled Non-Equilibrium Flashing Flow, and Reference [63]. A conservative approach should be taken where any doubt exists.

Two approaches are available for determining the required size of the relief device: (a) steady-state and (b) dynamic analysis. If a steady-state method is used, the relief-device size should be based on the gas and/or liquid flow passing through the rupture. Capacity credit can be taken for the low-pressure side piping per the guidelines of 5.19.5. A one-dimensional dynamic model can be used where the approach is to simulate the pressure profile and pressure transients developed in the exchanger from the time of the rupture. These methods generally include the dynamic model of the tube-rupture relief scenario and the response time of the relief device, the accuracy of which is critical in calculating the accuracy of pressures generated. Data on the dynamic model of the tube rupture relief scenario and the response time of the relief device can be found in References [119] and [120].

This type of analysis is recommended, in addition to the steady-state approach, where there is a wide difference in design pressure between the two exchanger sides [e.g. 7 000 kPa (approx. 1 000 psi) or more], especially where the low-pressure side is liquid-full and the high-pressure side contains a gas or a fluid that flashes across the rupture. Modelling has shown that, under these circumstances, transient conditions can produce overpressure above the test pressure, even when protected by a pressure-relief device [64], [65], [66]. In these cases, additional protection measures should be considered.

5.19.4 Relief devices and locations

The design of piping around the exchanger and the location of the relieving device are both critical factors in protecting the exchanger. Both rupture disks and pressure-relief valves should be considered.

It may be necessary to locate the relieving device to be located either directly on the exchanger or immediately adjacent on the connected piping. This is especially important if the low-pressure side of the exchanger is liquid-full. In this case, the time interval in which the shock wave is transmitted to the relieving device from the point of the tube failure increases if the device is located remotely. In addition, there is a time delay for the gas to overcome the momentum of the liquid-filled low-pressure side prior to establishing a full flow through the relief path. This can result in higher transient overpressure on the exchangers before operation of the rupture disk or relief valve.

It can be impractical to protect some heat exchangers (and associated piping) by relief devices alone e.g. if there is a high pressure difference between the shell and tube sides. In these cases, different layers of protection, such as improved metallurgy, more frequent inspection and increasing the design pressure of the low pressure side (including upstream and downstream piping until the pressure is dissipated), can be necessary.

An analysis should be made of the time interval needed for the relieving device to open. The opening time for the device used should be verified by the manufacturer and should also be compatible with the requirements of the system.

5.19.5 Influence of piping and process conditions

To determine the influence of piping, either in eliminating the need for a relieving device or in reducing relieving requirements, the configuration of the discharge piping and the contents (liquid or vapour) of the low-pressure side should be considered. If the low-pressure side is in the vapour phase, full credit can be taken for the vapour-handling capacity of the outlet and inlet lines, provided that the inlet lines do not contain check valves or other equipment that could prevent backflow. If the low-pressure side is liquid-full, the effective relieving capacity for which the piping system may be credited shall be based on the volumetric flow rate of the low-pressure side liquid that existed prior to the tube rupture. However, if a detailed analysis is performed, a capacity credit may be taken for acceleration of the low-pressure side liquid.

If the piping system to the low-pressure side of heat transfer equipment contains valves, their effect on the capacity of the system when overpressure occurs should be taken into account. Valves provided only for isolation may be assumed to be fully opened. In calculating relieving-capacity credit for the piping system, one should consider the valves used for control purposes to be in a position equivalent to the minimum normal flow requirements of the specific process. However, this assumption cannot be made if the valve could automatically close because of the emergency situation.

5.19.6 Double-pipe exchangers

The two types of double-pipe exchangers are those that actually use schedule pipe as the inner tube and those that use gauge tubes, usually in the heavier gauges. Units that use schedule pipe for the inner conduit or tube are no more likely to rupture the inner pipe than any other pipe in the system. Therefore, it is not necessary to consider a complete tube rupture as requiring a provision for pressure relief. Although complete tube rupture can be unlikely, weld failures can occur, especially if the two pipes are made from dissimilar metals. The designer is cautioned to evaluate each case carefully and to use sound engineering judgment to decide whether the particular case under study represents an exception. For example, where gauge tubes are used, the designer should determine whether or not they are equivalent to schedule pipe.

5.20 Vapour depressuring

5.20.1 General

Depressuring systems can be used to mitigate the consequences of a vessel leak by reducing the leakage rate and/or inventory within the vessel prior to a potential vessel failure. More often, depressuring systems are used to reduce the failure potential for scenarios involving overheating (e.g. fire). When metal temperature is increased due to fire or exothermic or runaway process reactions, the metal temperature can reach a level at which stress rupture can occur. This can be possible even though the system pressure does not exceed the maximum allowable accumulation. In this case, depressuring reduces the internal stress, thereby extending the life of the vessel at a given temperature. A typical target is to provide comparable benefit as fireproofing (designed to maintain the integrity for 2 h in a fire) or to maintain system integrity until the acceptance criteria for rupture is reached. In order to be effective, the depressuring system shall depressure the vessel such that the reduced internal pressure keeps the stresses below the rupture stress. In general, the depressurization rates should be maximized within the total flare system capacity (i.e. the sum of all required simultaneous depressurization and relief rates should be close to or equal to the flare/vent system capacity; for further information, see 7.1).

If a depressurization system is installed to protect vessels and/or piping against fire, the need for passive fire protection is determined by the capacity of the depressurization device, the type, size and intensity of the fire, the availability of firewater and fire fighting equipment, the type and layout of the drain system. The need for passive fire protection also depends on wall thickness, the vessel/pipe material and the prevailing acceptance criteria for the specific installation. The following should be considered when designing/specifying the depressurization system:

— rupture time (time to escape, time for rescue actions);

— rupture pressure of vessels (escalation, fragmentation);

— rupture pressure of pipes (escalation);

— total release of flammables (escalation);

— instantaneous release rate (sudden increase in fire size during evacuation or rescue);

— loss of production, reputation and rebuild cost;

— damage to internals of equipment (e.g. trays, packing supports), entrainment of packing or catalyst into the depressurization system, brittle failure due to cooling.

The above can vary from installation to installation, i.e. it may be different for a low-manned, remote installation compared to an installation located in populated areas; whether the fluid is LPG, gas or oil; whether the fluid is toxic or not, etc. It may also be different from one user to the other and from one country to another. Since the consequences of a vessel rupture [fragmentation and possible boiling-liquid expanding-vapour explosion (BLEVE)] normally are larger than for a pipe rupture, pipe rupture can be more acceptable than a vessel rupture.

EXAMPLE If a pool fire exposes the unwetted wall of a large [25,4 mm (1 in) wall thickness] vessel fabricated from ASTM A 515 Grade 70 carbon steel, it will take about 15 min to heat the vessel walls to around 649 °C (1 200 °F), as shown in Figure 1. At this temperature, rupture due to overheating is imminent as the material's allowable stress [120 MPa (17 500 psi) at ambient temperature] approaches its rupture stress, as given in Figure 2. In contrast, if the vessel is depressurized within the 15 min heat-up time to, say, 50 % of the initial pressure (i.e., half the initial internal stress), then the time to rupture would increase to about 2 h to 3 h (see Figure 2).

If vapour depressuring is required for both fire and process reasons, the larger requirement should govern the size of the depressuring facilities.

A vapour-depressuring system should have adequate capacity to permit reduction of the vessel stress to a level at which stress rupture is not of immediate concern. For pool-fire exposure and with heat input calculated from Equations (6) or (7), this generally involves reducing the equipment pressure from initial conditions to a level equivalent to 50 % of the vessels design pressure within approximately 15 min. This criterion is based on the vessel-wall temperature versus stress to rupture and applies generally to carbon steel vessels with a wall thickness of approximately 25,4 mm (1 in) or more. Vessels with thinner walls generally require a somewhat faster depressuring rate. The required depressuring rate depends on the metallurgy of the vessel, the thickness and initial temperature of the vessel wall and the rate of heat input.

Many light hydrocarbons chill to low temperatures as pressure is reduced. Design and depressuring conditions should consider this possibility.

Depressuring is assumed to continue for the duration of the emergency. The valves should remain operable for the duration of the emergency or should fail in a full-open position. Otherwise, fireproofing of the control signal and valve actuator or other protective measures (e.g. locate the valve, valve actuator and control signals outside the fire area) to assure the appropriate operability of the valve during a fire requires consideration.

Emergency depressuring for the fire scenario should be considered for large equipment operating at a gauge pressure of 1 700 kPa (approx. 250 psi) or higher. The effect of heat input to process vessels is discussed in 5.15.2 and 5.20.2. Depressuring to a gauge pressure of 690 kPa (100 psi) is commonly considered when the depressuring system is designed to reduce the consequences from a vessel leak.

The reduced pressure permits somewhat more rapid control of the situation in which the source of fire is the leakage of flammable materials from the equipment being depressurized.

Depressuring criteria other than those given above can be used depending upon the specific circumstances and user-defined requirements. For example, if there is a reactive hazard or other exceptional hazard that can cause loss of containment due to over-temperature, emergency depressuring can be appropriate for equipment designed for a wider range of pressures than that noted above. Refer to 5.13 for guidance on how to estimate the vent size and temperature rise in a reactive system.

Mitigation measures for equipment that can be exposed to a fire often include equipment design, equipment layout, structural fireproofing, area-drainage design, firewater-system design, emergency-response capabilities, emergency isolation, and/or emergency depressuring. It is necessary that the user assess how effective the site-specific mitigation measures can be when determining the appropriate emergency depressuring criteria.

5.20.2 Vapour flows

5.20.2.1 General

To reduce the internal pressure in equipment involved in a fire, vapour should be removed at a rate that compensates for the following occurrences:

a) vapour generated from liquid by heat input from the fire;

b) vapour expansion during pressure reduction;

c) liquid flash due to pressure reduction. (This factor applies only when a system contains liquid at or near its saturation temperature).

The total vapour load for a system to be depressurized may be expressed as the sum of the individual occurrences for all equipment involved. Thus, in terms of the loads in list items a) through c), the total mass, m, equals item a) plus item b) plus item c) as given by Equation (14):

$$\dot{m} \cdot t = \sum_{i=1}^{x} (q_{\mathrm{m,f}} \cdot t)_i + \sum_{i=1}^{x} (q_{\mathrm{m,d}} \cdot t)_i + \sum_{i=1}^{x} (q_{\mathrm{m,v}} \cdot t)_i \qquad (14)$$

NOTE The variables for all equations in 5.20 are defined in 5.20.3.

The combined expression, $\dot{m} \cdot t$, is used because $\dot{m}$ represents a flow rate per unit of time, and some of the noted vapour quantities are mass quantities that are not influenced by time, namely $q_{\mathrm{m,v}} \cdot t$ and $q_{\mathrm{m,d}} \cdot t$ (the vapour loads from density change and liquid flash). If the system to be depressurized includes more than one vessel, the vapour quantities for each vessel under all three occurrences should be calculated, especially if different relative molecular masses, latent heats, insulation thicknesses and vaporization temperatures are involved. The average relative molecular mass and temperature for $\dot{m}t$ (the total vapour relieved from the whole system) should be calculated from the total individual vapour relative molecular masses and vapour temperatures involved. The vapour loading on the depressuring system for each of the terms in Equation (14) is described in 5.20.2.2 through 5.20.2.4.

5.20.2.2 Vapour from fire-heat input

The heat input to equipment during a fire is generally calculated in accordance with 5.15.2; however, the following modifications and limitations can be used to compute loads for a vapour depressuring and pressure-relieving system under fire conditions.

a) The extent of an assumed fire zone is a function of the design and installation features that permit confining a fire within a given area (see 4.3.14). Although the size of the assumed fire zone can vary, experience generally indicates that a fire that can be confined to approximately 232 m^2 (2 500 ft^2) of plot area will not affect the design of the main relief headers in processing areas where a depressuring flow discharges into the same relief header.

b) Additional insulation or an increase in the thickness of insulation on individual vessels may also be considered as a means of reducing vapour generation resulting from exposure to fire.

c) During a fire, all feed and output streams to and from the system to be depressurized and all internal heat sources within the process are assumed to have ceased. Thus, the vapour generation is a function only of the heat absorbed from the fire and the latent heat of the liquid.

To calculate the vapour load generated by fire, the fire should be assumed to be in progress throughout the depressuring period. The mass, m_f, of vapour generated by the fire during the depressuring interval in a vessel, i, of the system can be determined by Equation (15):

$$(m_\mathrm{f} \cdot t)_i = t(Q/L)_i \tag{15}$$

This calculation should be repeated for all vessels in the system if significant differences in vapour and liquid properties are involved.

5.20.2.3 Vapour from density change and liquid flash

The calculations of vapour loads caused by vapour-density change and those that result from liquid flash cannot be completely separated. To determine the vapour quantities contributed by these causes, It is necessary to know the liquid inventory and vapour volume of the system. This includes all liquid and vapour in any directly connected facilities outside the fire area that cannot be isolated under fire conditions, as well as all liquid and vapour contained in equipment located in the assumed fire area. Although liquid inventory and vapour volume depend on plant design, the following assumptions may be made to estimate these values.

a) The liquid inventory of fractionating columns can be estimated as the normal column bottom and draw-off tray capacity, plus a hold-up per tray, equal to the weir height plus 50 mm (2 in), or its design quantity, if known.

b) Normal operating levels may be used as the basis for computing the inventory of accumulators.

c) To obtain an initial, rapid approximation for standard shell-and-tube heat exchangers, one-third of the total shell volume should be assumed to be occupied by the tube bundle. For condensers and heat exchangers in vaporizing service, 80 % of the volume involved should be assumed to be vapour. The remainder should be assumed to be liquid.

d) All liquid in heaters should be included in the estimate, regardless of temperature. If the heater is in vaporizing service, one should assume 80 % of the tube volume past the normal point of vaporization to be vapour.

Only after the vapour and liquid volumes in the system have been determined can one estimate the respective loadings they contribute to depressuring.

One can determine the mass of vapour to be removed from a given vapour space in a vessel, i, to compensate for the reduced vapour density at the lower pressure by using Equation (16) or (17):

In SI units:

$$(q_{\mathrm{m,d}} \cdot t)_i = 0{,}120\ 5 \cdot V_i \left[\left(\frac{p \cdot M}{Z \cdot T} \right)_\mathrm{a} - \left(\frac{p \cdot M}{Z \cdot T} \right)_\mathrm{b} \right]_i \tag{16}$$

In USC units:

$$(q_{\text{m,d}} \cdot t)_i = 0{,}093\ 2 \cdot V_i \left[\left(\frac{p \cdot M}{Z \cdot T} \right)_{\text{a}} - \left(\frac{p \cdot M}{Z \cdot T} \right)_{\text{b}} \right]_i \tag{17}$$

where the subscript "a" represents the higher-pressure condition and "b" represents the lower-pressure condition.

Note that V_i is assumed not to increase significantly as a result of liquid flash. This calculation should be repeated for each vessel in the system if different vapour properties are involved.

Since the calculation of the vapour load caused by liquid flash depends on liquid quantity and liquid properties in the system, the preceding data are also valid for this calculation. In systems that contain liquid at saturation conditions, the temperature of the liquid should be reduced to obtain the required reduction in pressure. To reduce pressure, one can remove vapour at a rate equal to the vapour-generation rate created by heat input from the fire to compensate for the flash vaporization of some liquid. Without this allowance for flash vaporization, the required reduction in pressure is not possible. It is necessary to consider only the liquid inventory that is at or near its saturation temperature for liquid flash. Two methods are shown for calculating the rate at which it is necessary to withdraw vapour in order to reduce the temperature within a time interval, t, to a point at which the corresponding liquid vapour pressure equals the desired final pressure. The first method applies only to relatively pure chemicals and to narrow-boiling-range hydrocarbons; the second covers liquids that consist of mixtures of hydrocarbons with a wider boiling range. For pure chemicals or hydrocarbons with a narrow boiling range, the amount of liquid flash in a vessel, i, of the system may be conservatively approximated by equating the heat of the flashed vapour with the heat loss of the average liquid quantity as shown in Equation (18):

$$(q_{\text{m,v}} \cdot t)_i \cdot \lambda_i \approx \left[(q_{\text{m,a}} \cdot t) - \frac{Q_i \cdot t}{2\lambda_i} - \frac{(q_{\text{m,v}} \cdot t)_i}{2} \right] Cp_i \left(T_{\text{a}} - T_{\text{b}} \right)_i \tag{18}$$

Rearranging Equation (18) as follows in Equation (19) yields the amount of liquid flash:

$$(q_{\text{m,v}} \cdot t)_i \approx \left[(q_{\text{m,a}} \cdot t)_i - \frac{Q_i \cdot t}{2\lambda_i} \right] \cdot \left[\frac{2Cp_i \left(T_{\text{a}} - T_{\text{b}} \right)_i}{2\lambda_i + Cp_i \left(T_{\text{a}} - T_{\text{b}} \right)_i} \right] \tag{19}$$

NOTE $(q_{\text{m,a}} \cdot t)$ is used only for consistency, and $q_{\text{m,a}}$ has no physical significance.

If a more rigorous calculation is desired, the same approach may be applied in stepwise form.

Equation (19) cannot be used for liquids consisting of a mixture of hydrocarbons that have a wide boiling range because the liquid properties and composition change as the liquid is vaporized. If more accurate fluid data are not available, a series of simplified adiabatic flash calculations should be made between the initial pressure and the final pressure, while neglecting the simultaneous fire effect. The simplified adiabatic flash calculation is a stepwise procedure that, by repeatedly applying Equation (20), yields a mass fraction flashed from the liquid quantity that was originally in the system during the required pressure decrease. This process assumes that the vapours flashed in each step are totally removed from the system to be depressurized before the next step occurs. The correction for the fire is made in Equation (21) in which the average of the remaining liquid quantity is used (that is, the original liquid quantity in the system minus half of the quantity vaporized by fire during the total depressuring period) instead of the total liquid quantity that was originally in the system. This compensates to some extent for neglecting the fire-vaporization effects on the composition for each flash step.

To determine the approximate amount of liquid vaporized from a mixture, an equilibrium-phase diagram is required and a graphical solution employing n steps is employed. The procedure uses Equation (20):

$$\left(\Delta T_n \right)_i = \left[\frac{L_n \left(\Delta q_{\text{m,v}} \cdot t \right)_n}{\left(q_{\text{m,L}} \cdot t \right)_{n-1} - \left(\Delta q_{\text{m,v}} \cdot t \right)_n \left(Cp \right)_n} \right]_i \tag{20}$$

5.20.2.4 Example for vapour from density change and liquid flash

Figure 3 is an example of an equilibrium-phase diagram for a given liquid having the following evolution:

a) Initial conditions:

— 0 % of the liquid is vaporized;

— $T_a = 256$ °C (492 °F);

— absolute pressure $p_a = 2\,861$ kPa (415 psi).

b) The conditions after step $n = 1$ are as follows:

— 5 % of the liquid is vaporized and 95 % of the liquid remains;

— $T_1 = 252$ °C (486 °F);

— absolute pressure $p_1 = 2\,496$ kPa (362 psi).

c) The development of the diagram continues stepwise until depressuring is completed at the following conditions after step $n = 5$ at item b):

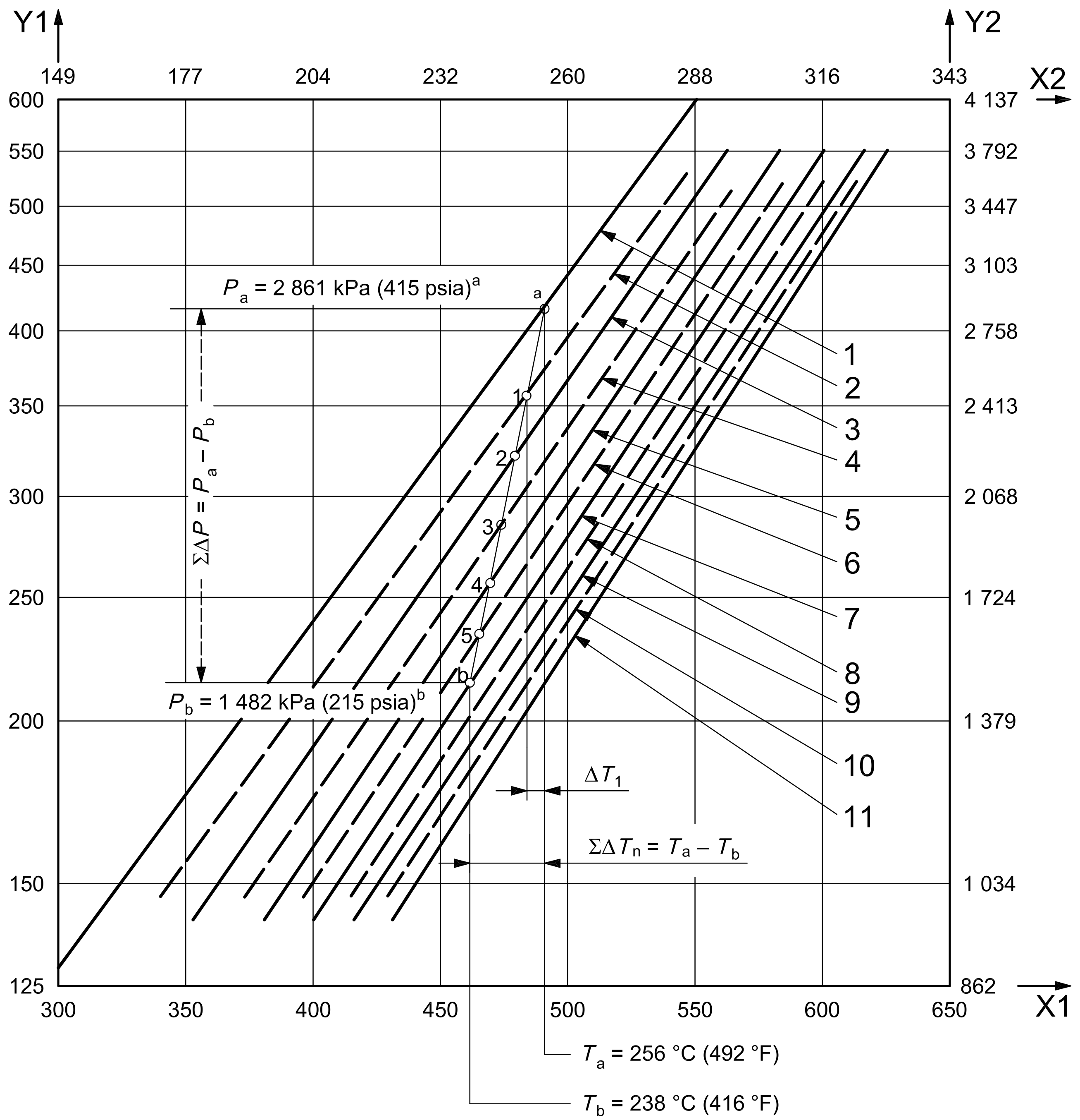
— 30 % of the liquid is vaporized and 70 % of the liquid remains;

— $T_b = 238$ °C (461 °F);

— absolute pressure $p_b = 1\,482$ kPa (215 psi).

For convenience, the mass percent vaporized was assumed to be equal to the volume percent vaporized. By assuming that an incremental part of the liquid (e.g. 5 %) was vaporized during each step, the change in the liquid temperature can be computed using Equation (20). Since the remaining liquid has a saturation temperature and pressure along the 5-%-vaporized line of the phase diagram and the temperature change has been determined using Equation (20), the pressure change is also known. The process is repeated in incremental steps until the pressure, p_b, at the end of the depressuring period is obtained. In Figure 3, the desired end pressure is reached when the mass fraction, X_i, of the initial liquid in the vessel, i, that has been vaporized is ≈ 0,30. Substituting this value of X_i into Equation (21) for the last term in Equation (19) gives the estimated mass of liquid flashed as a result of the depressuring from the vessel, i, of the system during a simultaneous fire.

$$(q_{\mathrm{m,v}} \cdot t)_i \approx \left[(q_{\mathrm{m,a}} \cdot t)_i - \frac{Q_i \cdot t}{2L_i} \right] w_i \tag{21}$$



Key

X1 liquid temperature, expressed in degrees Fahrenheit
X2 liquid temperature, expressed in degrees Celsius
Y1 pressure, Psig
Y2 pressure, kPa

1 0 % vapourized
2 5 % vapourized
3 10 % vapourized
4 15 % vapourized
5 20 % vapourized
6 25 % vapourized
7 30 % vapourized
8 35 % vapourized
9 40 % vapourized
10 45 % vapourized
11 50 % vapourized

[a] At start.

[b] At end.

Figure 3 — Equilibrium phase diagram for a given liquid

5.20.3 Nomenclature

The variables used in the equations throughout 5.20 are defined as follows:

C_p is the average specific heat of the liquid, expressed in kJ/kg·K (Btu/lb·°R);

L is the average latent heat of the liquid, expressed in kJ/kg (Btu/lb);

m is the mass of liquid or vapour, expressed in kg (lb);

$\dot{m}$ is the mass flow rate per unit time;

M is the relative molecular mass of the vapour;

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

q_{m} is the vapour mass flow rate, expressed in kg/h (lb/h);

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

T is the absolute temperature of the liquid or vapour, expressed in K (°R);

t is the depressuring time interval, expressed in hours (usually assumed to be 0,25 h);

V is the volume available for the vapour, expressed in m^3 (ft^3);

w is the mass fraction of the initial liquid in the system vaporized as a result of depressuring, dimensionless;

Z is the compressibility factor, dimensionless;

Δ represents a difference, e.g. as in $\Delta T_n = T_{n-1} - T_n$.

subscripts:

a is the original condition at the start of the depressuring time interval, assumed to be the saturated vapour-liquid equilibrium condition with respect to temperature and pressure;

b is the depressurized condition at the end of the depressuring time interval;

d relates to the density change of the vapour due to pressure reduction;

f relates to vaporization from the fire;

i relates to an individual vessel of the system if more than one vessel is involved and requires separate consideration because of differing fluid properties, insulation for fire effect, or related factors;

L relates to liquid;

n is the nth depressuring step of many steps between the original condition and the depressurized condition;

n–1 is the depressuring step preceding step n;

v relates to liquid flash or vapour generated from pressure reduction;

x is the total number of vessels in the depressuring system.

5.21 Special considerations for individual pressure-relief devices

5.21.1 General

Sizing procedures for pressure-relief devices shall be in accordance with API RP 520-1 or ISO 4126.

5.21.2 Liquid-vapour mixture and solids formation

A pressure-relief device handling a liquid at vapour-liquid equilibrium or a mixed-phase fluid produces vapour due to flashing as the fluid moves through the device. The vapour generation can reduce the effective mass-flow capacity of the valve and should be taken into account. Liquid carryover can result from foaming or inadequate vapour-liquid disengaging. The designer is cautioned to investigate the effects of flow reduction or choking. Choking occurs at a point in any flowing compressible or flashing fluid where the available pressure-drop increment is totally used up by accelerating the flashing fluid. Therefore, no additional pressure difference is available to overcome the friction in the incremental line length. See API RP 520-I and References [60], [61] and [68] for further discussion on this subject.

Some fluids (e.g. carbon dioxide and wet propane) can form solids when they are discharged through the relieving device. No uniformly accepted method has been established for reducing the possibility of plugging.

5.21.3 Location of a pressure-relieving device in a normally liquid system

If valves or other devices are sized to relieve vapours caused by vapour entry or generation of vapour in a normally all-liquid system (see 5.10, 5.12, 5.13, and 5.20), care should be taken to locate the device so that it actually relieves vapour and is not required to relieve the volumetric equivalent of the vapour as liquid.

5.21.4 Multiple pressure-relief devices

5.21.4.1 Basis

The pressure-design code may limit the allowable range of set pressures for multiple pressure-relief devices. For example, see the more detailed information in EN 764-7 [5] or ASME Code, Section VIII, Division 1, Paragraph UG-125 through Paragraph UG-137, Appendix 11 and Appendix M [20].

5.21.4.2 Justification

The considerations that make a multiple pressure-relief-valve installation with staggered settings desirable include the following:

- sizing factor and valve leakage;
- pressure vessel requirements;
- inlet pressure characteristics of the pressure-relief valve;
- reactive thrust at relief;
- range of required relieving rates for various contingencies.

In sizing pressure-relief valves, the designer should explore possible sources of overpressure, establish the governing flow rate and select the required orifice area. Although the governing flow can result from a single factor or a combination of circumstances, the difficulty of anticipating simultaneous occurrences tends to encourage conservative sizing (oversizing). As the size of process units increases, the calculated area required often cannot be obtained in a single pressure-relief-valve body of rated, commercial design. Hence, multiple pressure-relief valves are needed simply to handle the required relieving rate. Minor fluctuations in the controlled vessel pressure can approach or enter the operating range of a single pressure-relief valve. This creates continuous leakage that sustains itself until the pressure in the system drops low enough to enable the spring to force the valve closed. The larger the valve, the lower its lift to handle this small flow rate, and the greater the leakage at any given lift. Chatter and seat damage often accompany this circumstance. This problem is compounded with multiple pressure-relief valves uniformly set; however, multiple pressure-relief valves with staggered settings can provide a solution. If feasible, the lowest set valve should be the smallest one that can be selected on the basis of a reasonable relieving requirement or a reasonable portion of the total requirement. The higher set valves open only under conditions that require the combined orifice areas to handle the generated flow.

5.21.4.3 Application and practice

See EN 764-7 or API RP 520-I and API RP 520-II [8] for guidance.

The use of multiple pressure-relief valves can frequently be accomplished easily and economically if permitted by the pressure-design code. In considering multiple pressure-relief-valve releases, the effects of back pressure should be evaluated with all valves releasing concurrently under that single contingency. The normal design approach is to consider that all pressure-relief valves are flowing simultaneously, whether they be staggered set valves on one vessel or several pressure-relief valves on various vessels that should also release under this same contingency. The aggregate rate of flow determines the back pressure in the system. Any increase in back pressure in the system that results from the contingency can be considered to be built-up back pressure. Where conventional valves are employed, there are back pressure limits that shall be considered. Within these limitations, it is not necessary to consider the changes in back pressure caused by flow that results from one valve opening before another as superimposed back pressure on other valves.

5.22 Dynamic simulation

Dynamic simulation can be used in pressure-relief system design to calculate transient pressure increases as indicated in 5.19 or to calculate required relief rates from individual pressure-relief devices. Conventional methods for calculating relief loads are generally conservative and can lead to overly sized relief- and flare-system designs. Dynamic simulation provides an alternative method to better define the relief load and improves the understanding of what happens during relief.

Dynamic simulation is particularly useful in analysing existing flare systems (see 7.1.4.2). A dynamic simulation of a single system is discussed below.

Dynamic simulation is an alternative calculation method for determining the relief requirements for an individual column. Dynamic simulation can be applied wherever conventional methods can be applied. Column-relief-load calculations using dynamic simulation shall follow the same rules set elsewhere in this International Standard for performing relief-load calculations using conventional calculation methods. It can be necessary to perform sensitivity analyses with respect to control response in order to identify appropriate control response. In general, no credit is taken for automatic control action unless it tends to increase the relief load.

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

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

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

6 Selection of disposal systems

6.1 General

The selection of a disposal method is subject to many factors that can be specific to a particular location or an individual unit. The purpose of a disposal system is to conduct the relieved fluid to a location where it can be safely discharged. Disposal systems generally consist of piping and vessels. All components should be suitable in size, pressure rating and material for the service conditions intended. Clause 6 outlines the general principles and design approach for determining the most suitable type of disposal system.

6.2 Fluid properties that influence design

6.2.1 Physical and chemical properties

The flash-point, flammable-limits and ignition temperatures of certain flammable liquids, gases and solids are listed in NFPA HAZ01 [37]. Additional data on the flammability characteristics of pure compounds and mixtures, in both air and atmospheres that contain varying amounts of inert gases and water vapour, are found in the U.S. Bureau of Mines Bulletin 627 [69]. This reference also provides information on explosive limits and presents a method for calculating the flammability characteristics of mixtures, based on the properties of pure compounds.

Consideration should be given to any phase change, either vaporization of liquid or condensation of vapour, that occurs in the fluid when the pressure is reduced or as a result of cooling. With auto-refrigeration, vaporization of volatile liquids can be incomplete unless facilities are provided to add the necessary heat for vaporization.

Caution should be exercised to avoid mixing chemicals that can react in flare headers. Routing reactive materials to a flare header has caused high flare pressures that have resulted in flare-header ruptures. Materials that react violently when mixed with water (such as alkyls, sodium, potassium and silanes) should be routed to a segregated header that does not contain water.

Caution should be exercised to avoid mixing of water with other sources if there is a potential for solids formation in the flare system. If there is a potential for formation of ice, the water sources should be routed in a separate header to the flare knockout drum; see also 6.6.2.4.

See also 7.2.3.

6.2.2 Physiological and nuisance properties

The physiological and nuisance properties of material released from pressure-relieving and depressuring systems should be studied to establish the proper type of disposal system.

6.2.3 Recovery value

The monetary value of process wastes can warrant special means of collection for return to the process, as is the case, for example, of costly solvents. An economic-engineering evaluation can determine whether the recovery value of the material justifies the installation of a recovery system. If a recovery system is justified, or required by local regulations, refer to 7.4 for guidance. To avoid loss of valuable process material, the pressure-relieving device should be set sufficiently above the normal operating pressure to give a reliable margin of differential pressure.

6.3 Atmospheric discharge

6.3.1 General

In many situations, pressure-relief vapour streams can be safely discharged directly to the atmosphere if environmental regulations permit such discharges. This has been demonstrated by many years of safe operation with atmospheric releases from properly installed vapour-pressure-relief devices. Technical work sponsored by API [70] has also shown that, within the normal operational range of conventional pressure-relief devices, well-defined flammable zones can generally be predicted for vapour releases. With proper recognition of the appropriate design parameters, vapour releases to the atmosphere can provide for the highest degree of safety. Where feasible, this arrangement offers significant advantages over alternative methods of disposal because of its inherent simplicity, dependability and economy. The decision to discharge hydrocarbons or other flammable or hazardous vapours to the atmosphere requires careful attention to ensure that disposal can be accomplished without creating a potential hazard or causing other problems, such as the formation of flammable mixtures at grade level or on elevated structures, exposure of personnel to toxic vapours or corrosive chemicals, ignition of relief streams at the point of emission, excessive noise levels and air pollution.

6.3.2 Formation of flammable mixtures

6.3.2.1 General

The intent of 6.3.2 is to address design issues for individual relief-device tailpipes that vent directly to atmosphere.

To evaluate the potential hazards of flammable mixtures that result from atmospheric discharge of hydrocarbons, the physical state of the released material is of primary importance; e.g. the behaviour of a vapour emission is entirely different from that of a liquid release. Between these two extremes are situations involving liquid-vapour mixtures in which mists or sprays are formed. Vapours, mists and liquids each introduce special considerations in analyzing the risk associated with atmospheric relief.

6.3.2.2 Vapour emission

When hydrocarbon-relief streams comprised entirely of vapours are discharged to the atmosphere, mixtures in the flammable range unavoidably occur downstream of the outlet as the vapour mixes with air. Under most circumstances in which individual pressure-relief valves discharge vertically upward through their own stacks, this flammable zone is confined to a rather limited definable pattern at elevations above the level of release. At exit velocities from the pressure-relief-valve stack, the jet-momentum forces of release usually are dominant [70]. Under these conditions, the air-entrainment rate is very high, and the released gases are then diluted to below the lower flammable limit before the release passes out of the jet-dominated portion if the Reynolds number, Re, meets the criterion of Equation (22):

$$Re > 1{,}54 \cdot 10^4 \left(\frac{\rho_\mathrm{j}}{\rho_\infty} \right) \tag{22}$$

where

Re is the Reynolds number at the vent outlet;

ρ_j is the density of the gas at the vent outlet;

ρ_∞ is the density of the air.

NOTE Equation (22) might not be valid where jet velocity is less than about 12 m/s (40 ft/s) or when the jet-to-wind velocity ratio is less than 10.

On the other hand, if the release is at too low a velocity and has too low a Reynolds number, jet entrainment of air is limited, and the released material is wind dominated. Principles of atmospheric dispersion then determine the dilution rate and the distance within which flammable conditions can occur. Under these conditions, flammable mixtures can possibly occur at grade or at distant ignition sources. A complete evaluation requires consideration of the following:

a) velocity and temperature of the exit gas;

b) relative molecular mass and quantity of the exit gas;

c) prevailing meteorological conditions, especially any adverse conditions peculiar to the site;

d) local topography and the presence of nearby structures;

e) elevation at which the emission enters the atmosphere.

Previous technical investigations [71] have demonstrated the rapid dispersion caused by the turbulent mixing that results from dissipation of energy in a high-velocity gas jet. For a situation in which a pressure-relief valve is flowing at or close to full capacity, discharge velocities through independent atmospheric stacks usually exceed 150 m/s (500 ft/s). The studies on the discharge of jets into still air indicate that gases with velocities of 150 m/s (500 ft/s) or more have sufficient energy in the jet to cause turbulent mixing with air and effect dilution in accordance with Equation (23):

$$\frac{q_{\mathrm{m,y}}}{q_{\mathrm{m,o}}} = 0{,}264 \frac{y}{d} \tag{23}$$

where

$q_{\mathrm{m,y}}$ is the mass flow rate of the vapour-air mixture at distance, y, from the end of the tail pipe;

$q_{\mathrm{m,o}}$ is the mass flow rate of the relief device discharge, expressed in the same units as $q_{\mathrm{m,y}}$;

y is the distance along the tail pipe axis at which $q_{\mathrm{m,y}}$ is calculated;

d is the tail pipe diameter, expressed in the same units as y.

Equation (23) indicates that the distance, y, from the exit point at which typical hydrocarbon relief streams are diluted to their lower flammable limit (i.e. a mass fraction of 3 %) occurs approximately 120 diameters from the end of the discharge pipe, measured along the axis. In essence, when hydrocarbon vapours are diluted with air to a mass fraction of approximately 3 %, the concentration of the resultant mixture is at or below the lower flammable limit. This value actually varies from 3,0 % for methane to 3,6 % for hexane. When figured on a volume fraction basis, which is more commonly used than mass fraction to express limits of flammability, these values are equivalent to 5,3 % and 1,2 %, respectively. For materials that do not have combustion characteristics similar to light hydrocarbons, the extent of a flammable mixture can differ considerably from 120 diameters. Based on these dispersion data, it can be concluded that where discharge velocities are achieved, the hazard of flammable concentrations below the level of the discharge point is negligible. Fixed distances may be used for designs based upon experience, which precludes the need to perform dispersion analyses. This confirms the many years of experience with vapour releases from pressure-relief valves discharging directly to the atmosphere without accumulating flammable concentrations.

Through the years, skepticism arose about the validity of this past investigation, even though experience had indicated that large flammable volumes were not created by pressure-relief valves releasing vapour directly to the air. There was concern because any system designed for a discharge velocity of 150 m/s (500 ft/s) at maximum conditions can have a lower discharge velocity under other conditions.

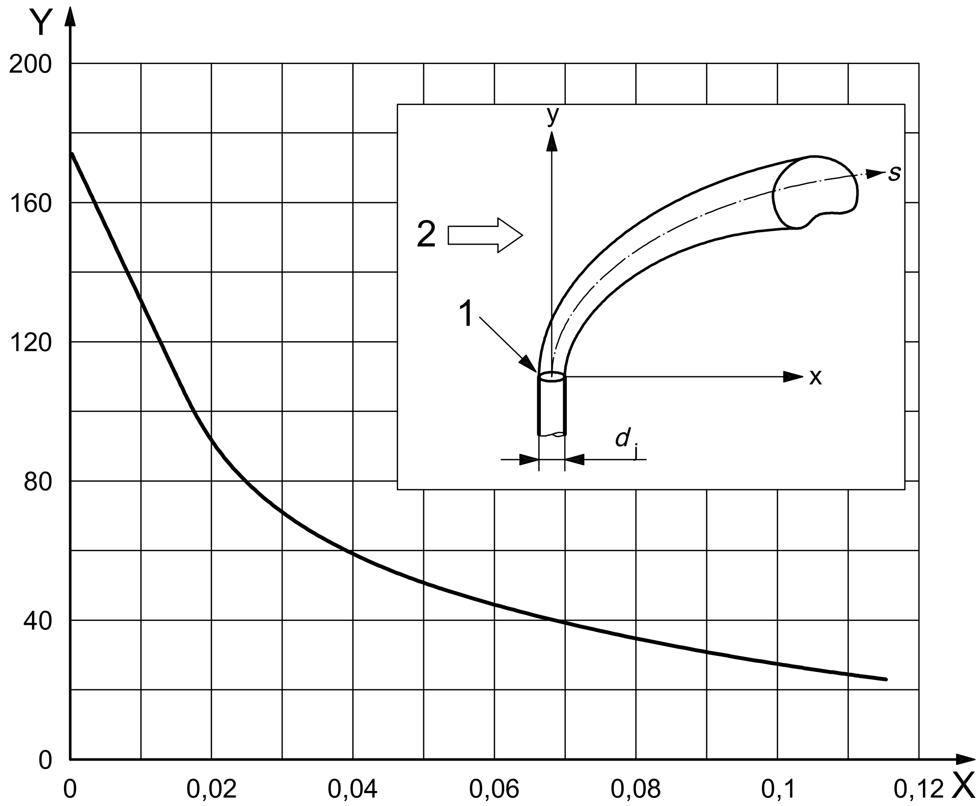
Although a high discharge velocity is characteristic of a pressure-relief valve when it is flowing at design capacity, one cannot assume that a pressure-relief valve is flowing at full capacity. For example, even though the initial release can be at a high velocity, once a spring-loaded pressure-relief valve has opened, kinetic forces are sufficient to offset the spring-closing force until the flow has been reduced to approximately 25 % of the valve's rated capacity. Reduced flow rates can occur as the conditions affecting relief are corrected. In many cases, overpressure can result from a minor operating upset, causing the flow rate to be appreciably less than the design capacity. The probability of these situations occurring can often be minimized by using two or more pressure-relief valves and staggering the set pressure to provide for sequential operation. Using a common vent stack for several pressure-relief valves can also result in a discharge at a relatively low velocity if only one valve is operating.

Because of these concerns, studies were undertaken at the Battelle Institute [70] to evaluate the effects of reduced velocities of discharge at the point where the pressure-relief valve is about to reseat at approximately one-quarter of the valve's rated capacity. Also covered in these studies are the effects of the temperature and relative molecular mass of the hydrocarbon gases as they affected the zone of flammability under various ratios of exit velocity to wind velocity. These studies verified that vapours released from pressure-relief valves through their individual stacks are safely dispersed even when the valves were operating at only 25 % of their full capacity, which corresponds to the reseat level of the valves. As long as the minimal value given by Equation (22) is exceeded, the release is jet-dominated and diluted outside the flammable range, within the jet pattern. For the most part, vent velocities are greater than 30 m/s (100 ft/s), even at the 25 % release rate.

Other studies of the safety of tanker venting [72] have shown the same jet-momentum dilution effects where release velocities exceed 30 m/s (100 ft/s). One would expect that only for low-set pressure-relief valves, or for multiple valves routed via a manifold into a common vent stack, is the Reynolds number of the released gases below the minimum necessary for jet momentum effects.

Figures 4, 5 and 6 demonstrate the limits of flammability vertically, horizontally and along the main axis of the jet. Hoehne and Luce [70] indicate these figures apply to both single and multi-component hydrocarbon jets of any molecular mixture between methane and heptane. For more detailed analysis, dispersion modelling can be performed. The axial and vertical distances in still air are indicated to be somewhat greater than the 120 diameters indicated by previous still-air studies. However, the horizontal limit of the flammable envelope is shown to be essentially independent of the wind velocity and is significantly lower than the axial distances indicated by the previous study.

The studies demonstrate the adequacy of the general industry practice of locating pressure-relief-valve stacks that discharge to the atmosphere at least 15 m (50 ft) horizontally from any structures or equipment running to a higher elevation than the discharge point. In most cases, this is adequate to prevent flammable vapours from reaching the higher structures. With these jet-momentum releases, there should also be no concern about large clouds of flammable vapours or flammable conditions existing at levels below the release level of the stack. These recent studies have generally verified long-standing experience relative to the safety of vapour releases vertically to the atmosphere from atmospheric pressure-relief valve discharge stacks.



Key

X velocity ratio, u_∞ / u_j

Y vertical plume centre distance factor, $y / (d_j) \sqrt{\rho_j / \rho_\infty}$, dimensionless

u_∞ is the wind speed, expressed in metres per second (feet per second)

u_j is the jet exit velocity, expressed in metres per second (feet per second)

ρ_j is the fluid density inside the tip exit, expressed in kilograms per cubic metre (pounds per cubic foot)

ρ_∞ is the density of the ambient air, expressed in kilograms per cubic metre (pounds per cubic foot)

d_j is the inside diameter of the tip (jet exit diameter), expressed in metres (feet)

y is the vertical distance, expressed in metres (feet)

The insert defines the flow system and axes:

x is the horizontal distance, expressed in metres (feet)

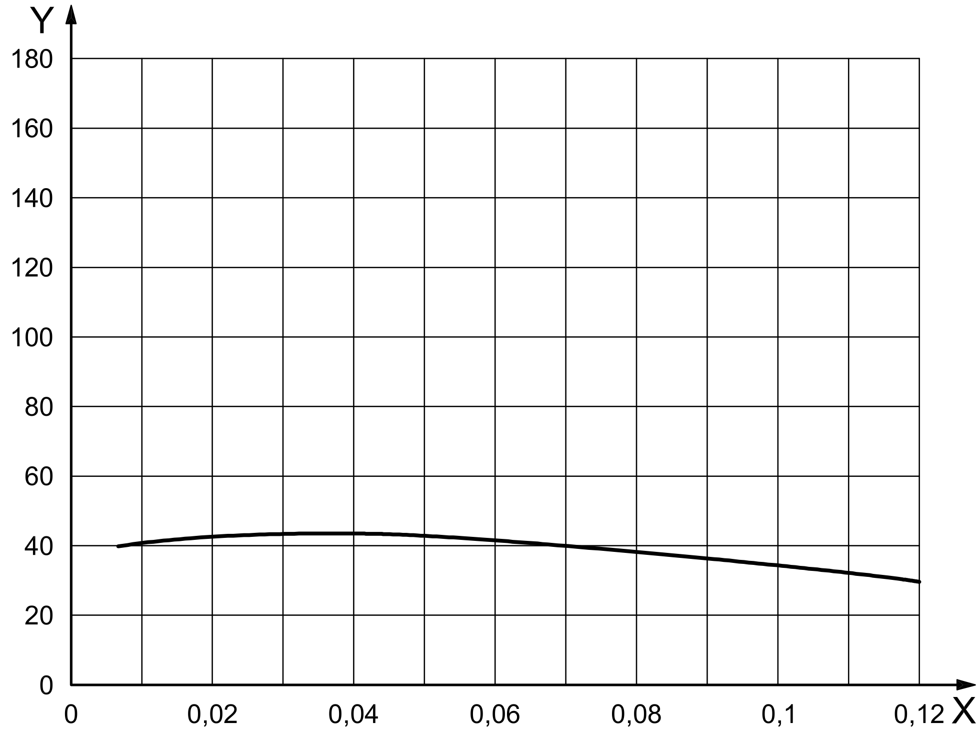
S is the distance along the jet centre from the tip apex, expressed in metres (feet)

1 tip of discharge stack

2 wind (cross-stream)

The maximum downwind vertical distance from jet exit to lean-flammability concentration limit is the distance factor multiplied by $\left[(d_j)\sqrt{\rho_j / \rho_\infty}\right]$, expressed in metres (feet).

Figure 4 — Maximum downwind vertical distance from jet exit to lean-flammability concentration limit for petroleum gases



Key

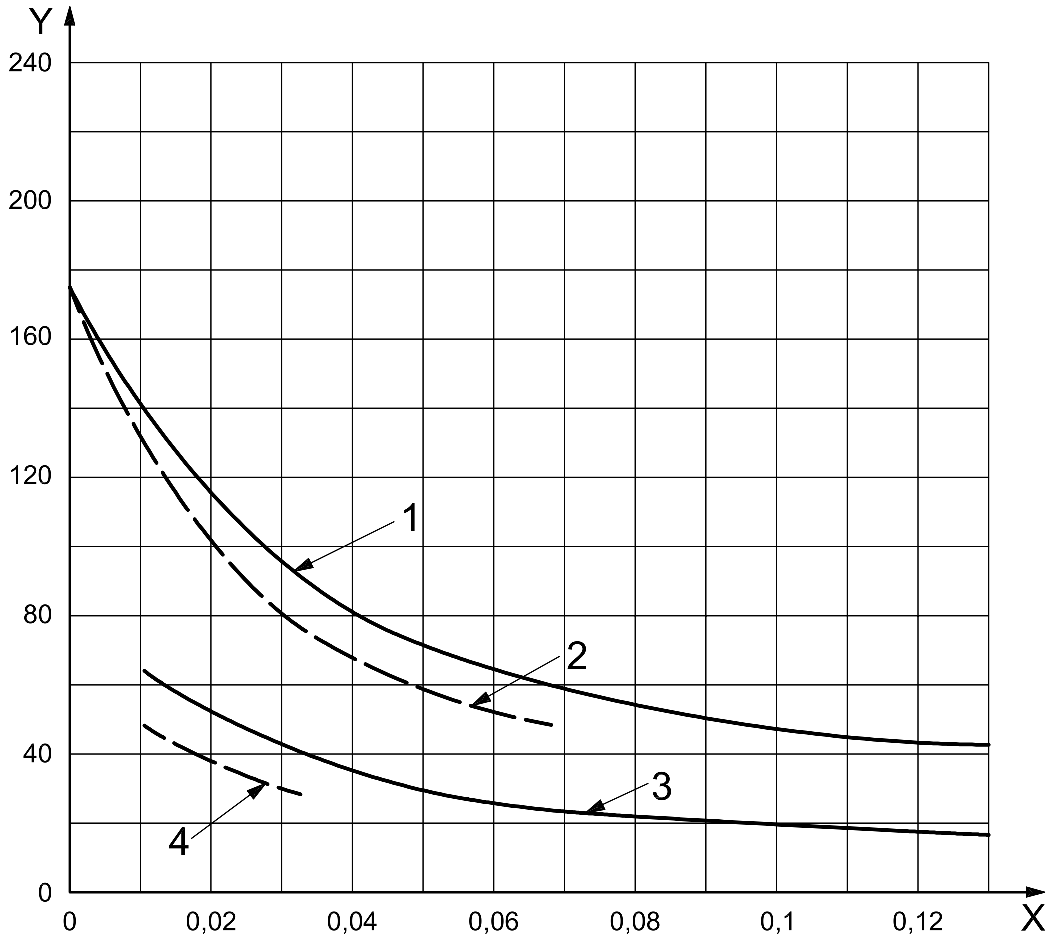
X velocity ratio, u_∞ / u_j

Y downwind horizontal plume centre distance factor, $x \big/ \left(d_\mathrm{j} \right) \sqrt{\rho_\mathrm{j} / \rho_\infty}$, dimensionless

NOTE See Figure 4 for the definition of the other variables and for the flow system and axes.

The maximum downwind horizontal distance from jet exit to lean-flammability concentration limit is the distance factor multiplied by $\left[\left(d_\mathrm{j} \right) \sqrt{\rho_\mathrm{j} / \rho_\infty} \right]$, expressed in metres (feet).

Figure 5 — Maximum downwind horizontal distance from jet exit to lean-flammability concentration limit for petroleum gases



Key

X velocity ratio, u_∞ / u_j

Y downwind horizontal plume centre distance factor, $S / \left(d_\mathrm{j}\right) \sqrt{\rho_\mathrm{j} / \rho_\infty}$, dimensionless

1 lean limit, cold jet

2 lean limit, hot jet

3 rich limit, cold jet

4 rich limit, hot jet

NOTE See Figure 4 for the definition of the other variables and for the flow system and axes.

The axial distance to lean- and rich-flammability concentration limits is the distance factor multiplied by $\left[\left(d_\mathrm{j}\right) \sqrt{\rho_\mathrm{j} / \rho_\infty}\right]$, expressed in metres (feet).

Figure 6 — Axial distance to lean- and rich-flammability concentration limits for petroleum gases

6.3.2.3 Mist emission

Mists, as referred to in this International Standard, result from condensation following vapour relief. Fine sprays associated with relief streams that contain liquids are considered in 6.3.2.4. Condensed mists are finely divided; the diameter of most drops is less than 10 µm, with few larger than 20 micrometers. Mechanical sprays do not usually contain many drops smaller than 100 µm in diameter.

Whether vapours condense in appreciable quantities when they are released to the atmosphere depends on the stream composition, atmospheric temperature and exit velocity. The assumption is frequently made that if the lowest anticipated atmospheric temperature is below the dew point of a released hydrocarbon, significant condensation will occur. This approach ignores two important effects associated with the release of vapours. As vapours are depressurized across the pressure-relief valve, they are superheated, and the tendency for immediate condensation to occur is minimized in the highly enriched zone at the point of emission. More important is the combined dilution effect of air and light components normally present in discharges from pressure-relief valves. Rapid dilution tends to lower the dew point of individual components to a point below the ambient temperature.

Loudon [73] gives a method of calculating whether condensation of a discharge from a pressure-relief valve can occur. These calculations indicate that most emissions do not condense, regardless of relative molecular mass, although relatively heavy molecular mass hydrocarbons can condense in the range previously noted (10 µm to 20 µm). This approach is supported by experience with refinery relief installations involving discharge to the atmosphere of vapour streams covering a wide range of conditions.

In cases in which the vapour discharges from pressure-relief valves condense, consideration should be given to how this condensation influences the formation of a flammable atmosphere. Combustible liquid mists in air are capable of propagating flame when they are ignited, even though the liquid is so non-volatile that no appreciable amount of vapour is formed at the ambient temperature. Mists of flammable liquids can thus present a hazard even at temperatures well below their flash point. Burgoyne [74] has shown that for flammable, condensed mists, the mass-percent lower flammability limit and the burning velocity are the same as for the corresponding vapour. According to Saletan [75], the ignition energy required to ignite a mist in air at ambient temperatures and pressures is approximately 10 times that needed to ignite a vapour.

In cases in which calculations indicate that vapour discharges from pressure-relief valves can condense, coalescence can possibly produce droplets that rapidly settle to grade rather than disperse as a mist similar to vapours. The hydrocarbon partial pressure at which the calculated adiabatic air-mixing curve intersects the dew-point curve should be considered indicative of bounding the region in which coalescence seems unlikely. Although no conclusive data are now available, condensation at hydrocarbon partial pressure of 34 kPa (5 psi) or less should be treated as finely divided mists without coalescence. In the absence of coalescence, the effect of gravity should be negligible, since the free-fall velocity of 10 µm hydrocarbon particles in air is approximately 3 mm/s (0,01 ft/s). Therefore, even with very light wind, the discharge from an elevated location travels a considerable distance before it reaches grade.

Based on the foregoing factors pertaining to the dispersion and combustion characteristics of a mist, it can be concluded that as long as the condensate remains in a finely divided form and is airborne, the mixture can be treated for flammability and dispersion characteristics as though it were completely vaporized. Because of the extremely small size of the droplets, use of the methods described in 6.3.2.2 can give an order of magnitude of the concentration at various distances from the point of emission. As noted above, the same mass percentages of hydrocarbons are necessary to make a mist flammable as are necessary to make a vapour flammable. It can, therefore, also be concluded that as long as the minimum Reynolds number of the release is in excess of that required by Equation (22), the envelope of flammability for the mist is within the same confined predictable limits as those for a vapour. Therefore, although condensation can create problems relative to air pollution, if only hydrocarbon-flammability considerations are involved, the risk from an area-explosion potential is no higher than if the vapours had not condensed.

6.3.2.4 Liquid emission

Unlike discharges composed of vapour or mist, which rapidly disperse when they are vented to the atmosphere at high velocity, liquid discharges settle to grade. If volatile components are present, a flammable atmosphere can result. The risk of fire or explosion can be high if appreciable quantities of liquid hydrocarbons are released to the atmosphere when the ambient temperature is at or above the flash point of the liquid.

Theoretically, liquids that have a flash point above the maximum anticipated ambient temperature do not vaporize enough to create a flammable atmosphere. However, widespread spraying of oil droplets can create concern in an emergency and constitute a serious nuisance. Also, fires can occur if the liquid comes in contact with very hot lines or equipment. Therefore, all liquid-relief streams should generally be disposed of by one of the methods described in 6.6.

To minimize the possibility of a release of flammable liquid, all pressure-relief valves that vent vapour to the atmosphere should be located so that the valve inlet connects to the vapour space of vessels or lines. In some instances, additional safeguards are warranted. For example, during unit upsets, liquid levels can increase and flood the vessels that are partially or totally filled with vapour during normal operations. The potential for such a situation can be greatly minimized by locating pressure-relief valves at a point in the process system where the probability of liquid occurring at the pressure-relief valve inlet is considered negligible because of factors related to time and the system's liquid capacity. For example, a pressure-relief valve located on top of a large fractionating column presents far less risk of liquid release than a valve positioned on the overhead receiver, which can flood in a matter of minutes. In other situations, high-level alarms or other instrumentation can provide a valuable safeguard against high-liquid levels reaching the pressure-relief-valve inlet.

In summary, a rigorous analysis should be made of the various causes of overpressure on any system containing flammable liquid in which pressure-relief valves that vent to the atmosphere are included in the design. All possibilities that can allow liquid to gain entrance to the pressure-relief valve should be determined and appropriate safeguards should be taken to prevent this occurrence.

6.3.3 Exposure to toxic vapours or corrosive chemicals

6.3.3.1 Toxic vapours

Although most vapour streams can be harmful to breathe at high concentrations, the majority present little or no risk to personnel when they are discharged from pressure-relief valves at a remote location.

Certain process streams contain vapours that are dangerous at extremely low concentrations; e.g. hydrogen sulfide vapours can cause unconsciousness within seconds following exposure to a concentration above 1 000 mg/kg [ppm[3)]]. This is approximately one-tenth the concentration representing the lowest flammable limits of any hydrocarbon. Where toxic materials are present in a relief stream, an investigation should be made to predict the maximum downwind concentration at any location where personnel can be exposed. Special attention should be given to adjacent elevated structures that can lie within the path of the plume and are thus subject to relatively high concentrations.

Each situation in which toxic vapours can be released to the atmosphere warrants careful analysis. Since toxicity varies greatly for different materials, the maximum concentration that can be tolerated should first be determined. Based on the length of exposure, the maximum tolerated level may vary for different locations. A higher concentration can pose less risk at locations that can be quickly and safely evacuated as opposed to those locations where it is necessary for personnel to remain on duty or where personnel cannot readily leave. Of further importance is the probable duration of a release. Most emergencies that cause overpressure on equipment can be controlled within 5 min to 10 min. The duration of an emergency varies, depending on the process and equipment involved. For example, when the source of overpressure can be eliminated by shutting down a pump or compressor, the duration of relief should be shorter than if a fractionating column were to overpressure. A period of 10 min to 30 min should be sufficient to control any emergency situation short of a catastrophe.

The actual exposure of an individual to a release is difficult to determine accurately but should be estimated. Where toxic releases have a sufficiently high Reynolds number, they meet the dispersion/dilution criteria for jet momentum releases described in 6.3.2.2. The materials in the release can be expected to be diluted between 30:1 and 50:1 before the jet-momentum effects are lost. In considering further dispersion from the end of the jet, one should take into account that the released materials have already been diluted to at least this level. From this level of mixing, calculations of ground concentrations can be evaluated based on the techniques given in Gifford's article [76].

3) "ppm" is a deprecated unit.

6.3.3.2 Corrosive chemicals

Certain chemicals, such as phenols, that are liquid at ambient conditions can create a serious hazard to personnel if they are discharged from pressure-relief valves to the atmosphere. When process systems contain such chemicals, atmospheric relief is not safe unless valves can be installed at locations where personnel exposure from release of such materials can be avoided. Many of the same considerations discussed in 6.3.2.4 concerning avoidance of liquid releases apply to corrosive chemicals.

6.3.4 Ignition of a relief stream at the point of emission

6.3.4.1 Sources of ignition

6.3.4.1.1 The possibility of accidental ignition of the outflow of flammable vapours from a pressure-relief valve or a vent can best be analysed in terms of the possible causes of ignition covered in 6.3.4.1.2 through 6.3.4.1.5.

6.3.4.1.2 The possible existence of outside ignition sources such as open flames, hot surfaces and unclassified electrical equipment installed in surrounding areas and on structures is known or can be anticipated. With jet-momentum releases from pressure-relief valves, emission points can be located by dispersion modelling so that the flammable pattern evolved does not reach such sources. This becomes more difficult where wind-dominated, low-velocity releases are involved, since flammable patterns can extend considerable distances from the release point. Also, in these instances, the ignition potential from temporary sources, such as automotive equipment or hot-work activities, should be recognized. With normal atmospheric releases, outside ignition sources can be readily avoided by the proper location of vents. On the other hand, with low-velocity, low-momentum releases, a careful design check should be made of conditions at various emissions rates and atmospheric conditions to avoid the potential of ignition by outside sources. In lieu of dispersion calculations, fixed distances may be used for designs based upon experience.

6.3.4.1.3 Discharges from open atmospheric vents have been known to be ignited by lightning. Except for emergency discharges associated with power outages that can occur during thunderstorms, the probability of lightning occurring simultaneously with the opening of a relief valve is negligible. Intermittent discharges over long periods and continuous discharges (e.g. from leaking relief valves) increase the probability of lightning ignition. For additional information about lightning, see the recommendations in NFPA 780 [36].

6.3.4.1.4 For general information on electrostatics, see Eichel's article [77] and API RP 2003 [15]. During high-velocity discharges from gas wells to the atmosphere, static discharges are developed that are sufficient to cause sparks and ignition [78]. The condensate zone in the jet of well-head gas apparently tends to produce a high level of charge, although ignition does not actually occur. Another theory relating to static ignition proposes that gas flow through a piping system during venting induces a static charge on any solid or liquid particles in the pipe stream that contact the pipe wall. As the gas reaches the sharp edges of the vent outlet, static discharges can occur, either by complete electrical breakdown (spark discharge) or by partial electrical breakdown (corona discharge). There is a lack of documented information on the ignition of relief-valve vapour discharges attributed to the development of electric potential at the discharge point. The experience of pipeline companies (who customarily discharge natural gas to the atmosphere at low elevations) includes gas-gauge pressures as high as 6 200 kPa (900 psi) and discharge rates as high as 82 kg/s (650 000 lb/h) from a single vent stack. The probability of ignition by static electricity is, therefore, very low because of a relatively weak charge build-up in the jet and reasonable isolation from the well-grounded vent stack.

This conclusion pertains to hydrocarbon vapour releases. Experience indicates that streams with a high hydrogen content are susceptible to ignition by static electricity as a result of the described mechanism because of electrostatic discharges at the sharp edge of the vent outlet. NASA investigated this phenomenon[79] and found that such electrostatic discharges can be prevented by installing a toroidal ring on the vent-stack outlet. This ring inhibits the static discharge at the vent stack exit by removing the sharp edged geometry of the vent outlet, which is conducive to spark formation.

Ignition of hydrogen from atmospheric vents can also result from the chemical reaction between hydrogen and iron oxides frequently found in vessels and piping. When a stream containing extremely small particles of ferrous oxide (FeO) or iron (Fe) is brought into close contact with the oxygen present in the atmosphere, an exothermic reaction occurs that under ideal conditions can provide sufficient energy to ignite a hydrogen-air mixture. The energy requirement has been experimentally determined at 0,017 millijoule (approximately 5 % of that necessary to ignite a methane-air mixture). This quantity of energy can conceivably be imparted to a tiny particle as a result of the heat released in the reaction of either FeO or Fe with oxygen (O_2). Furthermore, if the ratio of surface area to mass were high enough, a temperature sufficient for ignition can be reached. Also, because of the wide explosive range of hydrogen (volume fraction from 4 % to 75 %), flammable atmospheres are formed very close to the point of release. This, along with hydrogen's very low ignition energy, increases the probability of ignition.

6.3.4.1.5 Relief streams that are above the auto-ignition temperature on the upstream side of the valve can ignite spontaneously on contact with air unless sufficient cooling occurs before a flammable vapour-air mixture is formed. For this reason, these hot streams should usually be routed to a closed system, cooler or quench tower. Under some circumstances, with proper location of the discharge stack, ignition can be tolerated. Under these conditions, the thermal radiation effects discussed in 6.3.4.3 should be evaluated. See API Publ 2216 [17] for more information.

6.3.4.2 Explosive release of energy

If a quantity of gas accumulates and then ignites, the possible explosive release of energy in the atmosphere can cause concern about using atmospheric relief. Where unconfined jet-momentum releases are involved, as with a normal pressure-relief device, there is likely to be little potential for the accumulation of large vapour clouds and this can be validated by dispersion and consequence modelling. The total potential hazard can be related to the total quantity of hydrocarbon-air mixture that accumulates within the flammable envelope downstream of the point of emission. With jet-momentum releases, the total volume can be calculated. In a typical case, the flammable zone can be in the range of 40 diameters to 120 diameters downstream but can vary depending on densities and ratios of jet-to-wind velocity. The mixture in this zone can contain an average of about 6 % hydrocarbon, which would represent 3 s of the emitted outflow. The volume within the flammable range at any time is relatively small compared with the total gas volume emitted and considerably limits the problem even if an ignition does occur.

If the release rate does not achieve jet momentum and dilution is not achieved, vapour clouds can develop. Similarly, if even a relatively small amount of flammable gas accumulates in a confined or congested space, a significant hazard can be created. In these cases, care should be taken to avoid any confinement of the released gases, since the degree of confinement determines the pressure rise if accidental ignition occurs. Evaluating such confinement should take account of the proximity of buildings or high concentrations of equipment that produce congestion or confinement. The total potential hazard from such sources can then be related to the total quantity of gas released [80].

6.3.4.3 Thermal radiation effects

Wherever large quantities of flammables are vented, the potential heat release is sufficient to warrant considering its effects on personnel and equipment, even though ignition of the discharge from pressure-relief devices is highly improbable. Once allowable thermal radiation levels are established, the required distance from various exposure locations to the point of emission can be calculated (see 6.4.2.3 for information on evaluating thermal radiation effects).

6.3.5 Excessive noise levels

The noise generated by a pressure-relief valve discharging to the atmosphere can be loud. The noise levels produced by gases at the point of atmospheric discharge can be approximated by reference to 7.3.4.3. Since emergency relief is typically infrequent and of short duration, the noise might not be subject to regulation. In many areas, regulatory authorities define allowable levels of noise exposure for personnel or at property limits. If no regulatory limits are prescribed, the proposed standards of the American Conference of Governmental Industrial Hygienists [81] may be applied.

The allowable noise intensity and duration should be evaluated at areas where operating personnel normally work or at property limits. If two or more pressure-relief valves can discharge to the atmosphere simultaneously, it is necessary to evaluate the combined effects. For design information on noise levels associated with atmospheric discharge, see 7.3.4.3.

6.3.6 Air pollution

The continuing problem of air pollution has become a factor that warrants serious consideration. Regulations pertaining to air pollution usually provide exemption for discharges that occur only under emergency conditions; however, effluent concentrations at grade level or other locations obviously should be controlled, even though the acceptance level for limited and occasional emergency discharge can be much higher than that for prolonged or continuous emissions. Methods for calculating the grade-level concentration to determine whether air pollution exists are discussed in Gifford's article [76].

6.4 Disposal by flaring

6.4.1 General

The primary function of a flare is to use combustion to convert flammable, toxic or corrosive vapours to less objectionable compounds. Selection of the type of flare and the special design features required are influenced by several factors, including the availability of space; the characteristics of the flare gas, namely, composition, quantity and pressure level; economics, including both the initial investment and operating costs; and public relations. Public relations can be a factor if the flare can be seen or heard from residential areas or navigable waterways.

Flare mechanical design, operation and maintenance issues are covered by API Std 537.

6.4.2 Combustion properties

6.4.2.1 Flame properties

6.4.2.1.1 A flame is a rapid, self-sustaining chemical reaction that occurs in a distinct reaction zone. The two basic types of flames are (a) the diffusion flame, which is found in conventional flares and occurs on ignition of a fuel jet issuing into air, and (b) the aerated flame, which occurs when fuel and air are premixed before ignition. The burning velocity, or flame velocity, is the speed at which a flame front travels through the unburned combustible mixture.

6.4.2.1.2 In the case of a flare, the flame front is normally at the top of the stack; however, at low gas velocities, back mixing of air occurs in the top of the stack. Experiments [82] have shown that if a sufficient flow of combustible gas is maintained to produce a flame visible from ground level, there is usually no significant back mixing of air into the stack. At lower gas flows, there is the possibility of combustion at a flame front located part of the way down the flare tip with a resultant high tip temperature. Or there can be flame extinguishment with subsequent formation of an explosive mixture in the stack and ignition from the pilot light.

In an aerated flame from a premixing device, such as a flare pilot, a phenomenon known as flashback can occur. This results from the linear velocity of the combustible mixture becoming less than the flame velocity, causing the flame to travel back to the point of mixture.

In the case of diffusion flames, if the fuel flow rate is increased until it exceeds the flame velocity at every point, the resultant turbulent mixing and dilution with air can cause the flame to be lifted above the burner until a new stable position in the gas stream above the burner is reached. This phenomenon is called a detached, stable flame. (Extinguishment of the flame is referred to as blow off.) Both blow off and flashback velocities are greater for fuels that have high burning velocities. Small amounts of hydrogen in a hydrocarbon fuel widen the stability range because blow-off velocity increases much faster than flashback velocity.

The allowable flare-burner exit velocity is a function of relief-gas composition, flare burner design and the gas pressure available. These parameters are inter-related. Some flare tips incorporate a flame-retention device or other means that provides a stable burning flame either attached or detached relative to the flare tip. There is evidence [83], [84], [85], [86], [87], [88], [89] that flame stability can be maintained at relatively high velocities depending on the discharge properties and the type of tip used. Experience has shown that a properly designed and applied flare burner can have an exit velocity of more than Mach 0,5, if pressure drop, noise and other factors permit. Many pipe flares, assisted or unassisted, and air-assisted flares have been in service for many years with Mach numbers ranging from Mach 0,8 and higher.

Some flares are subject to regulations that limit exit velocity. Pipe flares applied in the U.S. as control technology for volatile organic compound (VOC) emissions can have gas exit velocity limited by the United States Code of Federal Regulations 40CFR60.18 [118]. In some locales, the 40CFR60.18 requirements on exit velocities have been extended to “emergency conditions”. The regulations provide guidelines for the determination of the maximum exit velocity as a function of waste-gas characteristics and the type of flare burner employed. It is important to note that there are many flare applications that do not involve VOC control. Such flares are not usually required to meet the exit velocity requirements of the CFR.

6.4.2.2 Smoke

Many hydrocarbon flames are luminous because of incandescent carbon particles formed in the flames. Under certain conditions, these particles are released from luminous flames as smoke. The exact reasons and mechanisms by which smoke is formed are still not fully understood. Many different processes have been suggested, but a discussion of them is beyond the scope of this International Standard. However, it is safe to say that smoke is formed during the combustion of hydrocarbons only when the system is fuel-rich, either overall or locally. Observation has revealed that suppression of the hydrogen-atom concentration in the flames accompanies the suppression of smoke formation [90]. Smoke formation can possibly be reduced by reactions that consume hydrogen atoms or render them ineffective.

The ways in which water vapour reduces smoke from flares have been discussed by Smith [91]. Briefly, one theory suggests that steam separates the hydrocarbon molecules, thereby minimizing polymerization, and forms oxygen compounds that burn at a reduced rate and temperature and that are not conducive to cracking and polymerization. Another theory claims that water vapour reacts with the carbon particles to form carbon monoxide, carbon dioxide and hydrogen, thereby removing the carbon before it cools and forms smoke.

6.4.2.3 Thermal radiation

6.4.2.3.1 Many investigations have been undertaken to determine the effect of thermal radiation on human skin. Using human subjects, Stoll and Greene [92] found that with an intensity of 6,3 kW/m^2 (2 000 Btu/h·ft^2), the pain threshold is reached in 8 s and blistering occurs in 20 s. On the bare skin of white rats, an intensity of 6,3 kW/m^2 (2 000 Btu/h·ft^2) produces burns in less than 20 s. The same report indicates that an intensity of 23,7 kW/m^2 (7 500 Btu/h·ft^2) causes burns on the bare skin of white rats in approximately 6 s. Table 8 gives Buettner’s [93] exposure times necessary to reach the pain threshold as a function of radiation intensity. These experimental data were derived from tests given to people who were radiated on the forearm at room temperature. The data indicate that burns follow the pain threshold fairly quickly. Buettner’s data agree well with those of Stoll and Greene. Tissue damage starts with a burn that resembles a mild sunburn. As exposure time and/or radiation intensity increases, the burn progresses to a severe sunburn and with further exposure into more serious tissue damage.

Table 8 — Exposure times necessary to reach the pain threshold

Radiation intensity kW/m^2 (Btu/h·ft^2)	**Time-to-pain threshold** s
1,74 (550)	60
2,33 (740)	40
2,90 (920)	30
4,73 (1 500)	16
6,94 (2 200)	9
9,46 (3 000)	6
11,67 (3 700)	4
19,87 (6 300)	2

Since the allowable radiation level is a function of the length of exposure, factors involving reaction time and human mobility should be considered. In emergency releases, a reaction time of 3 s to 5 s may be assumed. Perhaps 5 s more can elapse before the average individual seeks cover or departs from the area, which would result in a total exposure period ranging from 8 s to 10 s. In evaluating the emergency procedures, consideration may also be given to an exposed individual becoming incapacitated during an attempt to exit the area.

As a basis of comparison, the intensity of solar radiation is in the range of 0,79 kW/m^2 to 1,04 kW/m^2 (250 Btu/h·ft^2 to 330 Btu/h·ft^2) depending on geographical location and time of year. Solar radiation can be a factor for some locations, but its effect added to flare radiation has only a minor impact on the acceptable exposure time.

The flare owner/operator shall determine the need for a solar-radiation-contribution adjustment to the values given in Table 9 on a case-by-case basis. While an adjustment of 0,79 kW/m^2 to 1,04 kW/m^2 (250 Btu/h·ft^2 to 330 Btu/h·ft^2) to a 6,31 kW/m^2 (2 000 Btu/h·ft^2) level has a relatively small impact on flare cost, the same adjustment to a 1,58 kW/m^2 (500 Btu/h·ft^2) level results in a significant increase in cost. This determination can include, among other things, an analysis of the frequency of maximum radiation flaring, the probability of personnel or the public being near the flare during a maximum flaring incident, the probability of the sun and flame being aligned in such a manner as to have additive intensities and the ability of the personnel or the public to avoid or move away from the exposure.

Table 9 — Recommended design thermal radiation for personnel

Permissible design level K kW/m^2 (Btu/h·ft^2)	**Conditions**
9,46 (3 000)	Maximum radiant heat intensity at any location where urgent emergency action by personnel is required. When personnel enter or work in an area with the potential for radiant heat intensity greater than 6,31 kW/m^2 (2 000 Btu/h·ft^2), then radiation shielding and/or special protective apparel (e.g. a fire approach suit) should be considered. **SAFETY PRECAUTION — It is important to recognize that personnel with appropriate clothing [a] cannot tolerate thermal radiation at 6,31 kW/m^2 (2 000 Btu/h·ft^2) for more than a few seconds.**
6,31 (2 000)	Maximum radiant heat intensity in areas where emergency actions lasting up to 30 s can be required by personnel without shielding but with appropriate clothing [a]
4,73 (1 500)	Maximum radiant heat intensity in areas where emergency actions lasting 2 min to 3 min can be required by personnel without shielding but with appropriate clothing [a]
1,58 (500)	Maximum radiant heat intensity at any location where personnel with appropriate clothing [a] can be continuously exposed

[a] Appropriate clothing consists of hard hat, long-sleeved shirts with cuffs buttoned, work gloves, long-legged pants and work shoes. Appropriate clothing minimizes direct skin exposure to thermal radiation.

Flare system design and plant equipment layout should minimize the need for operator attendance and equipment installed in locations of high radiant heat intensity.

The design of towers or other elevated structures exposed to flare radiation should consider radiation effects on the ability to safely egress. If personnel exposure to radiant heat exceeds the guidelines provided above, then shielding or other protection should be considered. It is often most effective to accomplish this by locating ladders and platforms on a side away from the flare.

Personnel are commonly protected from high thermal radiation intensity by restricting access to any area where the thermal radiation can exceed 6,31 kW/m^2 (2 000 Btu/h·ft^2). The boundary of a restricted access area can be marked with signage warning of the potential thermal radiation exposure hazard. Personnel admittance to, and work within, the restricted access area should be controlled administratively. It is essential that personnel within the restricted area have immediate access to thermal radiation shielding or protective apparel suitable for escape to a safe location.

Another factor to be considered regarding thermal radiation levels is that clothing provides shielding, allowing only a small part of the body to be exposed to full intensity. In the case of radiation emanating from an elevated point, standard personnel protective measures, such as wearing of a hard hat, can reduce thermal exposure.

There are practical differences between laboratory tests and full-scale field exposure [70], [94]. Heat radiation is frequently the controlling factor in the spacing of equipment such as elevated and ground flares. Table 9 presents recommended design total radiation levels for personnel at grade or on adjacent platforms. The extent and use of personal protective equipment can be considered as a practical way of extending the times of exposure beyond those listed.

The effects of thermal radiation on the general public, who can be exposed at or beyond the plant boundaries, should be considered.

Each company may select the radiation level to which personnel can be exposed, either for a short duration or continuously. Table 9 is provided to guide companies in making this decision. However, many factors can influence the radiation levels to which personnel may be continuously exposed. The following are some of these factors:

a) environmental: Wind and ambient temperature can influence the amount of radiation a person can withstand.

b) design: Factors such as orientation of the work place with respect to the flare and shielding can both impact on personnel radiation exposure.

c) training: Properly trained workers wear appropriate clothing and know how to react to changing situations. For example, it can be safe to work with the wind blowing in a certain direction but unsafe if there is a drastic wind shift.

It is expected that each company evaluate the impact of these factors to determine a safe level of radiation exposure for their personnel.

6.4.2.3.2 In most cases, equipment can safely tolerate higher degrees of heat density than those defined for personnel. However, if any items vulnerable to overheating problems are involved, such as construction materials that have low melting points (e.g. aluminium or plastic), heat-sensitive streams, flammable vapour spaces and electronic or electrical equipment, then the effect of radiant heat on them might need to be evaluated. If an evaluation is necessary, a heat balance can be performed to determine the resulting surface temperature for comparison with acceptable temperatures for the equipment [94].

6.4.2.3.3 A common approach to determining the flame radiation to a point of interest is to consider the flame to have a single radiant epicentre and to use the following empirical equation by Hajek and Ludwig [85]. Equation (24) may be used for both subsonic and sonic flares, provided the correct F factor is used.

$$D = \sqrt{\frac{\tau \cdot F \cdot Q}{4\pi \cdot K}} \tag{24}$$

where

D is the minimum distance from the epicentre of the flame to the object being considered, expressed in metres (feet);

τ is the fraction of the radiated heat transmitted through the atmosphere;

NOTE Refer to C.3.6.3 for further information on the use of τ.

F is the fraction of heat radiated;

Q is the heat release (lower heating value), expressed in kW (Btu/h);

K is the radiant heat intensity, expressed in kW/m^2 (Btu/h·ft^2).

A discussion of the single-epicentre equation, Equation (24), and its terms together with a review and comparison of a number of interpretations of the method can be found in reference [88].

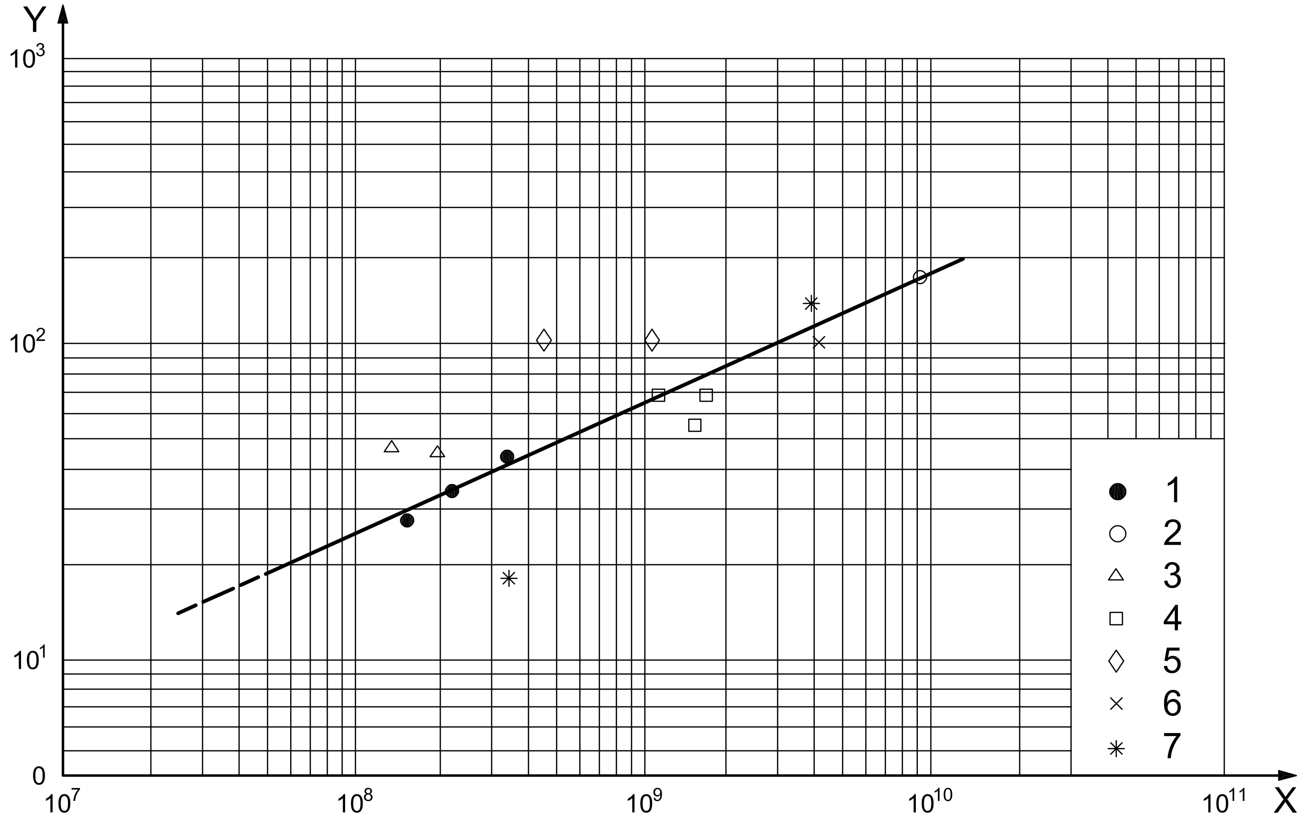
The F factor allows for the fact that not all the heat released in a flame can be transferred by radiation. Measurements of radiation from flames indicate that the fraction of heat radiated (radiant energy per total heat of combustion) increases toward a limit, similar to the increase in the burning rate with increasing flame diameter.

F factor data from the U.S. Bureau of Mines [95] for radiation from gaseous-supported diffusion flames are given in Table 10. These data apply only to the radiation from a flame from subsonic flares. If liquid droplets of hydrocarbon larger than 150 µm in size are present in the flame, the values in Table 10 should be somewhat increased. If the flame is not entirely smokeless, the effective overall F-factor can be less than the values in Table 10. Exit velocity and flare-tip design can also influence the F-factor.

Two methods are presented in Annex C for considering radiation levels. The example in C.2 is the simple approach that has been used for many years. It uses Figures 7 and 8 to determine an estimated flame length. The wind tilts the flame in the direction the wind is blowing. The wind effect is obtained from Figure 9, which relates horizontal and vertical displacement of the flame to the ratio of lateral wind velocity to stack velocity. A wind velocity of 9 m/s (20 mph) is a common assumption for most radiation calculations. The flame radiation epicentre is located at the centre of a straight line drawn between the flare tip and the end of the flame. Figures 7 through 9 should be used only for subsonic flares and the flare manufacturer should be consulted for sonic flares.

The methods presented here assume that a flame can be modelled by a single point source for radiation. The radiation flux that is modelled should comply with this assumption when determining spacing and radiation exposure. If the point of interest is too close to the flame for the single point assumption, more complex radiation analysis should be employed.

The location of the flame centre is quite significant when radiation levels are examined. Flame length varies with emission velocity and heat release. Information on this subject is limited and is usually based on visual observations in connection with emergency discharges to flares. Figures 7 and 8 were developed from some plant-scale experimental work on flame lengths covering relatively high release rates of various mixtures of hydrogen and hydrocarbons.



Key

X heat release, expressed in watts

Y flame length (including any lift-off), expressed in metres

1 fuel gas (508-mm stack)

2 Algerian gas well

3 catalytic reformer — recycle gas (610-mm stack)

4 catalytic reformer — reactor effluent gas (610-mm stack)

5 dehydrogenation unit (305-mm stack)

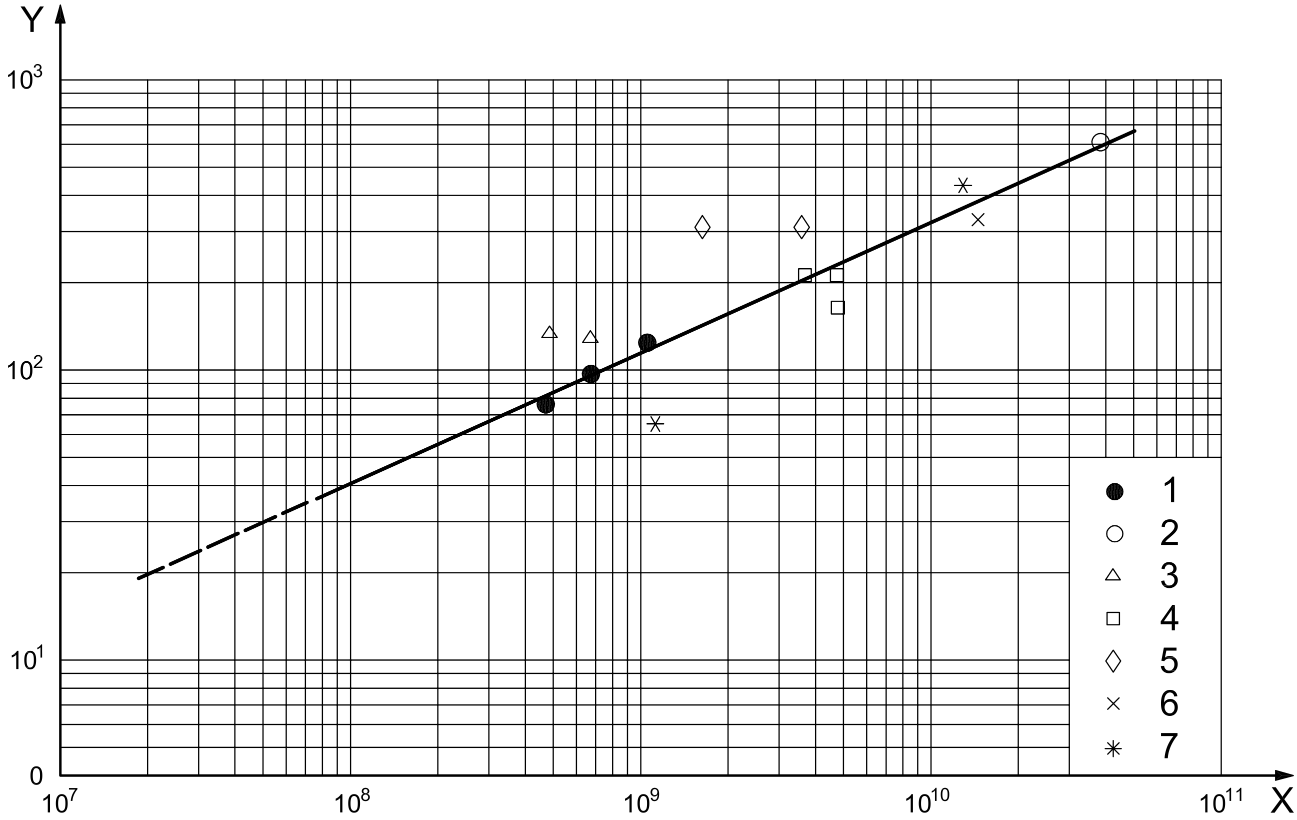
6 hydrogen (787-mm stack)

7 hydrogen (762-mm stack)

NOTE 1 This figure was converted from Figure 8.

NOTE 2 Multiple points indicate separate observations or different assumptions of heat content.

Figure 7 — Flame length versus heat release — Industrial sizes and releases (SI units)



Key

X heat release, expressed in British thermal units per hour

Y flame length (including any lift-off), expressed in feet

1 fuel gas (20-inch stack)

2 Algerian gas well

3 catalytic reformer — recycle gas (24-inch stack)

4 catalytic reformer — reactor effluent gas (24-inch stack)

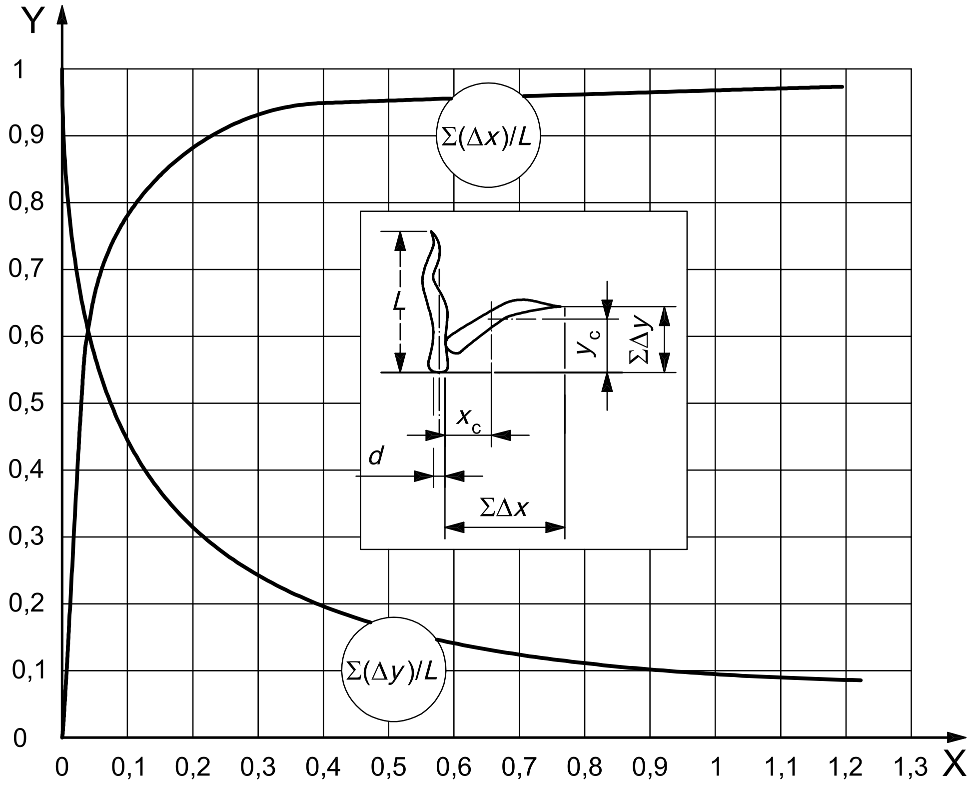
5 dehydrogenation unit (12-inch stack)

6 hydrogen (31-inch stack)

7 hydrogen (30-inch stack)

NOTE Multiple points indicate separate observations or different assumptions of heat content.

Figure 8 — Flame length versus heat release — Industrial sizes and releases (USC units)



Key

X $\sum(u_\infty/u_j)$

Y $\sum \Delta y/L$ or $\sum \Delta x/L$

u_∞ is the lateral wind speed

u_j is the jet exit velocity

Insert shows the flame geometry in still air and lateral wind.

Figure 9 — Approximate flame distortion due to lateral wind on jet velocity from flare stack

Table 10 — Radiation from gaseous diffusion flames

Gas	Burner diameter cm	Fraction of heat radiated
Hydrogen	0,51	0,095
	0,91	0,091
	1,90	0,097
	4,10	0,111
	8,40	0,156
	20,30	0,154
	40,60	0,169
Butane	0,51	0,215
	0,91	0,253
	1,90	0,286
	4,10	0,285
	8,40	0,291
	20,30	0,280
	40,60	0,299
Methane	0,51	0,103
	0,91	0,116
	1,90	0,160
	4,10	0,161
	8,40	0,147
Natural gas (95 % CH_4)	20,30	0,192
	40,60	0,232

Several formulas for calculating flame length and approximating flame tilt are presented in the literature [70], [94], [96], [97], [98]. Each formula has its own special range of applicability and should be used with caution, particularly since the combined impact of several factors (radiation, radiant heat fraction, flame length and centre and flame tilt) shall be considered.

The example in C.3 is another approach to calculating the probable radiation effects, using the more recent method of Brzustowski and Sommer [94]. The principal difference between these methods is the location of the flame centre. The curves and graphs necessary to simplify the calculations are included in Annex C.

There are other methods that can be utilized to calculate radiation from flares. More sophisticated models that consider wind velocity, exit flare gas velocity, flame shape and flame segmental analysis can be appropriate for special cases, especially with large release systems.

Most flare manufacturers have developed proprietary radiation programs based on empirical values. The F factor (fraction of heat radiated) values used in these programs are specific to the equations used, and might not be interchangeable with the F factor values used in Equation (24). These programs have not been subject to review and verification in the open literature. The user is cautioned to assess the applicability of these methods to his or her particular situation.

6.4.3 Combustion methods

6.4.3.1 General

Disposal of combustible gases, vapours and liquids by burning is generally accomplished in flares. Flares are used for environmental control of continuous flows of excess gases and for large surges of gases in an emergency. The flare is usually required to be smokeless for the gas flows that are expected to occur from normal day-to-day operations. This is usually a fraction of the maximum gas flow, but some environmentally sensitive areas require 100 % smokeless or even a fully enclosed flare. The smokeless burning expectations should be explicitly defined. Attempts to shortcut the establishment of factually based smokeless burning requirements by setting the smokeless flow rate as a percentage of the maximum emergency flow rate can lead to disappointment or needless expense.

Various techniques are available for producing smokeless operation, most of which are based on the premise that smoke is the result of a fuel-rich condition and is eliminated by promoting uniform air distribution throughout the flames (see 6.4.2.2). In 6.4.3.2 is provided a description of the most common techniques employed for providing smokeless operation. In addition to smokeless operating requirements, stricter flaring regulations (federal, state and local) are constantly evolving and, in most areas, typically include low noise levels, limits on smoking reliefs, continuous pilot monitoring and limits on tip-exit velocities and minimum heat content of the flare gas. Current regulations should always be consulted for detailed flaring requirements.

6.4.3.2 Flare systems designs

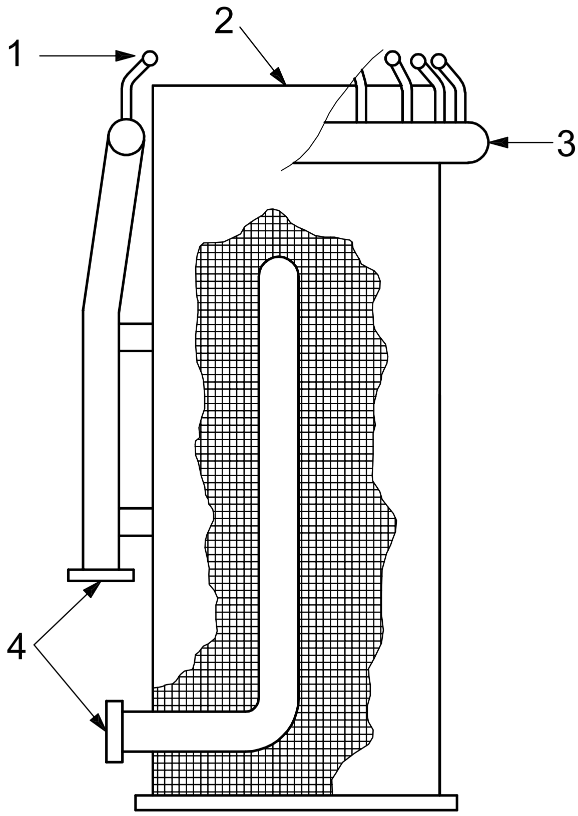
6.4.3.2.1 Smokeless operation is normally the overriding requirement when designing the burner for a flare system. Almost every flare design is aimed at inducing smokeless operation under a certain set of flare-gas or utility-availability conditions. To promote even air distribution throughout the flames (and thus prevent smoke formation), energy is required to create turbulence and mixing of the combustion air within the flare gas as it is being ignited. This energy can be present in the gases, in the form of pressure, or it can be exerted on the system through another medium, such as injecting high-pressure steam, compressed air or low-pressure blower air into the gases as they exit the flare tip. To create conditions favourable for smokeless combustion, flare designs range in complexity from a simple open pipe with an ignition source to integrated, staged flare systems with complex control systems. In 6.4.3.2.2 and 6.4.3.2.3 is a short summary of the most common types of flaring systems.

6.4.3.2.2 The simplest flare-tip design is commonly referred to as a utility or pipe-flare tip and can consist of little more than a piece of pipe fitted with a flame retention device for flame stability at higher exit velocities (the upper portion is typically stainless steel to endure the high flame temperatures) and a pilot for gas ignition. This plain design has no special features to prevent smoke formation, and consequently should not be used in applications where smokeless operation is required, unless the gases being flared, such as methane or hydrogen, are not prone to smoking. Flare tips of this style should include a flame-retention device (to increase flame stability at high flow rates) and one or more pilots (depending upon the diameter of the tip). Windshields or heatshields are usually added on flare tips to reduce flame lick on the outside of the tip. An inner refractory lining is also common with larger diameter tips to minimize thermal degradation caused by internal burning at low rates (known as burnback).

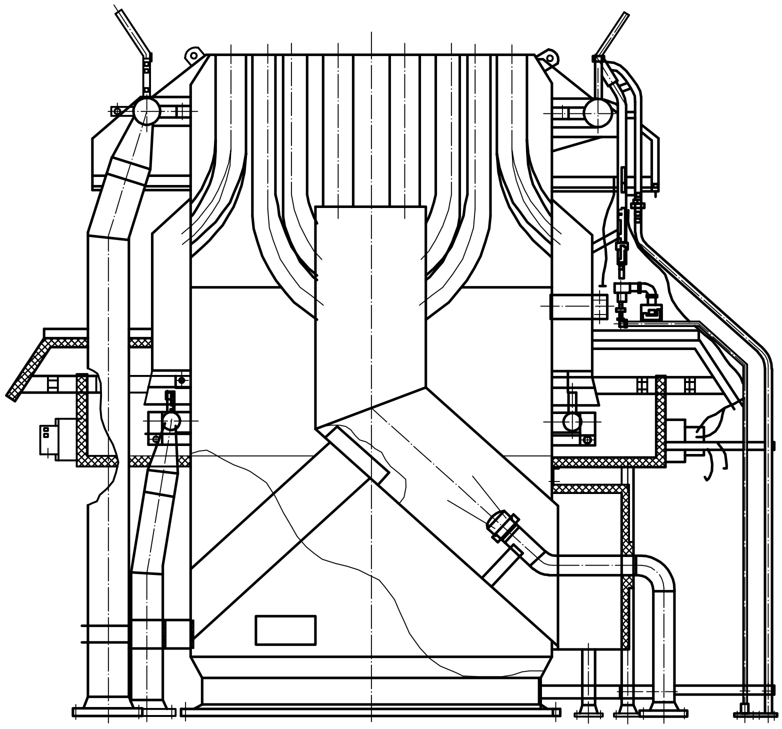
6.4.3.2.3 Flare tips that use steam to control smoking are a common form of smokeless flare tip. The steam can be injected through a single pipe nozzle located in the centre of the flare, through a series of steam/air injectors in the flare, through a manifold located around the periphery of the flare tip or a combination of all three, as appropriate for a particular application [see Figures 10 a) and 10 b)]. The steam is injected into the flame zone to create turbulence and/or aspirate air into the flame zone via the steam jets.

This improved air distribution allows the air to react more readily with the flare gases to eliminate the fuel-rich conditions that result in smoke formation. Another factor assisting smokeless operation is the steam water-gas shift interaction where carbon monoxide and water vapour react to form carbon dioxide and hydrogen, which is more easily burned. Proprietary tip designs that offer unique steam injection methods and varying resultant steam efficiencies are available from various manufacturers.

The amount of steam required is primarily a function of the flare-gas composition, flow rate, and steam pressure and flare-tip design (see Table 11). Although steam is normally provided from a supply header at a gauge pressure of 700 kPa to 1 000 kPa (approx. 100 psi to 150 psi), special designs are available for utilizing steam-gauge pressure as low as 200 kPa (approx. 30 psi). The major impact of lower steam pressure is a reduction in steam efficiency during smokeless turndown conditions.



a) Normal



b) Low noise

Key

1 steam tips 2 flame holder 3 steam manifold 4 steam connections

Figure 10 — Steam-injected smokeless flare tips

Table 11 — Suggested injection steam rates

Gases being flared	Steam required [a] kg (lb) of steam per kg (lb) of hydrocarbon gas
Paraffins	
Ethane	0,10 to 0,15
Propane	0,25 to 0,30
Butane	0,30 to 0,35
Pentane plus	0,40 to 0,45
Olefins	
Ethylene	0,40 to 0,50
Propylene	0,50 to 0,60
Butene	0,60 to 0,70
Diolefins	
Propadiene	0,70 to 0,80
Butadiene	0,90 to 1,00
Pentadiene	1,10 to 1,20
Acetylenes	
Acetylene	0,50 to 0,60
Aromatics	
Benzene	0,80 to 0,90
Toluene	0,85 to 0,95
Xylene	0,90 to 1,00

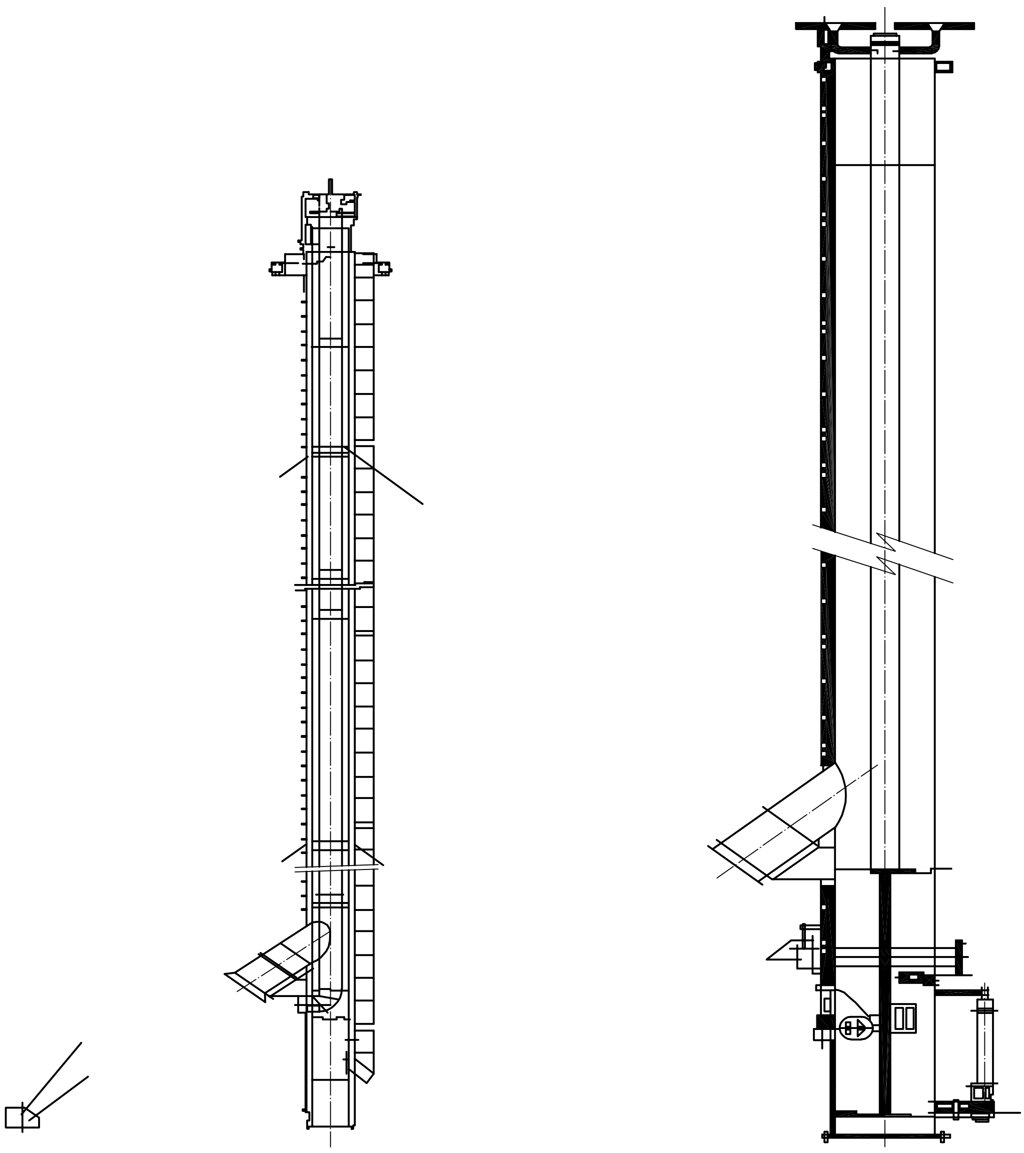
[a] The suggested amount of steam that should be injected into the gases being flared in order to promote smokeless burning (Ringlemann 0) can be determined from this table. The given values provide a general guideline for the quantity of steam required. Consult the flare vendor for detailed steam requirements.

In cold climates, an internal steam nozzle can cause condensate to enter the flare stack and header, collect and freeze. In some instances, this has resulted in complete blockage of the flare stack or flare header. Therefore, consideration should be given to supplying steam to an internal steam nozzle through a separately controlled steam line so that it can be turned off in cold conditions.

6.4.3.2.4 High-pressure air can also be used to prevent smoke formation. This approach is less common because compressed air is usually more expensive than steam. However, in some situations with low smokeless capacities, it can be preferable, for example, in arctic or low-temperature applications where steam can freeze and plug the flare tip/stack. Also, other applications include desert or island installations where there is a shortage of water for steam, or where the waste-flare gas stream reacts with water. The same injection methods described for steam (6.4.3.2.3) are used with compressed air. The air is usually provided at a gauge pressure of 689 kPa (100 psi) and the mass quantity required is approximately 200 % greater than required by steam, since the compressed air does not produce the water-gas shift reaction that occurs with steam.

6.4.3.2.5 High-pressure water, while quite uncommon, is also used to control smoking, especially for horizontal flare applications and when it is necessary to eliminate large quantities of waste water or brine. One lb (0,45 kg) of water at a gauge pressure 350 kPa to 700 kPa (approx. 50 psi to 100 psi) is usually required for each 0,45 kg (1 lb) of gas flared. Freeze protection is required in cold climates and, because of the difficulty in controlling the water flow at low flaring rates, usually requires a staged water-spray injection system.

6.4.3.2.6 A low-pressure forced-air system is usually the first alternative evaluated if insufficient on-site utilities are available to aid in producing a smokeless operation. The system creates turbulence in the flame zone by injecting low-pressure air supplied from a blower across the flare tip as the gases are being ignited, thus promoting even air distribution throughout the flames. Usually, air at a gauge pressure of 0,5 kPa to 5,0 kPa (2 in H_2O to 20 in H_2O) flows coaxially with the flare gas to the flare tip where the two are mixed. This system has a higher initial cost due to the requirement for a dual stack and an air blower. See Figure 11. However, this system has much lower operating costs than a steam-assisted design (requiring only power for a blower). The additional quantity of air supplied by the blower for smokeless operation is normally 10 % to 30 % of the stoichiometric air required for saturated hydrocarbons and 30 % to 40 % of the stoichiometric air required for unsaturated hydrocarbons.



a) Guy-supported air-assisted with ladders/platforms and dual blowers

b) High-pressure sonic air-assisted with self-supporting stack and with KO drum in base

Figure 11 — Typical flare systems

6.4.3.2.7 A high-pressure system does not require any utilities such as steam or air to promote smokeless flaring. Instead, these systems utilize pressure energy available within the flare gas itself [typically a gauge pressure of 35 kPa to 140 kPa (5 psi to 20 psi) minimum at the flare tip] to eliminate fuel-rich conditions and resulting smoke within the flames. High-pressure system limitations are also present but vary by manufacturer and nature of design. By injecting the flare gas into the atmosphere at a high pressure, turbulence is created in the flame zone, which promotes even air distribution throughout the flames. Since no external utilities are required, these systems are normally advantageous for disposing of very large gas releases, both from the economics of smokeless operation and the control of flame shape. The individual tips used have relatively small capacities, and larger system designs can require that many tips be manifolded together. Maintaining sufficient tip pressure during turndown conditions is critical and often requires that a staging system be employed to proportionately control the number of flare tips in service with relationship to the amount of gas flowing. Staged-flare systems can be mounted either at grade or elevated; however, the larger systems can require ground-level designs since numerous tips are required (it is not uncommon to have more than 300 tips in a large staged-flare system) and the tips shall be evenly spaced to allow air entry into the system.

Staged flares should provide backup for system failures by inclusion of bypasses or emergency vents. Bypasses around control valves are a common safety measure. These typically utilize a rupture disk or similar device.

6.4.3.2.8 All of the preceding descriptions have been for flare equipment to dispose of exothermic flare gases; that is, gases that have a high enough heating value [usually greater than 74,5 MJ/m^3 (200 Btu/Scf) for unassisted flares and 112 MJ/m^3 (300 Btu/Scf) for assisted flares] to sustain combustion on their own without any auxiliary fuel additions. Endothermic gases can be disposed of in thermal incineration systems; however, there are situations where the preferred approach is to use a special flare design. These flares utilize auxiliary fuel gas to burn the flare gases. With small gas flow rates, simple enrichment of the flare gases by adding fuel gas in the flare header to raise the net heating value of the mixture can be sufficient. In other situations, it can be necessary to add a fuel-gas injection manifold around the flare tip (similar to a steam manifold) and build a fire around the exit end of the flare tip through which it is necessary for the gases to flow. Dilute ammonia or high CO_2 composition flare gases with small amounts of H_2S are common applications where the addition of fuel gas is required.

6.4.3.2.9 High-pressure fuel gas can also be used to prevent smoke formation by entraining outside air into the flare flame and generating turbulence to assist overall combustion. Usually, the injection methods are similar to steam tips, but special high-performance tips are used to reduce the amount of assist gas. If natural gas is used as the assist gas, typically 0,5 kg to 0,75 kg of assist gas per kilogram of flare gas is required, based on a flare gas consisting of normal paraffinics such as propane and butane. The gas-gauge pressure for natural gas assist is typically 500 kPa (approx. 75 psi) (minimum) with 1 000 kPa (approx. 150 psi) preferred.

6.4.3.3 Enclosed ground flares

Ground flares encompass a broad range of vastly different types of flare systems. In general, any of the flare tips or systems discussed in 6.4.3.2 can be mounted atop an elevated stack or mounted at grade. With increasingly strict requirements regarding flame visibility, emissions and noise, enclosed ground flares can offer the advantages of hiding flames, monitoring emissions and lowering noise. However, the initial cost often makes them undesirable for large releases when compared to elevated systems. With an enclosed ground-flare system, a variety of tips or burners may be utilized and are enclosed or hidden behind a refractory-lined carbon-steel shell. A significant disadvantage with a ground flare is the potential accumulation of a vapour cloud in the event of a flare malfunction. As a result, special safety dispersion systems are usually included in the ground-flare system. For this reason, instrumentation for monitoring and controlling ground flares is typically more stringent than for an elevated system. These flares are typically the most expensive because of the size of the shell or fence and the additional instrumentation that can be required to monitor these key parameters. Another significant limitation is that enclosed ground flares have significantly less capacity than elevated flares.

If emissions monitoring is not required, a fenced ground flare system can be designed with very large capacity. A radiation/wind fence can partially or totally hide the flames from view to a person located near grade. By restricting the amount of flame visible to a point of interest at grade level locations, it is possible to greatly reduce the external radiation load from the flare. Fenced ground flares frequently use multiple, high-pressure burners to obtain smokeless performance at firing rates that cannot normally be handled smokelessly by elevated flares.

6.4.3.4 Elevated flares

The most common type of flare system currently in use is an elevated flare. In these systems the flare tip is mounted atop the stack, which reduces ground-level radiation and improves the toxicity-dispersion profile. There are three common stack support methods:

a) self-supported: Self-supported stacks are normally the most desirable. However, they are also the most expensive because of greater material requirements needed to ensure structural integrity over the anticipated conditions (wind, seismic and the like). They require only enough land area for the foundation and the ability to meet safe ground-level thermal-radiation levels, but are normally limited (economically versus alternatives) to a stack height of 60 m to 90 m (approx. 200 ft to 300 ft). See Figure 12 a).

b) guy-wire-supported: These are the least expensive but require the largest land area due to the guy-wire radius requirements. The typical guy-wire radius is equal to one-half the overall stack height. Guyed stacks of heights of 180 m to 250 m (approx. 600 ft to 800 ft) have been used. See Figure 12 b).

c) derrick-supported: These are used only on larger stacks where self-supported design is not practical, or available land area excludes a guy-wire design. Some derrick designs allow the flare stack and tip to be lowered to grade on movable trolleys for inspection and maintenance. This self-lowering design is especially useful when multiple stacks are installed on the same derrick. In locations where land is not available, the multi-flare derrick can be used. See Figure 12 c).

Additional information regarding elevated flares and support structures and their structural design can be found in API Std 537.

6.4.3.5 Unassisted pipe flare

An unassisted pipe flare is used where smokeless burning assist is not required. Pilots and a pilot-ignition system provide flare flame ignition.

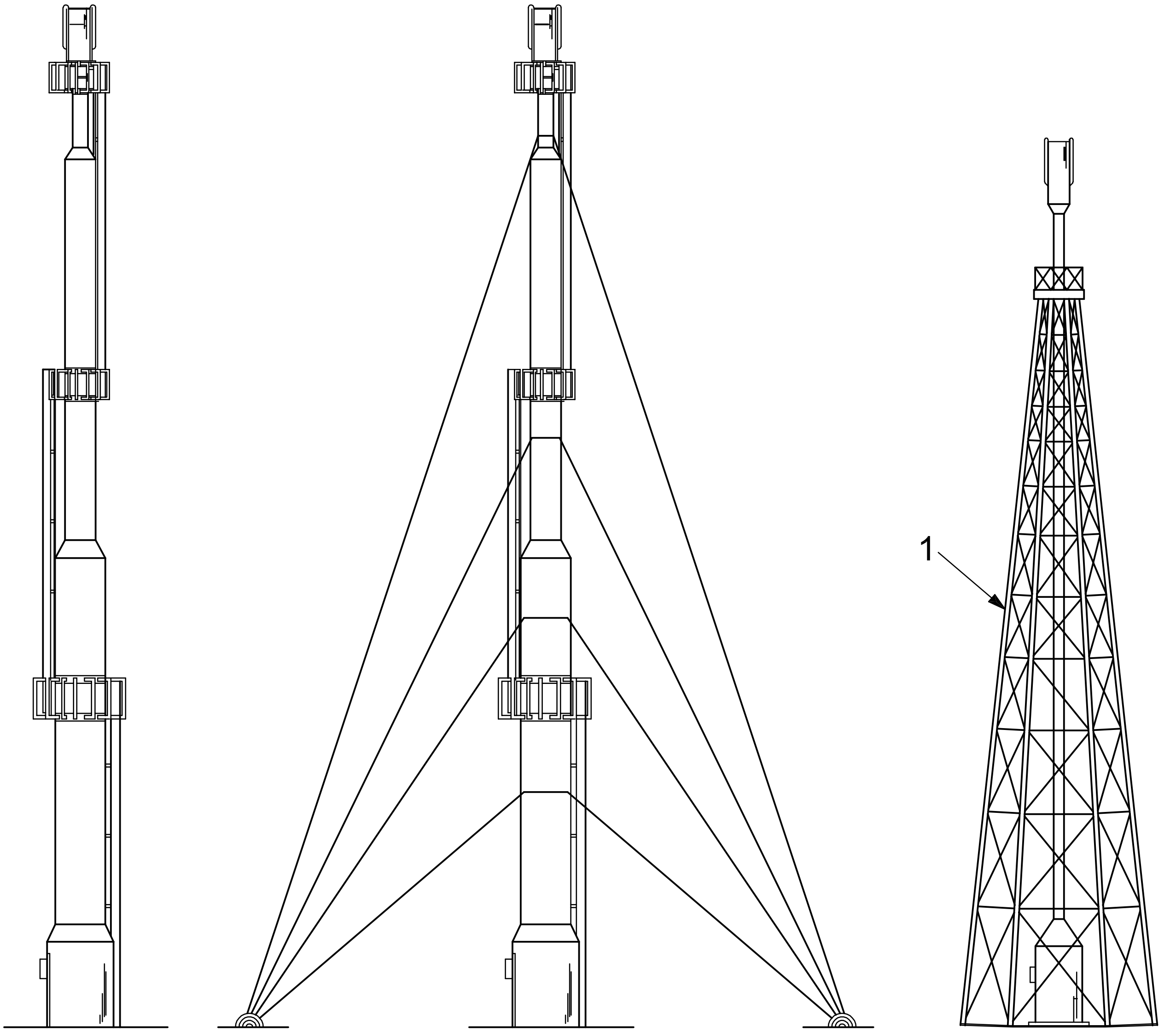
The pipe-flare tip may have a mechanical device or other means of establishing and maintaining a stable flame. The ignition fire from the gas discharge is initially ignited by interaction with the pilot(s) flames. Once the pilot lights and the flame stabilizes, the flare should maintain flame stability over the operating design range. Flame stability for a pipe flare is primarily dependent upon the selection of the gas exit velocity.

The form of the flame produced by an unassisted pipe flare is a function of the relief-gas composition and the gas-exit velocity. At greater gas-exit velocities, the flame uses the gas-discharge energy to pull combustion air into the flame. It produces a shorter, more erect flame that has greater resistance to wind deflections. At lower gas-exit velocities, air is drawn to the flame by the buoyancy of the heated products of combustion. A buoyant flame is typically softer, longer and more affected by wind than a flame that uses higher gas-exit velocities.

Low gas-exit velocities and buoyancy-dominated flames are preferred for successful combustion of low-heating-value relief gas. High gas-exit velocities are preferred for higher-heating-value hydrocarbon relief gases or for relief gases rich in hydrogen. Because of the high flame velocity, wide flammability range, buoyancy effects and noise, hydrogen flares require special design considerations. The manufacturer should be consulted for details.

Flare combustion noise is influenced by gasexit velocity. Increased relief-gas-exit velocity can produce greater combustion turbulence and higher combustion noise. The highest combustion-noise levels are realized when a flare tip operates at gas-exit velocities where combustion instabilities occur.

The relief-gas discharging from the flare tip shall occur within the hydraulic design for the flare system (within the allowable pressure drop and flame combustion velocity limits). It shall be ignited and burned with the designed flame characteristics.



a) Self-supporting b) Guy-supported c) Derrick-supported

Key

1 derrick

Figure 12 — Flare structures

6.4.3.6 Auxiliary flaring equipment

6.4.3.6.1 If used, the purpose of a liquid seal in a flare system includes the following:

— to prevent any flashback originating from the flare tip from propagating back through the flare system;

— to maintain a positive system pressure to ensure no air leakage into the flare system and permit use of a flare gas recovery system;

— to provide a method of flare staging between an enclosed flare and a full size emergency flare;

— to prevent an ingress of air into the flare system during sudden temperature changes or condensation of flare gas, such as can occur following a major release of flare gas or following a steaming operation.

Liquid seals are located between the main knockout drum and the flare stack and are quite often incorporated into the base of the stack. They are sized for the maximum vapour-release case. When equipment, piping elevations and other factors permit, liquid-seal volume and seal-leg height should be sufficient to prevent the seal from being broken as a result of the vacuum formed in the flare header following a major release of flare gas or steaming operation.

For facilities that have cryogenic products in the flare header, consideration should be given to the effect of the cold material on the seal liquid medium. Water seals are not recommended where there is a risk of obstructing the flare system due to an ice plug. Alternate sealing fluids such as glycol/water mixture may be considered. Alternatively, methods such as heating the seal fluid or draining the seal when cold temperature is detected have been used.

6.4.3.6.2 For safety purposes, a pre-commissioning purge and subsequent continuous purge with a non-condensable oxygen-free gas is required through the flare system. The pre-purge displaces any existing air from the stack and the continuous purge ensures that atmospheric air does not enter the stack through the flare tip during low-flow conditions. There should, then, be a continuous purge of auxiliary gas, which may be gas from normal process vents (provided that the required flow rate can be maintained). The requirements for a continuous purge can be eliminated if a liquid seal is located near the base of the stack (refer to 6.4.3.6.1). This requires special precautions in the design of the stack to assure viability in the event of an internal explosion. It also can allow air to infiltrate to the liquid seal, which, for some seal mediums, carries other requirements.

Air present in the stack can create a potentially explosive mixture with incoming flare gas during low-flare gas flow rate conditions. There are two common types of mechanical seals, usually located at/or below the flare tip, that are used to reduce the amount of continuous purge gas required to prevent air infiltration into the flare stack:

a) buoyancy seal:

   This type of seal uses the difference in the relative molecular masses of the purge gas and infiltrating air to form a gravity seal that prevents the air from entering into the stack. A baffled cylinder arrangement forces the incoming air through two 180° bends (one bend up and one bend down) before it can enter into the flare stack. If the purge gas is lighter than air, the purge gas accumulates in the top of the seal and prevents the air from infiltrating the system. If the purge gas is heavier than air, the purge gas accumulates in the bottom of the seal and prevents air from infiltrating. This seal normally reduces the purge-gas velocity required through the tip to 0,003 m/s (0,01 ft/s). Also, with most purge gas compositions, this rate limits oxygen levels below the device to less than 0,1 %. Higher purge-gas velocities can be required to avoid burnback within the flare tip. The two 180° turns in a buoyancy seal can cause liquid collection in the seal (see Figure 13) in which case a drain is required. For those flare tips that have a refractory lining, debris from the flare tip refractory can potentially plug the buoyancy-seal drain, resulting in a possibly unsafe condition. The drain shall be kept open and protected from freezing in cold climates.

b) velocity seal:

   This seal works under the premise that infiltrating air enters through the flare tip and hugs the inner wall of the flare tip. The velocity seal is a cone-shaped obstruction, with single or multiple baffles, which forces the air away from the wall. It subsequently encounters the focused purge-gas flow and is swept out of the tip. This seal normally reduces the purge gas velocity through the tip to between 0,006 m/s to 0,012 m/s (0,02 ft/s and 0,04 ft/s), which keeps oxygen concentrations below the seal to 4 % to 8 % (approximately 50 % of the limiting oxygen concentration required to create a flammable mixture). Higher purge gas velocities can be required to avoid burn-back within the flare tip. Caution should be exercised when the waste-gas stream can contain hydrogen, ethylene or other gases with wide explosive limits. In such cases, a higher purge rate can be required to avoid an explosive mixture with air. A hole should be made in each baffle plate to permit drainage of possible liquids in order to avoid corrosion and/or freezing.

Without either one of these seals, the purge gas velocity in the tip required to prevent air infiltration into the stack should be determined using the procedure described 7.3.3.3.3.

CAUTION — Purge-reduction seals are not flame arrestors; that is, they will not stop a flashback. They are designed as energy-conservation devices to reduce purge-gas flows required to mitigate air infiltration into the stack. In the event of loss of purge-gas flow (and assuming that there is no other waste-gas flow) the oxygen level below the velocity type seal will almost immediately begin to increase. In the case of the buoyancy seal there is a time delay between the moment the purge gas flow stops and the time the oxygen level below the seal begins to increase.

Buoyancy and velocity seals are typically not applicable to multi-point, staged flares. In staged systems, the piping downstream of the staging device may be purged with a non-condensable inert gas following each closure of the staging device.

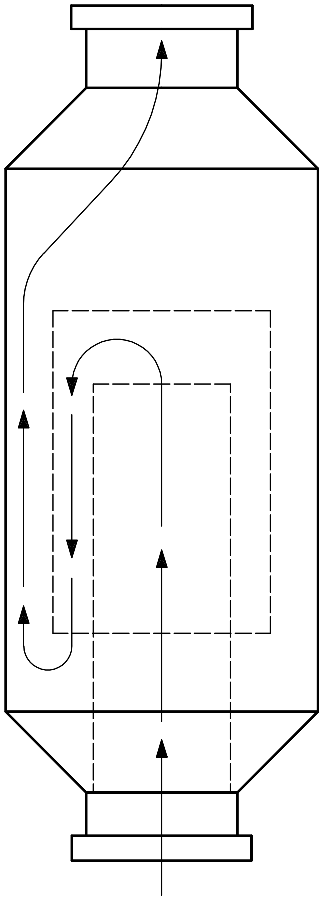


Figure 13 — Purge-reduction seal — Buoyancy seal

6.4.3.6.3 The following methods of controlling steam (or compressed air, water and so on) for smokeless flare control are common (many other strategies are possible):

a) manual operation:

Manual control usually involves remote operation of a steam valve by operating personnel assigned to a unit from which the flare is readily visible. This method is satisfactory if short-term smoking can be tolerated when a sudden increase in flaring occurs. With a manual arrangement, close supervision is required to ensure that the steam flow is reduced following the correction of an upset. Operating costs can be excessive if monitoring is not timely.

b) television monitoring with manual control:

The philosophy is the same as with manual operation except that a television monitoring system is added so that control room operators can monitor and control the steam flow more effectively.

c) feed forward control system for pressure, mass flow or velocity:

By measuring the amount of flare gas flowing to the flare, the steam rate can be automatically adjusted to compensate for rate changes. This system might not be desirable if the composition of the gas being flared varies widely over time (in other words, paraffins to olefins or aromatics, hydrogen, or various mixtures thereof).

d) feedback system using an infrared sensor:

Infrared sensors can be used to detect smoke formation in the flames and automatically adjust the steam control valve to compensate. A disadvantage of this system is that infrared waves are absorbed by moisture and the resultant feedback signal is reduced in rainy or foggy conditions.

At low flaring rates, fluctuations in either pressure or flow are so minute that very sensitive instruments are required to provide sufficient steam for smokeless combustion while at the same time avoiding waste. Therefore, controls shall be carefully sized, precisely adjusted and properly installed to obtain satisfactory operation.

Proper steam (or other assist media) control is essential if the flare tip is to achieve its maximum reaction-efficiency potential. This is particularly true at low flaring rates where it is possible to over-steam the flare to the point of near flame extinguishment. Such operation can be based on the belief that it is preferable to over-aerate a flame that it becomes difficult to see. This is not correct. Studies and tests of flare reaction efficiency have established that an over-steamed flare has a lower efficiency than a properly operated flare. Over-steaming can also increase the noise generated by the flare.

6.4.3.6.4 A pilot-ignition panel is an integral part of any flare system utilizing flame-front generation for ignition. A flame-front generator fills an ignition line leading from the panel to the flare pilot with a flammable mixture of air and fuel gas. A spark is then introduced at the panel to ignite the mixture and send a flame-front through the piping to ignite the pilot. These panels can be operated manually or they can be automated for pilot re-ignition when pilot flame-out is detected. Electronic ignition systems that do not require flame-front propagation are also used. These systems typically ignite the pilot gas at the pilot itself. The user is advised to evaluate the selection of the flare ignition system for the specific application based on experience from similar plants.

6.4.3.6.5 Several methods of pilot monitoring are available, including thermocouples installed within the pilot head, ionization monitoring within the pilot head, and remote acoustic, infrared or optical monitors. Experience has shown that thermocouples can fail due to high-temperature exposure. The user should base the pilot-monitoring system on relevant experience for the specific application. Further details concerning the design of pilot-monitoring systems can be found in API Std 537.

6.4.3.6.6 Aircraft warning lights are usually required only when flare heights are greater than 61 m (200 ft) or when the stack is close to an airport. The type of lights and number required are usually regulated. Consideration should be given to maintenance accessibility in determining the types of lights to use.

6.4.3.6.7 Some flare systems require a flare knockout drum to separate liquid from gas in the flare system and to hold the maximum amount of liquid that can be relieved during an emergency situation.

Knockout drums are typically located on the main flare line upstream of the flare stack or any liquid seal. If there are particular pieces of equipment or process units within a plant that release large amounts of liquid to the flare header, it is desirable to have knockout drums inside the battery limits to collect these liquids. This reduces the sizing requirements for the main flare knockout drum, as well as facilitates product recovery.

In general, a flare can handle small liquid droplets. However, a knockout drum is required to separate droplets larger than 300 µm to 600 µm in diameter in order to avoid burning liquid outside the normal flame envelope. If unit knockout drums are provided upstream of the main flare knockout facilities, these drums may be sized to separate droplets typically greater than 600 µm in diameter. The use of unit knockout drums effectively reduces the sizing requirement for the main flare knockout drum and facilities. See 7.3.2.1.

The liquid holdup capacity of a flare knockout drum is based on consideration of the amount of liquid that can be released during an emergency situation without exceeding the maximum level for the intended degree of liquid disengagement. This hold-up should also consider any liquid that can have previously accumulated within the drum that was not pumped out. The hold-up times vary between users, but the basic requirement is to provide sufficient volume for a 20 min to 30 min emergency release. Longer hold-up times might be required if it takes longer to stop the flow.

It is important to realize as part of the sizing considerations that the maximum vapour release case might not necessarily coincide with the maximum liquid. Therefore, the knockout drum size should be determined through consideration of both the maximum vapour release case as well as the release case with the maximum amount of liquid.

6.4.3.7 Flaring toxic gases

The flaring of toxic gases requires special considerations. Some information can be obtained from a test programme sponsored by the Environmental Protection Agency (EPA) through the Chemical Manufacturers Association (CMA). The destruction efficiency for certain combustible toxic material in a properly operated flare can be in the range of 98 % [99].

Depending upon the gases being flared and the flare design being used, the minimum allowable net lower heating value should be in the range of 7,5 MJ/m^3 to 11,2 MJ/m^3 (200 Btu/Scf to 300 Btu/Scf). If the net lower heating value can drop below this range, a special flare design can be required. (See 6.4.3.2.8.)

To ensure safe operation during periods when the flare might not have a flame present, ground-level concentration calculations for hazardous components should be performed, assuming the flare is functioning as a vent only. Other safeguards can be necessary to mitigate ground-level exposure hazards. Reliable, continuous pilot monitoring is considered critical when flaring toxic gases.

6.4.3.8 Burn pits

Burn pits normally require excavation or bermed areas to contain liquid hydrocarbons or other objectionable materials produced by incomplete combustion. Seepage from a poorly designed or maintained burn pit can pose a threat to groundwater supplies.

6.5 Disposal to a lower-pressure system

6.5.1 General

Discharge of the relieved material to the same or another system at lower pressure can be a safe and economical method, provided that the receiving system is designed for the additional load.

6.5.2 Sewer

Non-volatile liquid discharges from pressure-relief devices may be piped to sewer drains, provided that the sewer system has adequate capacity and is properly sealed and vented. Caution should be exercised to avoid discharging volatile, toxic or hot fluids into a sewer.

6.5.3 Process

The particular type of process unit selected determines whether a lower-pressure process system exists that can safely receive material relieved from a higher-pressure system. This is usually true with liquid reliefs (e.g. liquid relieved from the discharge side of a pump being disposed to the suction side). Selection of the type of valve used (that is, a balanced or a conventional valve) depends on the back pressure (constant, variable or built-up) of the lower pressure system.

6.6 Disposal of liquids and condensable vapours

6.6.1 General

The selection of a disposal system for liquids and condensable vapours, not covered in 6.3 through 6.5, is determined on the basis of the items covered in 6.6.2 through 6.6.5.

6.6.2 Temperature

6.6.2.1 General

Only a comprehensive study of the plot plan and individual pressure-relief-device data can determine the most desirable system for a particular plant. The methods of coping with temperature problems given in 6.6.2.2 through 6.6.2.4 are not meant to be limiting. They are included merely to illustrate a principle of separation of discharges.

6.6.2.2 Ambient

Non-volatile liquids at ambient temperatures may be discharged into a separate, closed relief header that discharges into a sump from which the liquids are recovered. Volatile or non-volatile liquids may be alternately discharged into the regular, closed disposal system. The liquid is disengaged at the knockout drum before the flare (see 7.3.2.1 for additional details).

6.6.2.3 Hot

Hot liquids and vapour may be cooled and condensed by one of the methods covered in a) through c) below.

a) Pressure-relief devices that discharge hot condensable hydrocarbon vapours or liquids may be piped into a separate header that terminates in a quench drum. In this service, quenching can reduce the temperature of the relief stream and may permit the use of less expensive materials in downstream equipment. Cooling also condenses some of the less volatile components and can reduce or prevent the release of hot condensable vapours to the atmosphere. A quench drum is a vessel equipped to spray a quenching liquid down through the hot discharged gases as they pass at reduced velocity through the drum. The quenching fluid may be water, gas oil, or another suitable liquid. The quenching liquid collects in the bottom of the drum for subsequent removal.

   One type of quench drum is a vertical vessel containing baffles that is connected by a means of a conical transition to a vent stack or flare. The condensable hydrocarbon material is fed into the drum below the baffles. Water is introduced into the drum above the baffles at a rate that depends on the temperature and the amount of hydrocarbon material being fed to the quench drum. The water spilling over the baffles desuperheats and condenses the hydrocarbon vapour, knocks out entrained hydrocarbon liquid and cools down the hydrocarbon liquid collected in the bottom of the drum. The uncondensed vapour and any steam formed pass up the vent stack or enter a flare system (see Figures D.2 and D.3).

b) The submerged discharge system is not extensively used in present day design. Care should be taken in its use and location when non-condensable gases that can escape to the atmosphere are present. Cooling of hot liquid and condensation of vapour by submerged discharge in a large body of cold liquid can have limited utility when considered as a disposal method to a lower-pressure system in the same process unit. Occasionally, steam is mixed with the effluent in sufficiently large quantities to make the discharge noncombustible. In this type of design, the pressure-relieving system on a unit that handles heavy hydrocarbons generally serves a dual purpose: as a disposal system for the pressure-relieving devices and as a dropout or blowdown system for furnaces and vessels.

   The submerged discharge system is a relieving system that terminates in parallel laterals submerged in a water-filled sump. Holes are cut in the bottom of the laterals throughout their length, imparting downward flow to the discharged effluent to obtain maximum agitation, cooling and condensing. Provisions shall be made to maintain a liquid level in the sump while the blowdown system is being used. The discharge is drained from this sump into a separator, where the oil and condensed vapours are removed from the water.

c) The use of shell-and-tube heat exchangers or coil-in-box coolers has the merit of separating cooled or condensed material immediately. In addition, the coil-in-box (emergency box cooler) cooler can remove some heat, which is desirable in emergencies when no cooling water is flowing.

6.6.2.4 Cold

Low-temperature fluids require considerations similar to those outlined for hot streams, particularly if there is a possibility of low-boiling liquids entering the disposal system. Autorefrigeration will occur as liquid boils at the reduced pressure. If the equilibrium temperature is sufficiently low, piping and drums fabricated of materials designed for low temperature may be required to eliminate the risk of brittle fracture [100]. In such circumstances, consideration should be given either to a completely separate low-temperature system or isolation of the stream until it reaches a knockout drum where the liquid can disengage. Vapours vented off the drum can often be safely combined with other disposal systems if, in the absence of liquid, the heat pickup (of the piping system) from the surrounding atmosphere will prevent temperatures from dropping to a dangerously low level.

6.6.3 Hazardous properties

The safe disposal of material that has toxic, acidic, alkaline, or corrosive properties may require chemical neutralization, absorption, or reaction in a special disposal system. Dilution with air or water to a safe level may be satisfactory in some cases.

6.6.4 Viscosity and solidification

In the selection of a disposal system for liquids and condensable vapours, the production of highly viscous or solid materials warrants consideration. The design of a disposal system for such materials may require heat tracing of valves and discharge lines. The formation of gums, polymers, coke, or ice that might prevent safe operation of the discharge system should also be considered in the design.

6.6.5 Miscibility

Solubility or miscibility of the material with water and avoidance of the formation of emulsions should be considered in the selection of a disposal system.

7 Disposal systems

7.1 Definition of system design load

7.1.1 General

7.1.1.1 The entire process of developing disposal-system design loads can be a complex process requiring input from process, operations, instrumentation and other engineering disciplines. The disposal-system boundary includes the piping, vessels and other equipment from the relief-device outlet to the final disposal point. Sizing of the disposal system can impact the relief-device operation (i.e., back pressure on pressure-relief valves, derating, etc.).

7.1.1.2 Although required relieving rates from individual devices for single-jeopardy events are known, it is necessary to determine the combined effects on the disposal system, for example

a) loss of cooling water and/or instrument air compressor as a result of loss of power;

b) loss of process reboil heat from a downstream column;

c) loss of instrument air might not cause loss of power; however, loss of power can directly cause loss of instrument air (e.g. loss of power to air compressor);

d) instrumentation impacts (favourable or unfavourable) can require complex analysis;

e) load reduction credits.

7.1.1.3 The following are general steps in establishing disposal system design loads.

a) As a starting point, establish the required relieving loads for the contingencies described in Clause 4 for each individual relief device discharging into the disposal system. Consider the potential disposal-system loads from pressure-control valves or emergency depressuring valves.

b) Determine which pressure systems are jointly affected by single contingencies (see 7.1.2).

c) Establish the maximum load into the disposal system during these contingencies (see 7.1.3).

d) Establish the design load for the disposal system (see 7.1.4).

7.1.2 Loads from pressure systems

The contingencies to be considered in defining relieving requirements are discussed in Clause 4. To define the system load, it is not necessary to assume the simultaneous occurrence of two or more unrelated contingencies. For example, an inadvertently closed valve at the same time a utility failure occurs is not typically considered. Therefore, the analysis should focus on individual initiating events and the resultant effects.

Particular study is required for cases of failure of major utilities, such as power or cooling medium. Partial failure and total failure of electrical power, steam, cooling medium, heating medium and instrument air to an entire plant should be considered. There are cases where partial failures result in higher loads than total failures. This type of study, with reference to electrical-power failure, commonly results in a design based on the failure of one bus, although loss of an entire distribution centre or of the incoming line can govern the design.

Interaction of utilities should also be considered. For example, loss of power can lead to loss of instrument air, steam, heating medium and/or cooling medium. The most common basis for analysing cooling-medium or heating-medium failure is the failure of the entire supply. Instrument-air failure is commonly considered to be a plant-wide failure unless conditions exist that allow the air supply to continue, such as when automatic makeup from an uninterrupted source or when multiple-compressor-source supplies are provided. Failure of the plant power to electronic or electrical instruments may also be considered, although credit can be given for sufficiently reliable backup power supplies (e.g. uninterruptible power supply).

To define the combined relieving loads under fire exposure, the probable maximum extent of a fire should be estimated. As a conservative approach, in the absence of any other governing factors, consideration of a fire-impact area is frequently limited to a ground area of 230 m^2 to 460 m^2 (2 500 ft^2 to 5 000 ft^2). A more detailed analysis can show a smaller fire-impact area. This detailed analysis includes consideration of the actual layout of facilities, the location of sources of combustibles, the provision of drainage and the effects of natural barriers.

Facilities that handle only flammable gases can be assumed to generate more localized fires than those produced when the release of flammable or combustible liquids results in a pool fire.

7.1.3 Establishing design load for the disposal system

The maximum potential load should be calculated for each common-mode event by adding up the individual system loads that would result during that contingency. For a common-mode event, it is necessary for the designer to determine the loads for each individual system including, as applicable, pressure-relief devices, emergency depressuring valves and/or other control valves (e.g. pressure-control vent valves). It is important to recognize that systems with pressure-control valves and/or depressuring valves, can maintain the pressure below the opening pressure of a pressure-relief device. In such cases, it is not necessary to include the load from the pressure-relief device in the flare load in addition to that from the pressure-control valves and/or depressuring valves. Note that in these cases the resulting disposal-system load from the pressure-control valve or the emergency depressuring valve can be larger than the calculated load for the pressure-relief device. For example, reboiler temperature pinch at full relieving pressure results in no relief load but the pressure-control vent valve opens at a pressure where the reboiler can still generate a vapour load to the disposal system.

The maximum flow through an emergency depressuring valve or pressure-control valve is limited by the maximum expected pressure upstream of the valve at the moment it is first opened. The designer shall determine this maximum expected pressure by reviewing the scenarios in which the emergency depressuring valve or pressure-control valve is opened. If the scenario involves vessels reaching pressure-relief-valve set pressure or full relieving pressure prior to opening the emergency depressuring valve or pressure-control valve, then these pressures should be used as the design basis. If the scenario is such that no abnormal pressure build-up is expected prior to opening the emergency depressuring valve or pressure-control valve, then the maximum normal operating pressure is likely be used as the design basis. It is important to recognize that reclosable pressure-relief devices might not have an additive load contribution to pressure-control valves connected to the same vessel or equipment.

If the capacity of a vapour-depressuring valve exceeds the normal vapour flow rate within the protected equipment or if the depressuring rate is additive to normal flows within the equipment, considerable liquid entrainment can occur. Therefore, disposal systems for depressuring valves should generally provide for liquid carryovers.

The disposal-system design basis is not necessarily based on the maximum mass load, because of the influence of fluid properties. For example, the flow that imposes the greatest head loss in flowing through the system might not be the highest mass flow. Thus, a flow of 12,6 kg/s (100 000 lb/h) of a vapour with a relative molecular mass of 19 at a temperature of 149 °C (300 °F) develops a greater head loss and is a greater “load” than a flow at 18,9 kg/s (150 000 lb/h) of a vapour with a relative molecular mass of 44 at a temperature of 38 °C (100 °F). Also, different scenarios may set the design basis for individual components such as laterals, knockout drums, flare stack, etc.

7.1.4 Refinement of the disposal system design load

7.1.4.1 General

There are several techniques that can be applied to establish a disposal-system design load that is less than the maximum as calculated in 7.1.3. The use of the techniques described in 7.1.4.2 and 7.1.4.3, particularly in combination, requires detailed analysis and can be very complex.

7.1.4.2 Dynamic system load modelling

Dynamic-system load modelling allows the user to predict the timing of individual system peak loads to determine the disposal system hydraulic performance. Various time intervals should be considered as a peak load for various parts of the disposal system can occur at different times. Dynamic-system load modelling differs from individual-system dynamic modelling (as described in 5.22) because the former considers the timing of multiple reliefs whereas the latter is focusing on determining only the peak load from one system without regard to the effect of timing on other relief loads. During a common mode event (fire or major utility failure) not all affected pressure systems reach full relieving conditions at the same time, and not all affected systems are able to sustain their loads for the same duration. Dynamic-system load modelling can require more sensitivity studies than individual-system dynamic modelling because the combined peak load for the disposal system might not necessarily occur during the peak load of an individual pressure system.

7.1.4.3 Load reduction credits

Load reduction credits include HIPS (see Annex E), operator intervention, basic process control, etc. As stated in 4.2, credit for some favourable instrumentation response to reduce disposal-system design loads may be taken. The decision to exclude a particular load due to the favourable response of instrument systems should consider the number and reliability of applicable instrument systems. Safety instrumented systems with high SIL values (see 3.66 and 3.67) are more likely to function than simple instrument trip systems or basic process control instrumentation. One approach is for the user to determine, based on instrument-system reliability, the percentage of these systems that would not function as designed. After doing so, the user then determines which instrument system is assumed to fail. Typically, the instrument systems that contribute the highest loads and/or back pressures are assumed to fail. When following this approach, the user should assess the potential for failure of multiple instrument systems affecting common relief headers to assure an adequate design.

Consideration may also be given to the capability for and response time available for operator intervention as a means of reducing system loads. When doing so, the user shall consider what other demands can be placed on the operator during the upset. Operator intervention credits can have already been taken in establishing the individual system relief load (see 5.4) and major upsets can require the operator to respond to many alarm conditions.

The basis for taking system-relief load-reduction credits should be evaluated carefully to assure an adequate design. One method of assessing the acceptability of system-relief load-reduction credits is to quantitatively assess the disposal system performance as a whole. This method considers the likelihood of the overpressure contingencies and the reliability of the safeguards that reduce or eliminate individual relief loads. This quantitative approach calculates the probabilistic disposal-system loads, probabilistic hydraulics and probabilistic equipment overpressures. The system performance is compared to the user's acceptance criteria.

7.2 System arrangement

7.2.1 General

Once the various combinations of loads have been defined for all pertinent contingencies and the corresponding allowable back pressures have been determined for all pressure-relief devices, selection of the disposal system can proceed. The factors influencing the choice of the disposal system are discussed in Clause 6.

In selecting the arrangement of the disposal system or systems, special attention should be given to situations where pressure-relief devices can discharge flashing liquids or where a combination of cold liquid and hot vapour discharge can result in vaporization of the liquid. Such situations can generate additional vapour loads beyond those that correspond to the relieving loads (see also 6.6.2.4 for special considerations in handling liquids that are capable of auto-refrigeration).

Table 12 can be used to determine where the required relieving rate and pressure-relief device rated capacity should be used in design of laterals and main header (if applicable).

Table 12 — Design basis for pressure-relief device laterals and disposal system headers

Device	Lateral/tail pipe	Main header (if applicable)
Pop-action, pilot-operated relief valve	Pressure-relief-valve rated capacity	Required relieving rate
Modulating-action pilot-operated relief valve	Required relieving rate [a]	Required relieving rate
Spring-operated relief valve	Pressure-relief-valve rated capacity [b]	Required relieving rate
Rupture disk (stand-alone)	Required relieving rate [c]	Required relieving rate
Buckling pin (stand-alone)	Required relieving rate [c]	Required relieving rate

[a] Consult the manufacturer, as some types of spring-operated pressure-relief valves can have some modulating capability. In these instances, the required relieving rate may be used.

[b] The mechanical and hydraulic design of the system should consider that the instantaneous flow rate upon opening can exceed the required relieving rate, particularly in cases where the relief device is oversized.

[c] The user is cautioned that if the required relieving rate is used for design of the lateral (see 3.41), any process changes that raise the required relieving rate can increase the back pressure above the acceptable limits.

7.2.2 Single-device disposal systems

Where only a single pressure-relief device or a single depressuring valve is connected to the disposal system, the outlet may also be to the atmosphere, to another system operating at lower pressure or to a local flare.

If the outlet is connected to a lower-pressure system, the allowable pressure drop in the disposal system should generally be based on the maximum allowable working pressure of the lower-pressure equipment. However, a reduced back pressure (e.g. normal operating pressure in the lower-pressure equipment) may be used if it can be shown that (a) none of the contingencies causing a relieving load also overpressure the lower-pressure equipment and (b) the load (the required relieving rate of the device) imposed by the higher-pressure-relief device does not result in overpressuring the lower-pressure equipment.

Each pressure-relief device that vents directly to the atmosphere should normally have an individual vent pipe sized for a relatively high exit velocity; however, the outlet piping should not be smaller than the pressure-relief-device outlet. The developed back pressure of this system should include all pressure losses, such as exit losses, friction losses and kinetic energy loss.

Pop-action, pilot-operated pressure-relief valves normally have the back pressure calculation based on the rated capacity of the valve. The design of the disposal system should be checked for adequacy under such conditions. Modulating pilot-operated-type pressure-relief valves generate loads that are equivalent to the required relieving rate for a particular contingency [101]. The initial design of the disposal system should, as a minimum, be based upon this required relieving rate.

Typically, the required relieving rate is used for the flare header, flare tip and knockout drum design with spring-loaded pressure-relief valves. However, there can be instances where a higher flow rate than required can be encountered that affects operation of the downstream equipment. For example, most spring-loaded pressure-relief valves discharge 50 % or more of their rated capacity at set pressure. Consequently, the initial flow rate can be greater than the required relieving rate. In this case, the rated capacity can be used as an upper-limit flow rate when designing downstream components such as scrubbers, thermal oxidizers and liquid-seal drums.

Some spring-loaded pressure-relief valves have limited modulating capabilities and can initially vent at rated capacity; consult the manufacturer to determine if the valve has modulating characteristics.

For rupture-disk or buckling-pin devices installed as a stand-alone device (i.e., not upstream of a pressure-relief valve), the required relieving rate is typically used to size the piping and the relief device. The design of downstream equipment, particularly scrubbers and thermal oxidizers, should consider the higher load that can be encountered based on the upstream pressure at which the relief device opens. The piping mechanical design should also consider this higher initial capacity.

7.2.3 Multiple-device disposal system

For disposal to a flare or to a remote atmospheric vent stack, combining the discharges from a number of pressure-relief devices or depressuring valves is usually economical. The specific arrangement of the headers and the routing of piping for the multiple-device system is normally a question of minimizing investment. This requires taking into consideration the system loads, the back-pressure limitations, the requirement for special materials and other design parameters discussed in 7.2.1 and 7.2.2.

In a multiple-device system that services a single unit, from the standpoint of economics, safety or other pertinent factors, it is frequently desirable to isolate certain pressure-relief or depressuring streams. This involves one or more of the following situations:

a) occurrence of corrosive materials;

b) significant differences in the pressure levels of the equipment connected to the system;

c) pressure relief or depressuring streams that can subject piping to abnormally high or low temperatures;

d) reactive materials (see 6.2.1).

In defining the header arrangements, consideration should be given to any requirements for separate shutdowns or separate maintenance on the protected equipment. It is usually not advisable to route pressure-relief-device headers from one operating area through another area where major maintenance shutdowns are performed separately. Furthermore, it is usually advisable to be able to isolate headers that serve separate process areas from the disposal system, rather than to be required to isolate individual pressure-relief devices within a common process area.

Multiple pressure-relief-device disposal systems that handle combustible vapours should not be used for venting air or steam during the start-up of process equipment. Any tie-ins of process vents to the multiple-device system should be accompanied by strict instructions against using such tie-ins for venting air to avoid flammable mixtures in a system.

Most multiple-device systems involve collecting pressure-relief-device discharges from various elevations. In general, laterals and headers should be arranged so that the outlet from each pressure-relief device is not a liquid trap. All collecting piping should be considered subject to the inflow of liquid and should avoid liquid traps. If it is not practical to arrange the piping so that laterals and headers drain to a remote knockout drum, a local knockout drum is usually required. It is normally not necessary for this local knockout drum to be sized for efficient vapour-liquid separation at the maximum flow rate, but only for collecting the probable maximum liquid carryover from any devices that can discharge liquids. The use of traps or other devices with operating mechanisms should be avoided.

If the liquids to be handled include water or oil with a relatively high pour point, a provision should be made to avoid solidification in the system. Likewise, the introduction of high-viscosity liquids can require protection against low ambient temperature, particularly on instrument-impulse lines.

7.3 Design of disposal system components

7.3.1 Piping

7.3.1.1 General

The design of disposal piping should comply with ISO 15649[2] or other piping codes specified by the owner.

NOTE For the purpose of this provision, ASME B31.3[21] is equivalent to ISO 15649.

Installation details and criteria pertinent to pressure-relief devices should conform to those specified in API RP 520-II or ISO 4126.

7.3.1.2 Design of relief device inlet piping

The inlet piping includes all components and fittings that comprise the flow passage between the entrance to the vessel nozzle and the face of the inlet flange of the pressure-relief valve or device. The first consideration is to provide full inlet area as a minimum and evaluate the reduction in relief-valve capacity caused by any rupture-disk devices, block valves or other components.

The inlet system should be self-draining and designed to prevent excessive pressure loss, which causes chattering with a consequent reduction of flow and damage to pipe joints and seating surfaces. See API RP 520-II and ISO 4126, which provide guidance on acceptable inlet-pressure losses. Pilot-operated valves can require consideration of the location of the pressure pickup to ensure proper sensing for stable operation of the main valve where inlet-pressure loss is excessive. Pressure drop through the burst rupture disk should be taken into account when calculating the inlet losses to the pressure-relief valve. The pressure-relief device should also be located as close to the source of pressure as is practicable; oversizing should be avoided.

API RP 520-II, ISO 4126 and EN 764-7 provide information on the design of inlet piping for pressure-relief devices; it is acknowledged that their approaches are different and the user shall specify which standard is to be used. Several sizing methods have been developed that minimize inlet-piping calculations in most cases and allow the designer to quickly identify marginal situations [102], [103].

In addition to flow considerations, the vessel nozzle and other inlet piping should be designed to withstand thermal loadings, reaction forces resulting from valve operation, vibration, dead weight and externally applied loadings.

The strength of the inlet piping is less than that of the valve because the inlet piping has a smaller section modulus. Any moments created by loads applied to the outlet flange and by the reactive force of the discharging fluid transmit bending stresses and rotational forces to the inlet piping. Design for reactive force is discussed in API RP 520-II and ISO 4126.

7.3.1.3 Design of relief device discharging piping

7.3.1.3.1 The basic criterion for sizing the discharge piping and the relief manifold is that the back pressure (which can exist or be developed at any point in the system) shall not reduce the relieving capacity of any of the pressure-relieving devices below the amount required to protect the corresponding vessels from overpressure. Thus, the effect of superimposed or built-up back pressure on the operating characteristics of the valves should be carefully examined. The discharge piping system should be designed so that the built-up back pressure caused by the flow through the valve under consideration does not reduce the capacity below that required of any pressure-relief valve that can be relieving simultaneously. Where conventional pressure-relief valves are used, the relief manifold system should be sized to limit the built-up back pressure to approximately 10 % of the set pressure of each pressure-relief valve that can be relieving concurrently. Additionally, the effect of superimposed back pressure from other valves upon the set pressure should be considered.

With balanced pressure-relief valves (bellows, piston or pilot-operated), higher manifold pressures can be used. The capacity of these balanced valves begins to decrease when the back pressure exceeds 30 % to 50 % of the set pressure due to subsonic flow and/or physical responses to the high back pressure. Refer to API RP 520-I or ISO 4126 and/or to manufacturer's curves for the effects of this back pressure. Additionally, the back pressure should not exceed the rating tabulated in API Std 526 or ISO 4126 as specified by the owner, which can be lower than the outlet flange rating.

When discharge manifolds and relief headers are sized, the relief contingency that produces the greatest back pressure should be identified. Any single relief contingency may involve several pressure-relief devices. Typical relief contingencies that may be considered include cooling water failure, power failure and instrument air failure.

The design of pipe anchors and supports on discharge manifolds can require special consideration. Sudden changes in flow rate and temperature can produce large reaction forces; if liquids are present in the relief system, the momentum forces can be significant. API RP 520-II discusses this subject in more detail.

In addition to the back-pressure criterion, the determination of the flow rate to be considered forms the basis for discharge-line sizing. In general, laterals and tailpipes from individual pressure-relief valves are sized based on the rated capacity (see Table 12). The lateral or tailpipe from a modulating pilot-operated pressure-relief valve is sized based upon the required relieving rate for a particular contingency. If process conditions change and the required relieving rate is higher than the initial design, the lateral or tailpipe from the modulating pilot-operated pressure relief should be checked for adequacy.

If the user has established a velocity criterion for tailpipes, the maximum velocity in a tail pipe should be calculated with the single source (the relief or depressurization device) as the only source discharging into the disposal system. Due to pressure drop in the tail pipe, the maximum velocity should be calculated at the end of each pipe diameter (if the diameter varies).

Common header systems and manifolds in multiple-device installations are generally sized based on the worst-case cumulative required capacities (instead of rated capacities) of all devices that can reasonably be expected to discharge simultaneously in a single overpressure event (in other words, for certain scenarios, it can be appropriate to assume some level of favourable instrument and/or operation response). See API RP 520-I for information regarding rated capacities. Causes of overpressure events are discussed in Clause 4; required relieving capacities for various events are discussed in Clause 5. If process conditions change and dictate a higher required cumulative capacity from the initial design, the common header system should be re-evaluated.

Simple rules cannot be expected to cover all installations. Good engineering judgment should be applied to select the flow basis most appropriate to each case.

In designing vapour depressuring systems, precise pressure-drop calculations are usually not necessary. The only limits on built-up back pressure, in addition to those mentioned above, are as follows.

a) The ratings of fittings exposed to back pressure should not be exceeded.

b) Any source that can reasonably be depressurized concurrently should be capable of entering the header when its depressuring valve is opened.

c) Back-flow from the header into any connected process should be avoided.

When the maximum vapour-relieving requirement has been established and the maximum allowable header back pressure has been defined (as determined by the type of valves in the system and the applicable design requirements), the selection of line size is then reduced to fluid-flow calculations. Several methods can be used to calculate the size of discharge piping when the flow conditions are known. These range from treating the flow as isothermal, with appropriate allowances for kinetic energy effects, to the more rigorous solutions afforded by the adiabatic approach. A number of methods listed in the Bibliography under the heading “Piping” permit the user to select the method best suited to his needs. In the absence of any preference, the following methods are recommended.

7.3.1.3.2 Several commercial relief-system network-flow simulation programs are available to the designer. The programs allow the user to model a complex flare system, including pressure-relief-device inlet piping, pressure-relief devices, pressure vents, outlet tail piping, flare laterals, main flare header, knockout drum, seal drum, flare stack and flare tip. Hydraulic capacity of modern flare tips is not directly discernible from the mounting flange size alone. Consult the flare vendor for flare-tip capacity. See API Std 537 for more details. Multiple-relief scenarios can be studied by alternating the devices that are allowed to flow. The programs typically use a steady-state equation of state to predict the properties of the fluid in the relief system. For gases or vapours, both isothermal and adiabatic flow models are usually available. Additionally, there are options to consider heat transfer (i.e., conditions between adiabatic and isothermal). Discontinuities in pressure, for gases, vapours and two-phase fluid, due to critical-flow conditions at restrictions such as the pressure-relief device outlet and the tail pipe outlets shall be taken into account by the program. For two-phase flow conditions, typically a two-phase pressure-drop method, such as Beggs and Brill [104] or the homogeneous equilibrium method (see 7.3.1.3.5) may be used. The Beggs and Brill method should have an adjustment in the acceleration term where necessary, due to the high velocity frequently found in flare headers. Critical flow conditions are typically handled by assuming homogeneous flow and applying basic thermodynamic relationships.

For less complex systems handling gases or vapours, spreadsheet models or other user-developed models can be employed to size and verify the relief system. A typical model of this type uses mass flow, density and heat capacity relationships to calculate the flow, temperature and physical properties of the relief fluid at nodes in the system. The pressure at various points is calculated using a compressible flow model (either isothermal or adiabatic). It is important that the model also check for pressure discontinuities due to critical flow in the system. Various relief scenarios can be evaluated by changing which inputs are active.

7.3.1.3.3 A number of methods listed in the Bibliography under the heading “Piping” permit the user to select a manual method to calculate the back pressure for gases and vapour in flare systems. In the absence of any preference, the methods described in the remainder of this subclause are suggested.

Vapour flow in relief-discharge piping is characterized by rapid changes in density and velocity; consequently, the flow should be rated as compressible. Several methods for calculating the size of discharge piping have been developed using isothermal or adiabatic flow equations. Actual flow conditions in relief systems are normally somewhere between isothermal and adiabatic conditions. For most cases, the slightly more conservative isothermal equations are recommended; however, the adiabatic flow equations can be preferable for some less-common applications (e.g. cryogenic conditions).

The sizing of relief-discharge piping can usually be simplified by starting at the system outlet, where the pressure is known, and working back through the system to verify acceptable back pressure at each pressure-relief device. Calculations are performed in a stepwise manner for each pipe segment of constant diameter. The isothermal flow equation based on inlet Mach number[105] is as given in Equation (25):

$$\frac{f \cdot l}{d} = \frac{1}{Ma_1^2}\left[1-\left(\frac{p_2}{p_1}\right)^2\right] - \ln\left(\frac{p_1}{p_2}\right)^2 \tag{25}$$

The equation can be transposed to Equation (26) for outlet Mach number:

$$\frac{f \cdot l}{d} = \frac{1}{Ma_2^2}\left[\left(\frac{p_1}{p_2}\right)^2\right]\left[1-\left(\frac{p_2}{p_1}\right)^2\right] - \ln\left(\frac{p_1}{p_2}\right)^2 \tag{26}$$

where

f is the Moody friction factor, dimensionless;

l is the equivalent length of pipe, expressed in metres (feet);

d is the pipe inside diameter, expressed in metres (feet);

Ma_1 is the Mach number at pipe inlet;

Ma_2 is the Mach number at pipe outlet;

p_1 is the pipe inlet absolute pressure, expressed in kilopascals (pounds per square inch);

p_2 is the pipe outlet absolute pressure, expressed in kilopascals (pounds per square inch).

The isothermal outlet Mach number is given by Equations (27) and (28):

In SI units:

$$Ma_2 = 3{,}23 \times 10^{-5}\left(\frac{q_\mathrm{m}}{p_2 \cdot d^2}\right)\left(\frac{Z \cdot T}{M}\right)^{0{,}5} \tag{27}$$

In USC units:

$$Ma_2 = 1{,}702 \times 10^{-5}\left(\frac{q_\mathrm{m}}{p_2 \cdot d^2}\right)\left(\frac{Z \cdot T}{M}\right)^{0{,}5} \tag{28}$$

where

q_m is the gas mass flow rate, expressed in kilograms per hour (pounds per hour);

Z is the gas compressibility factor;

T is the absolute temperature, expressed in kelvin (degrees Rankin);

M is the gas relative molecular mass.

Both graphical and computerized methods have been developed for solving Equations (25) and (26) and calculating pipe inlet pressure [105], [106]. Figure 14 is a typical graphical representation of Equation (25). The figure may be used to calculate the inlet pressure, p_1, for a line segment of constant diameter where the outlet pressure is known. If the relief system is to be operated at high pressure, the flow may be sonic in some parts of the system. In those cases, a check should be made to see if the flow is critical. The critical pressure at the pipe outlet can be determined by setting $Ma_2 = 1{,}0$ (sonic flow) in Equations (27) and (28) resulting in Equations (29) and (30):

In SI units:

$$p_{\text{crit}} = 3{,}23 \times 10^{-5} \left(\frac{q_{\text{m}}}{d^2} \right) \left(\frac{Z \cdot T}{M} \right)^{0{,}5} \tag{29}$$

In USC units:

$$p_{\text{crit}} = 1{,}702 \times 10^{-5} \left(\frac{q_{\text{m}}}{d^2} \right) \left(\frac{Z \cdot T}{M} \right)^{0{,}5} \tag{30}$$

where p_{crit} is the critical absolute pressure, expressed in kilopascals (pounds per square inch).

If the critical pressure is less than the pipe outlet pressure, the flow is subsonic. If the critical pressure is greater than the pipe outlet pressure, the flow is sonic and $Ma_2 = 1$. Therefore, the pipe inlet pressure, p_1, is calculated from Equation (25) with p_2 equal to the critical pressure.

7.3.1.3.4 A rapid solution for sizing depressuring lines is offered in the remainder of this subclause, using the method developed by Lapple [107]. This method employs a theoretical critical mass flow based on an ideal nozzle and adiabatic flow conditions and assumes a known upstream low velocity source pressure. The critical mass flux for isothermal flow conditions (i.e. where vapour $k = C_p/C_v = 1{,}00$) can be determined using Equations (31) and (32):

In SI units:

$$G_{\text{Ci}} = 6{,}7 p_1 \left(\frac{M}{Z \cdot T_1} \right)^{0{,}5} \tag{31}$$

In USC units:

$$G_{\text{Ci}} = 12{,}6 p_1 \left(\frac{M}{Z \cdot T_1} \right)^{0{,}5} \tag{32}$$

where

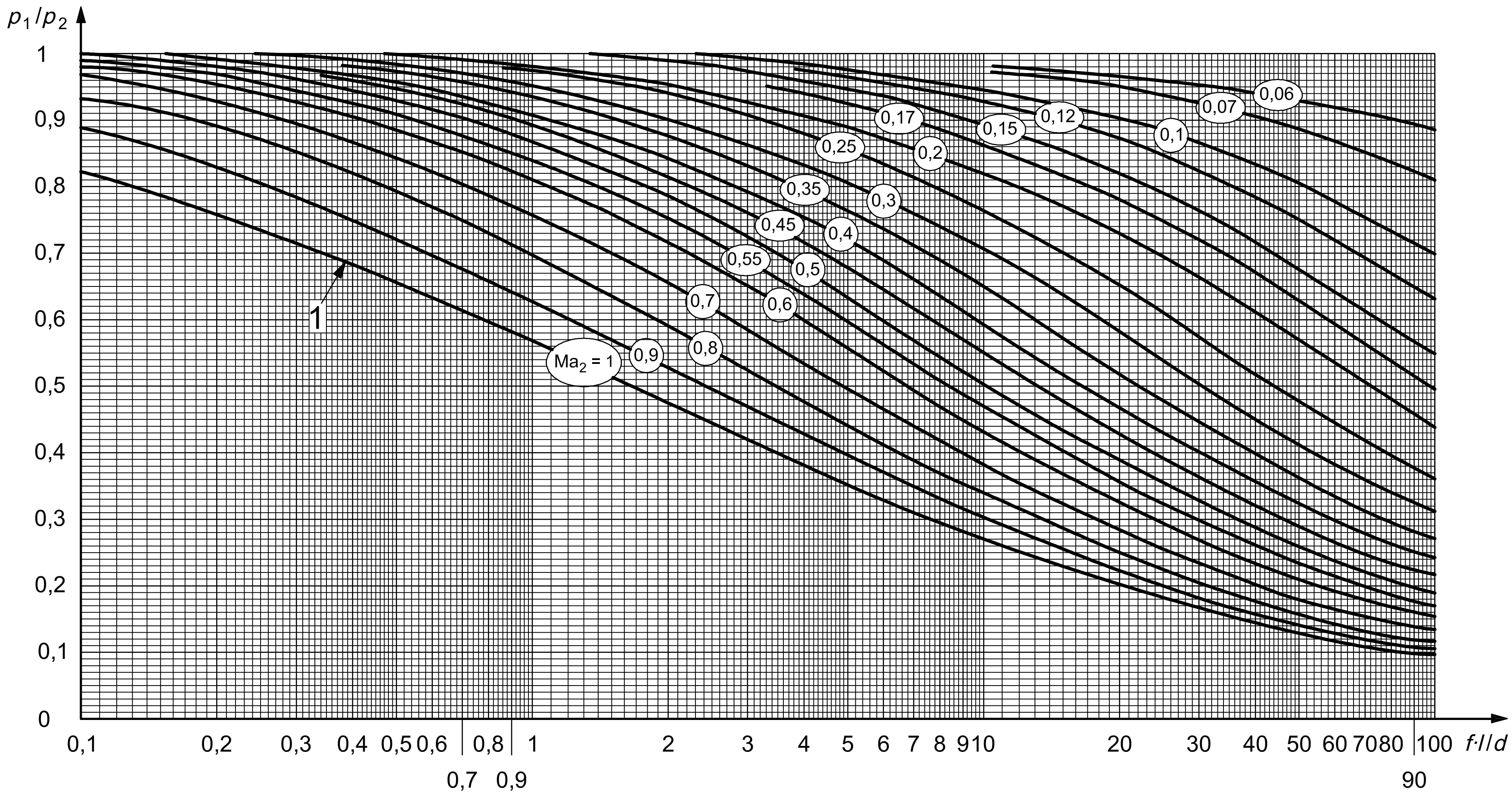
G_{Ci} is the critical mass flux, expressed in kilograms perf second-square metre (pounds per second-square foot);

p_1 is the absolute pressure at the upstream low velocity source (see Figure 15), expressed in kilopascals (pounds per square inch);

M is the relative molecular mass of the vapour;

T_1 is the upstream temperature, expressed in kelvin (degrees Rankin);

Z is the compressibility factor.



Key

1 critical flow line

NOTE This figure is reprinted with permission from the *Oil and Gas Journal*, Nov. 20, 1978, p. 166.

Figure 14 — Isothermal flow chart

The compressibility factor should be taken at flow conditions and, thus, changes as the fluid moves down the line with a resulting pressure drop. A stepwise calculation may be employed to allow for this variation. An accurate solution using this method is tedious, but sufficiently accurate results can usually be obtained by performing the calculation over relatively large increments of pipe lengths, using an average compressibility factor over those lengths. Regardless of which equation is used, actual mass flux (G) is a function of critical mass flux (G_{Ci}), frictional resistance (N), and the ratio of downstream to upstream pressure. These relationships are plotted in Figure 15. (Similar charts for adiabatic cases with ratios of specific heats of 1,4 and 1,8 have been developed by Lapple [107].) In the area below the diagonal line in Figure 15, the ratio G/G_{Ci} remains constant, which indicates that sonic flow has been established. The total frictional resistance for use with the chart is expressed by Equation (33):

$$N = \frac{f \cdot l}{d} + \sum K_i \qquad (33)$$

where

N is the pipe frictional resistance factor, dimensionless;

f is the Moody friction factor, dimensionless;

l is the actual length of the pipe, expressed in metres (feet);

d is the diameter of the pipe, expressed in metres (feet);

K_i are the resistance coefficients of fittings (see Tables 13 and 14), dimensionless.

If a Fanning friction factor is used, Equation (33) reduces to the following expression:

$$N = 4\frac{f \cdot l}{d}$$

These methods assume that there are no enlargements or contractions in the piping and no variation in the Mach number that results from a change in area. Coulter [108] provides a more comprehensive treatment of ideal gas flow through sudden enlargements and contractions.

Another method of calculating pressure drops for ideal gases at high velocities is the use of Fanno lines. Fanno lines are the loci of enthalpy/entropy conditions that result from adiabatic flow with friction in a pipe of constant cross-section. Fanno lines extend into both supersonic and subsonic flow zones. For relief-disposal systems, only the subsonic flow is of interest. The use of Fanno lines permits the calculation of pressure drops for ideal gases under adiabatic or isothermal flow conditions, with the total piping resistance as a parameter [109]. In general, the velocity in gas-discharge piping cannot exceed the sonic or critical velocity limit. (This limit is shown on Lapple's charts [107] or on Fanno lines.)

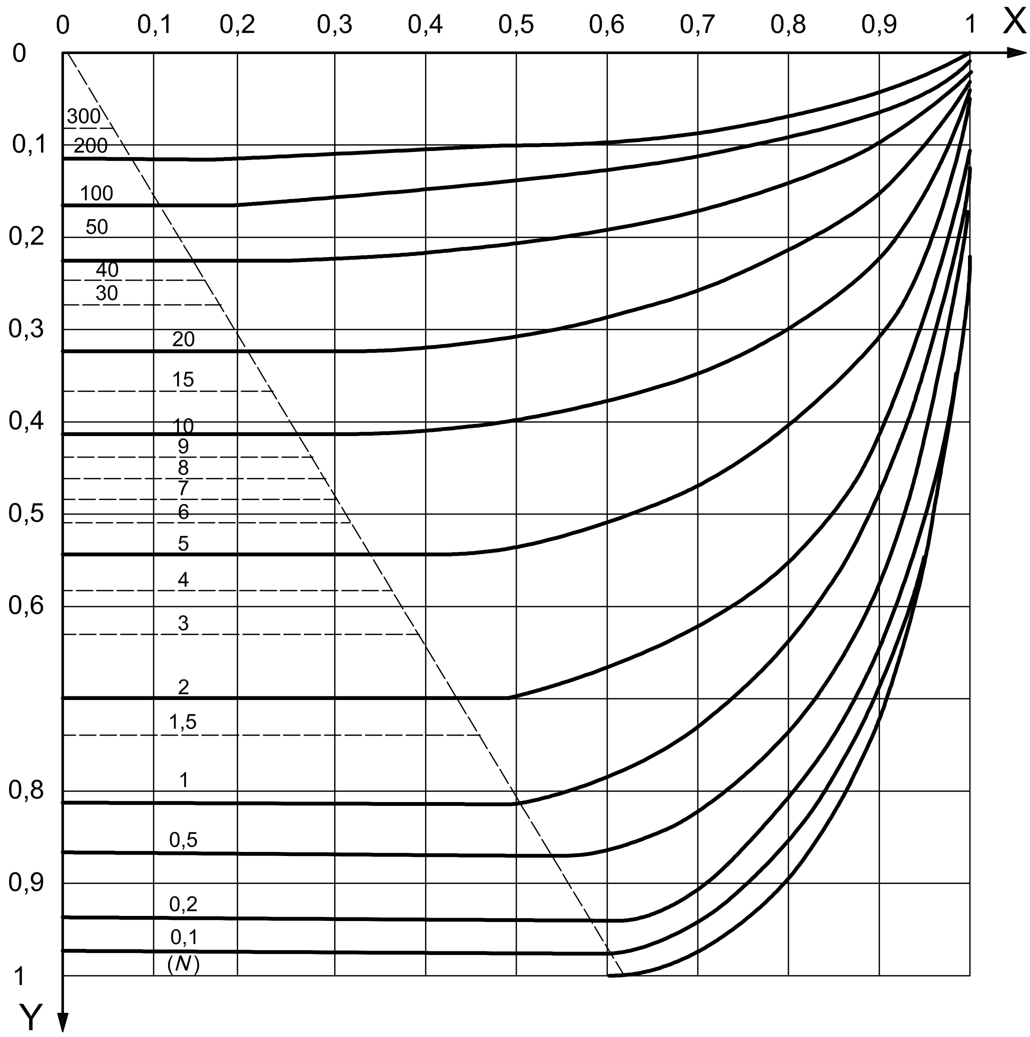
In most disposal systems, the gases being handled are not ideal. For gases, deviations from the ideal are expressed as compressibility factors, which, in turn, are normally correlated with reduced pressure and reduced temperature. For hydrocarbon gases, the compressibility factor is less than 1,0 if the reduced temperature does not exceed 2,0 and the reduced pressure does not exceed about 6,0. Since most pressure-relief-device disposal systems fall within these limits, the compressibility of the gases is usually less than 1,0. As long as compressibility is less than 1,0, the pressure drop calculated for an ideal gas is larger than that calculated for the same gas incorporating the compressibility factor.

For most applications, the pressure drops that are calculated assuming ideal gases under isothermal flow conditions exceed those calculated by more rigorous procedures. In any design of a disposal system, the sizing of piping based on ideal gas flows under isothermal conditions is normally adequate. However, for very high-pressure or high- or low-temperature situations, the possible effects of deviations from ideality should be checked.

If a rigorous calculation of pressure drop, including the effect of non-ideal behaviour is necessary or desirable, an incremental or stepwise approach is usually required. It should also be noted that for ideal gases, the specific heat ratio is equal to the isentropic expansion exponent and is independent of pressure. For non-ideal gases this is, at best, an approximation. Rigorous pressure-drop calculations should be based on the use of the real-gas isentropic-expansion exponent and should consider its pressure dependency.

In any calculation method, the total frictional resistance should include the length of piping and the equivalent length of all fittings, valves, expansion or contraction losses and any other flow resistances. The frictional resistance of fittings and some other items in the piping system can also be expressed in terms of K-factors. Tables 13 and 14 show typical K-factors for pipe fittings and for reducers (enlargements and contractions).

The friction factor, f, enters into all calculations of pressure drop. At high gas-flow velocities, which usually prevail in the design of disposal systems, the friction factor approaches a constant number that depends only on pipe size and internal roughness.



Key

X p_3/p_1 (p_2/p_1 only above diagonal dashed line)

where

p_3 is the pressure in the reservoir into which the pipe discharges [101 kPa (14,7 psia) with atmosphere discharge];

p_1 is the pressure at upstream low velocity source, expressed in kPa (psia);

p_2 is the pressure in the pipe at the exit or any point a distance, L, downstream from the source, expressed in kPa (psia).

Y G/G_{ci}

NOTE 1 Equations (31) and (32) are based on adiabatic flow and a vapour k-value ($= C_p/C_v$) approaching 1,00. For adiabatic flow and with a vapour of $k = 1,40$, the critical mass flux is 12,9 % higher than that calculated with Equations (31) and (32).

NOTE 2 The area below the diagonal dashed line represents sonic flow.

Figure 15 — Adiabatic flow of $k = 1,00$ (i.e. isothermal flow) compressible fluids through pipes at high pressure drops

Figure 15 is used as follows:

a) Calculate N (number of velocity heads) using Equation (33);

b) Calculate p_3/p_1 or p_2/p_1;

c) Calculate G_{ci} using Equation (31) or (32);

d) Knowing N and either p_3/p_1 or p_2/p_1, obtain G/G_{ci} from Figure 15;

e) Calculate G, in kilograms per square metre (pounds per second per square foot);

f) Calculate W [the actual flow, in kilograms per second (pounds per second)] as follows:

$$W = G \times A_p$$

where A_p is the cross-sectional area of pipe, expressed in square metres (square feet).

Table 13 — Typical K-factors for pipe fittings

Fitting	K	Fitting	K
Globe valve, open	9,7	90° double-mitre elbow	0,59
Typical depressuring valve, open	8,5	Threaded tee through run	0,50
Angle valve, open	4,6	Fabricated tee through run	0,50
Swing check valve, open	2,3	Lateral through run	0,50
180° close-threaded return	1,95	90° triple-mitre elbow	0,46
Threaded or fabricated tee through branch	1,72	45° single-mitre elbow	0,46
90° single-mitre elbow	1,72	180° welded return	0,43
Welded tee through branch	1,37	45° threaded elbow	0,43
90° standard-threaded elbow	0,93	Welded tee through run	0,38
60° single-mitre elbow	0,93	90° welded elbow	0,32
45° lateral through branch	0,76	45° welded elbow	0,21
90° long-sweep elbow	0,59	Gate valve, open	0,21
Rupture disk, subcritical flow	1,5 [a]		

NOTE 1 Except for rupture-disk data, this table is taken from the Tube-Turn Catalogue and Engineering Data Book No. 211 [143].

NOTE 2 K can vary with nominal pipe diameter. The values above are typical only.

[a] $K = 1,5$ has been used successfully for design purposes. Other K values have also been reported [110]. Consult a rupture-disk manufacturer for specific data, if required.

Table 14 — Typical K-factors for reducers (contraction or enlargement)

Contraction or enlargement	K-factors for various values of d/d' [a]				
	0	0,2	0,4	0,6	0,8
Contraction (ANSI)	—	—	0,21	0,135	0,039
Contraction (sudden)	0,5	0,46	0,38	0,29	0,12
Enlargement (ANSI)	—	—	0,90	0,50	0,11
Enlargement (sudden)	1,0	0,95	0,74	0,41	0,11

[a] d is the inside diameter of the small end of the reducer; d' is the inside diameter of the large end of the reducer.

For preliminary studies, it is often necessary to assume K-factors or an equivalent length of fittings, expansion loops, and the like. Based on actual layouts, these elements can add equivalent length equal to 100 % or more of the geographical length of the pipe.

Where gas flow velocity in long runs of piping approaches the critical flow limit, it is often economical to increase the pipe size in steps or progressively along the run. In general, a calculation of pressure drop is required for each section of uniform size. The piping directly connected to a pressure-relief device should not be smaller than the size of the outlet flange.

7.3.1.3.5 If the system includes mixed-phase fluids (vapour and flashing or non-flashing liquid), the line sizing is more complex. Commercial relief-system network-flow simulation programs (see 7.3.1.3.2) generally use an equation-of-state model to predict vapour/liquid equilibrium and then apply a two-phase pressure-drop equation, with an adjustment in the acceleration term due to the high velocity typical in flare lines, to determine line pressure loss.

There are a number of manual two-phase flashing-flow pressure-drop correlations available (see Bibliography items under headings “Flashing flow in pipes” and “Flashing flow in valves”). One method is based on the homogeneous equilibrium flow assumption, i.e. equal velocity (no-slip) and equal temperature in both liquid and vapour phases [111]. For the case of a single-diameter, horizontal line, the compressible flow relationship given in Equation (34) can be used to determine pressure drop in multi-phase flow systems:

$$\frac{C_1 \cdot f \cdot l}{d} = \frac{C_2 \cdot p_R \cdot \rho_R}{G^2} \left\{ \frac{\eta_1 - \eta_2}{1-\omega} - \frac{\omega}{(1-\omega)^2} \ln\left[\frac{(1-\omega)\eta_1 + \omega}{(1-\omega)\eta_2 + \omega} \right] \right\} + \ln\left[\frac{(1-\omega)\eta_1 + \omega}{(1-\omega)\eta_2 + \omega} \left(\frac{\eta_2}{\eta_1} \right) \right] \tag{34}$$

where

l is the total equivalent length of pipe having diameter d (including fittings), expressed in meters (feet);

d is the inside pipe diameter, expressed in millimetres (inches);

f is the Fanning friction factor, assumed constant over the length of pipe;

p_R is the reference condition absolute pressure, expressed in kilopascals (pounds per square inch);

ρ_R is the reference condition density, expressed in kilograms per cubic metre (pounds per cubic foot);

G is the mass flux in the pipe, expressed in kilograms per hour-square millimetre (pounds per hour-square inch);

For pressure-relief valve laterals, use G as the relief-valve capacity divided by the pipe cross-sectional area. For headers, use G as the required relief load divided by the pipe cross-sectional area.

$\eta_1 = p_1/p_R$, dimensionless;

$\eta_2 = p_2/p_R$, dimensionless;

p_1 is the pipe-inlet absolute pressure, expressed in kilopascals (pounds per square inch);

p_2 is the pipe-exit absolute pressure, expressed in kilopascals (pounds per square inch);

ω is the correlating parameter referenced to p_R, ρ_R [see Equation (36)];

C_1 is a constant, equal to 2 000 in SI units ($C_1 = 24$ in USC units);

C_2 is a constant, equal to 0,012 96 in SI units ($2{,}898 \times 10^6$ in USC units).

As in the case of gases, the pipe-outlet pressure, p_2, depends on whether or not the flow at the end of the pipe is choked. The pipe-outlet pressure, p_2, of a constant-diameter pipe is the higher of the pressure at the exit of the pipe and the critical choking pressure given in Equation (35):

$$p_c = C_3 \cdot G \sqrt{\frac{\omega \cdot p_R}{\rho_R}} \tag{35}$$

where

p_c is the critical choking absolute pressure, expressed in kilopascals (pounds per square inch);

C_3 is a constant, equal to 8,784 in SI units (5,8742 × 10^{-4} in USC units);

and the definition of the other symbols is the same as for Equation (34).

If the pressure at the pipe exit (e.g. atmospheric pressure or other known pressure) is less than p_c, then the flow is choked. In this case, replace p_2 with p_c in the η_2 term used in Equation (34). Otherwise, the flow is not choked so the pipe exit pressure should be used as p_2 in Equation (34).

The following is a procedure to select the reference conditions for calculating ω for use in Equations (34) and (35):

Step 1) Perform an isenthalpic flash starting from relieving conditions to the maximum expected back pressure (p_B). In many cases involving multi-phase relief, a balanced bellows or pilot-operated relief valve will be required so for a first try let $p_B \sim$ 30 % to 50 % of the pressure-relief valve set pressure. Let this pressure be the new reference pressure, p_R, and determine the density of the multi-phase mixture. This density is the new reference density, ρ_R.

Step 2) Perform an isenthalpic flash from relieving conditions to 50 % of the reference pressure, p_R, from Step 1 or atmospheric pressure, whichever is greater. Assign this pressure as p, and the multi-phase mixture density at this pressure as ρ.

Step 3) Then

$$\omega = \frac{(\rho_R / \rho) - 1}{(p_R / p) - 1} = \frac{(v / v_R) - 1}{(p_R / p) - 1} \tag{36}$$

where

v is the specific volume, expressed in cubic metres per kilogram (cubic feet per pound);

v_R is the reference condition specific volume, expressed in cubic metres per kilogram (cubic feet per pound).

NOTE The value of ω used in Equations (34) and (35) is NOT the same as that used when sizing the pressure-relief valve in accordance with API RP 520-I:2000, Appendix D.

Step 4) If there is a large pressure drop, repeat steps 2 and 3 to obtain additional ω-values. Use the appropriate ω, p_R and ρ_R values that most closely correspond to the calculated pressures in each pipe segment.

Equations (34), (35) and (36) are used to calculate either the upstream pressure (i.e., p_1) or the maximum equivalent length of pipe allowed for the specific relief device. If using the equations to determine the upstream pressure, then

a) determine ω using Equation (36) and the guidance above;

b) calculate G and determine the critical pressure, p_c using Equation (35);

c) if p_c exceeds the outlet pressure p_2, then set $p_2 = p_c$; otherwise use p_2 directly in Equation (34).

Then, $\eta_2 = p_2/p_R$

d) calculate the Fanning friction factor;

e) solve Equation (34) by trial and error for η_1 and then $p_1 = (\eta_1 \cdot p_R)$;

f) determine if the selected relief device is appropriate for the calculated p_1.

If using the equations to determine the maximum allowed equivalent pipe length for a specific type of pressure-relief valve, then

a) determine the maximum allowed back pressure for the specific pressure-relief valve type (e.g. 10 % of set pressure for conventional pressure-relief valves, 30 % of set pressure for most balanced bellows pressure-relief valves (without derating), 50 % of set pressure for most pilot-operated pressure-relief valves (without derating). Set this back pressure equal to p_1;

b) determine ω, using Equation (36) and the guidance above;

c) calculate G and determine the critical pressure, p_c using Equation (35);

d) if p_c exceeds the outlet pressure p_2, then set $p_2 = p_c$; otherwise use p_2 directly in Equation (34).

Then, $\eta_2 = p_2/p_R$

e) calculate the Fanning friction factor;

f) solve Equation (34) directly for l.

7.3.1.3.6 The mechanical design of the disposal system warrants the same attention as that given to the design of piping systems that handle process fluids. The problems encountered in the design of discharge piping from pressure-relief device or depressuring valves are frequently more complex than those encountered in the design of a process system, since discharge piping can be subject to a greater range of temperature, pressure and shock caused by the wide range of operating conditions. In addition, the disposal system can, at one time or another, contain any material handled in the process system.

The major stresses to which the discharge piping of a relieving system is subject are results of thermal expansion or contraction from the entry of cold or hot materials and thrust developed by the discharge fluid. In relieving systems that serve facilities within the scope of this International Standard, temperatures can range from well below zero to several hundred degrees. Designing for flexibility is more complicated than it is for process piping systems because thrust as well as thermal expansion shall be controlled.

Most situations make it possible to maintain stress levels in relieving systems within allowable limits over the full temperature range by providing guides, anchors and appropriate piping configurations.

Special attention to stresses is recommended if piping constructed of carbon steel can be cooled below its transition temperature. Cooling can be caused by the entry of cold materials or by auto-refrigeration, which occurs when the pressure on low-boiling liquids is reduced. Reference should be made to ISO 15649 for material specifications, allowable stresses and impact test requirements for carbon-steel piping materials that can be used for temperatures as low as – 46 °C (– 50 °F). Stress relieving of welded piping systems has proven beneficial as a supplementary precaution in reducing the risk of brittle fracture of carbon-steel piping that can operate below its transition temperature. If temperatures below – 46 °C (– 50 °F) are possible, the usual practice is to construct relief lines of materials that exhibit ductile behaviour at the minimum anticipated operating temperature.

The design of discharge piping requires careful analysis of the possible imposition of both thermal and mechanical stresses on the associated pressure-relief devices. The stresses set up in the pressure-relief devices can cause malfunction or leakage of the devices (see API RP 520-II). Forces on the device can be controlled by proper anchors, supports and provisions for flexibility of discharge piping.

Discharge piping that is supported by the outlet of the pressure-relief device instead of being supported separately induces stresses in associated pressure-relief devices and inlet piping. Forced alignment of the discharge piping imposes similar stresses. Discharge piping, including short tail pipes, should be examined, supported and carefully aligned as requirements dictate. Strains sufficient to cause mechanical failure usually occur first at the inlet piping; however, moments at much lower levels can cause serious malfunction and leakage of the pressure-relief device. Stresses can also be imposed on the disposal piping as a result of reaction forces created when the pressure-relief devices are discharging. Provisions should be made for anchoring or restraining disposal lines related to these devices where analysis indicates that this is necessary. A formula for computing reactive loads due to the operation of pressure-relief devices is given in API RP 520-II or ISO 4126-9.

Shock loading should also be considered in relief lines. Shock loading can result either from the sudden release of a compressible fluid into a multidirectional piping system or from the impact action of liquid slugs at points of change in direction. Reaction forces can occur at each change of direction in the piping.

7.3.1.3.7 The design of appropriate and adequate anchors, guides and supports for a pressure-relieving discharge piping system is complex. There are several methods of calculating piping flexibility; reference should be made to ISO 15649 for a background discussion. Once the range of relieving conditions to be handled is established, the problems are no different from those for most other piping systems, other than also having to consider thrust forces.

Experience has shown that it is necessary to carefully consider answers to the following questions to permit the design of a satisfactory system of anchors, guides, and supports.

a) What are the probable combinations of relieving conditions that the manifold needs to handle? What sort of temperature ranges do these conditions impose, considering changes in the ambient temperature? What are the probable inlet conditions, in terms of thermal movement, when these reliefs occur?

b) What are the probable magnitude and sources of any liquid slugs?

c) Are there any valves that can release large volumes of high-pressure gas and produce shock loads? If so, where are they located?

In general, it is preferable to select anchor points so that header movements and the resultant forces and moments are not imposed on the bodies or the discharge piping of pressure-relief valves. If pressure-relief devices discharge to the atmosphere, the tailpipe configuration should be checked for discharge reaction forces to ensure that it will not be overstressed.

7.3.1.3.8 Disposal system piping should be self-draining toward the discharge end. Pocketing of discharge lines should be avoided. If pressure-relief devices handle viscous materials or materials that can solidify as they cool to ambient temperature, the discharge line should be heat-traced. A small drain pot or drip leg can be necessary at low points in lines that cannot be sloped continuously to the knockout or blowdown drum. The use of liquid drain traps or other devices with operating mechanisms should be avoided.

7.3.1.3.9 Many design details and features merit particular emphasis with respect to relieving systems. The following points are not to be taken as definitive or restrictive.

a) The laterals from individual relieving devices should normally enter a header from above. This tends to keep any liquids that flows or develops in the header out of the laterals to each valve.

b) Laterals that lead from individual valves located at an elevation above the header should drain to the header. Locating a pressure-relief device below the header elevation in closed systems should be avoided wherever possible. Laterals from individual devices that must be located below the header should be arranged to rise continuously to the top of the header entry point; however, means should be provided to prevent liquid accumulation on the discharge side of these valves.

c) A slope of 21 mm in 10 m (¼ in in 10 ft) is suggested for all laterals and headers, taking into account piping deflections between supports.

d) Where individual devices are vented to the atmosphere, an adequate drain hole [a nominal pipe size of DN 15 (NPS ½) is usually considered suitable] should be provided at the low point to ensure that no liquid collects downstream of the device. The vapour flow that occurs through this hole during venting is generally not considered significant, but each case should be checked to see if the drain connection should be piped to a safe location. Vapours escaping from the drain hole should not be allowed to impinge against the vessel shell, since accidental ignition of such vent streams can seriously weaken the shell.

e) The use of angle entry, e.g. an entry at 0,79 radian (45°) or even 0,52 radian (30°) to the header axis, for laterals is much more common in relieving systems than in most process-piping systems. The two main reasons for this approach are (1) lower pressure drop (including velocity head losses), and (2) reduced reaction forces. Since laterals in relieving systems can often be sized at velocities approaching sonic, pressure losses or recoveries caused by velocity change can become a significant factor in system analysis. These resultant density changes can produce large reaction forces.

f) The use of block valves to section the header system for maintenance or safety should be considered. Such block valves should be provided with locking or sealing devices. Where block valves cannot be justified, the provision for blinding should be studied. In locating sectioning block valves or blinds, extreme caution should be exercised in their use to ensure that equipment which is operating is not isolated from its relieving system. If block valves are used in the header system, they should be mounted so that they cannot fail in the closed position (an example would be a gate falling into its closed position). Procedures should address the possibility of exposing workers to flare gas from a relief scenario that occurs while they are blinding the flare system if there are no isolation valves. In some cases, the header can be operating under a vacuum (stack draft), in which case air can be drawn into the header resulting in the potential formation of explosive-combustible mixtures.

7.3.2 Drums and seals

7.3.2.1 Knockout drums

7.3.2.1.1 The economics of drum design can influence the choice between a horizontal and a vertical drum. If a large liquid storage capacity is desired and the vapour flow is high, a horizontal drum is often more economical. Also, the pressure drop across horizontal drums is generally the lowest of all the designs. Vertical knockout drums are typically used if the liquid load is low or limited plot space is available. They are well suited for incorporating into the base of the flare stack.

Although horizontal and vertical knockout drums are available in many configurations, the differences are mainly in how the path of the vapour is directed. The various configurations include the following:

a) horizontal drum with the vapour entering one end of the vessel and exiting at the top of the opposite end (no internal baffling);

b) vertical drum with the vapour inlet nozzle entering the vessel radially and the outlet nozzle at the top of the vessel’s vertical axis. The inlet stream should be baffled to direct the flow downward;

c) vertical vessel with a tangential nozzle. Vertical centrifugal separators differ from vertical settling drums in that the flow enters tangentially and spins around a centre tube, which extends below the liquid inlet nozzle. The gas and liquid flow radially downward through the annulus causing liquid droplets to coalesce along the walls and collect in the bottom of the drum. The vapour changes direction once below the centre tube and flows upward to the outlet nozzle. To avoid liquid re-entrainment, vapour velocity has to be kept low in the turnaround section of the drum. An additional measure to prevent liquid re-entrainment is a baffle plate below the turnaround section of the drum. The maximum liquid level is the same as vertical settling drums;

d) horizontal drum with the vapour entering at each end on the horizontal axis and a centre outlet;

e) horizontal drum with the vapour entering in the centre and exiting at each end on the horizontal axis;

f) combination of a vertical drum in the base of the flare stack and a horizontal drum upstream to remove the bulk of the liquid entrained in the vapour. This combination permits the use of larger values for the numerical constant in the velocity equation.

A split-entry or -exit configuration can be used to reduce the drum diameter (but increase the length) for large flow rates and should be considered if the vessel diameter exceeds 3,66 m (12 ft). Careful consideration should be given to the hydraulics of split-entry configurations to ensure the flow is indeed split in the desired proportion. Inlet nozzles should include means such as baffles or long sweep elbows to prevent re-entrainment of liquid. Long sweep elbows are typically used up to DN 300 (NPS 12) inlet diameter. Baffles are typically used for larger inlet diameters.

If used, long sweep elbows should be directed away from the outlet location to maximize disengagement by reducing the momentum of the stream and to avoid short-circuiting or “streamlining” flow through the vessel. The long sweep elbow should be located a sufficient distance away from the end of the vessel to mitigate the ricochet of liquid from the end cap. Alternatively, a secondary internal baffle can be installed. Outlet nozzles should have a deflection plate or baffle to minimize carry-over. Other internal baffles can be required to minimize liquid sloshing. If installed, substantial forces associated with the gas velocities and liquid sloshing should be considered in the baffle mechanical design.

The maximum liquid level should not exceed the level where re-entrainment of liquid occurs. Knockout drum facilities typically have level-control instrumentation including local and remote level indication, high and low alarms. Other instrumentation can include pressure indication as well as temperature measurement. One or more pumps are used to remove accumulated liquid from the drum. Depending on the operating philosophy, these pumps typically are started manually with automatic shut-off on low level. The driver for at least one pump should be supplied from a reliable source such as emergency electrical power or steam turbine drive.

Depending on the climate and the nature of the liquid in the system, winterization and/or some means of supplemental heat input can be required for knockout drums to vaporize any volatile liquids (e.g. liquid propane). However, caution should be exercised in assuring that material vaporized in the knockout drum does not condense and possibly solidify in the flare header or stack downstream of the knockout drum.

7.3.2.1.2 Sizing techniques for a horizontal knockout drum are given below. Sizing techniques for a vertical knockout drum are limited to a single entry and single vapour exit [112]. Sizing a knockout drum is generally a trial-and-error process. The first step is to determine the drum size required for liquid entrainment separation. Liquid particles separate (a) when the residence time of the vapour or gas is equal to or greater then the time required to travel the available vertical height at the dropout velocity of the liquid particles and (b) when the gas velocity is sufficiently low to permit the liquid dropout to fall. This vertical height is usually taken as the distance from the maximum liquid level. The vertical velocity of the vapour and gas should be low enough to prevent large slugs of liquid from entering the flare. The presence of small liquid droplets increases thermal radiation fluxes and smoking potential. Long-term field experience has shown that the dropout velocity in the drum may be based on that necessary to separate droplets from 300 µm to 600 µm in diameter. Droplets of larger size can result in incomplete combustion with excessive smoking, possible “burning rain”, and even flame-out. Some types of flares can accommodate larger liquid droplets, so the vendor should always be consulted regarding the adequacy of a specific flare for the burning of liquids.

The dropout velocity [113], expressed in metres/second (feet per second) of a particle in a stream is calculated using Equation (37):

$$u_{\mathrm{c}} = 1{,}15\sqrt{\frac{g \cdot D(\rho_{\mathrm{l}} - \rho_{\mathrm{V}})}{\rho_{\mathrm{V}} \cdot C}} \qquad (37)$$

where

g is the acceleration due to gravity [= 9,8 m/s^2 (32 ft/s^2)];

D is the particle diameter, expressed in metres (feet);

ρ_l is the density of the liquid at operating conditions, expressed in kilograms per cubic metre (pounds per cubic foot);

ρ_v is the density of the vapour at operating conditions, expressed in kilograms per cubic metre (pounds per cubic foot);

C is the drag coefficient (see Figure 16).

This basic equation is widely accepted for all forms of entrainment separation.

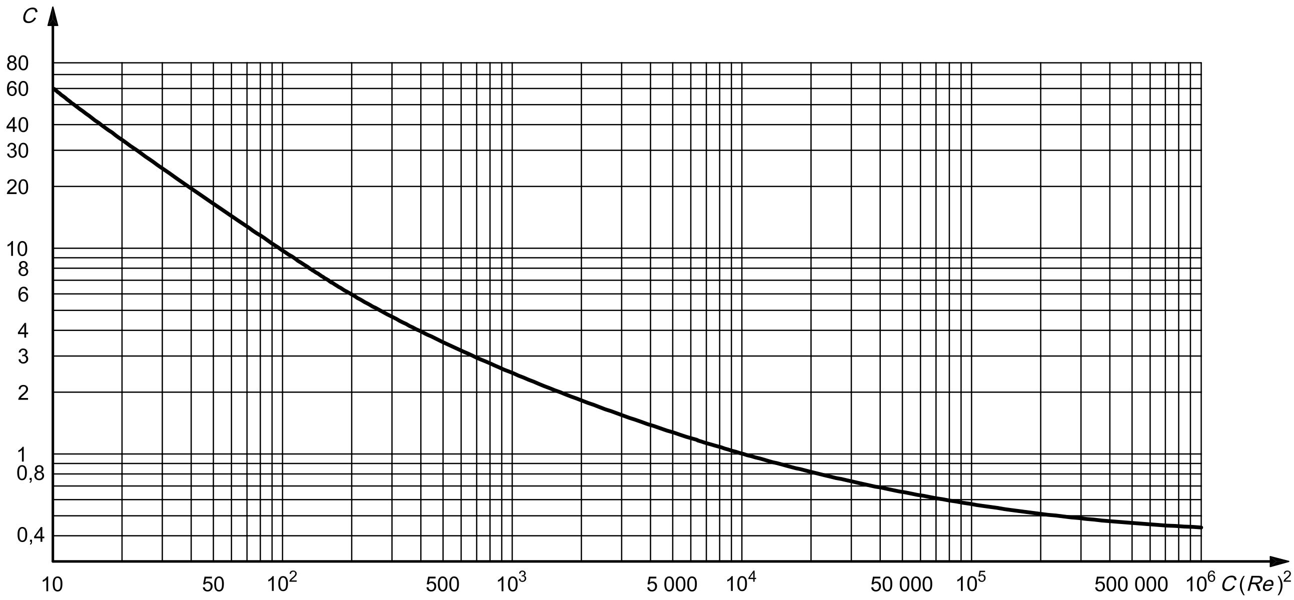


Figure 16 — Determination of drag coefficient

In SI units:

$$C(Re)^2 = \frac{0{,}13 \times 10^8 D^3 (\rho_l - \rho_v)}{\mu^2} \tag{38}$$

In USC units:

$$C(Re)^2 = \frac{0{,}95 \times 10^8 D^3 (\rho_l - \rho_v)}{\mu^2} \tag{39}$$

where

μ is the viscosity of the gas, expressed in megapascal-seconds (centipoise);

ρ_v is the density of the gas, expressed in kilograms per cubic metre (pounds per cubic foot);

ρ_l is the density of the liquid, expressed in kilograms per cubic metre (pounds per cubic foot);

D is the particle diameter, expressed in metres (feet).

NOTE Refer to the section on particle dynamics in the Chemical Engineers' Handbook [46].

The second step in sizing a knockout drum is to consider the effect any liquid contained in the drum can have on reducing the volume available for vapour/liquid disengagement. This liquid may result from

a) condensate that separates during a vapour release, or

b) liquid streams that accompany a vapour release.

The volume occupied by the liquid should be based on a release that lasts 20 min to 30 min. Longer hold-up times can be required if it takes longer to stop the flow. Any accumulation of liquid retained from a prior release (from pressure-relief devices or other sources) should be added to the liquid indicated in items a) and b) above to determine the available vapour-disengaging space. If the knockout drum is used to contain large liquid dumps from pressure-relief devices from other sources and there is no significant flashing, and the liquid is removed promptly, it is not necessary to consider this liquid volume when determining the volume available for vapour disengagement.

It is important to realize as part of the sizing considerations that the maximum vapour release case does not necessarily coincide with the maximum liquid. Therefore, the knockout-drum size should be determined through consideration of both the maximum-vapour-release case as well as the release case with the maximum amount of liquid.

The following sample calculations have been limited to the simplest of the designs [7.3.2.1.1, drum configurations (a) and (b)]. The calculations for drum configurations (d) and (e) are similar, with one-half the flow rate determining one-half the vessel length. The normal calculations are used for drum configuration (c) and are not duplicated here.

The following conditions are assumed.

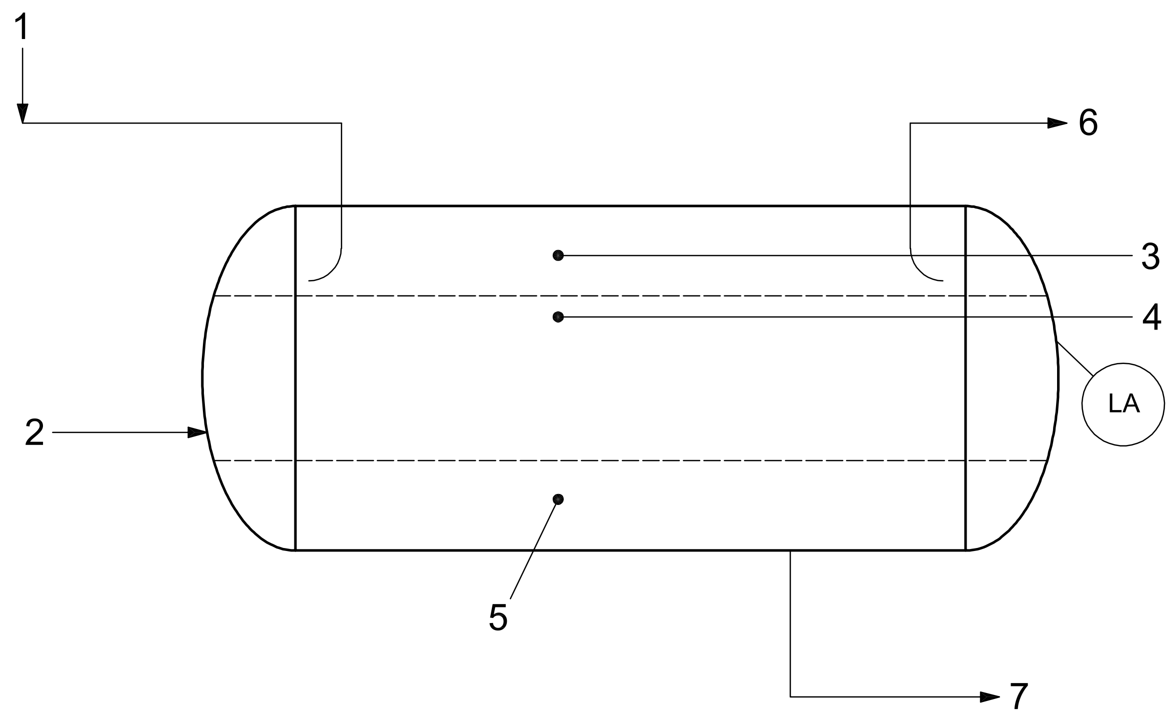
— A single contingency results in the flow of 25,2 kg/s (200 000 lb/h) of a fluid with a liquid density of 496,6 kg/m^3 (31 lb/ft^3) and a vapour density of 2,9 kg/m^3 (0,18 lb/ft^3), both at flowing conditions.

— The gauge pressure is 13,8 kPa (2 psi), and the temperature is 149 °C (300 °F).

— The viscosity of the vapour is 0,01 mPa·s (0,01 cP).

— The fluid equilibrium results in 3,9 kg/s (31 000 lb/h) of liquid and 21,3 kg/s (169 000 lb/h) of vapour.

In addition, 1,89 m^3 (500 US gal) of storage for miscellaneous drainings from the units is desired. The schematic in Figure 17 applies. The droplet size selected as allowable is 300 µm (0,012 in) in diameter.



Key

1 vapour and liquid safety relief valve releases
2 level instrument to indicate when slop and drain volume has been consumed
3 minimum vapour space for dropout velocity
4 liquid hold-up from safety relief valves and other emergency releases
5 slop and drain liquid
6 to flare
7 pumpout

Figure 17 — Flare knockout drum

The vapour rate, R_{V}, in actual cubic metres per second (cubic feet per second), is determined as follows:

In SI units:

$$R_{\text{V}} = \frac{21{,}3}{2{,}9} = 7{,}34 \qquad \text{m}^3\text{/s}$$

In USC units:

$$R_{\text{V}} = \frac{169\ 000}{3\ 600 \times 0{,}18} = 261 \qquad \text{ft}^3\text{/s}$$

The drag coefficient, C, is determined from Figure 16 using Equation (39):

$$C \cdot Re^2 = \frac{0{,}95 \times 10^8 \times 0{,}18 \times (0{,}000\,984)^3 \times (31 - 0{,}18)}{(0{,}01)^2} = 5\ 021$$

From Figure 16, $C = 1{,}3$.

The dropout velocity, u_c, is calculated as follows:

In SI units:

$$u_c = 1{,}150 \left[\frac{9{,}8 \times 300 \times 10^{-6} \times (496{,}6 - 2{,}9)}{2{,}9 \times 1{,}3} \right]^{0{,}5} = 0{,}71 \text{ m/s}$$

In USC units:

$$u_c = 1{,}150 \left[\frac{32{,}2 \times 0{,}000\,984 \times (31 - 0{,}18)}{0{,}18 \times 1{,}3} \right]^{0{,}5} = 2{,}35 \text{ ft/s}$$

A horizontal vessel with an inside diameter, D_i, and a cylindrical length, L, should be assumed. This gives the following total cross-sectional area, A_t:

$$A_t = \frac{\pi}{4} \cdot (D_i)^2 \tag{40}$$

Liquid hold-up for a 30-min release from the single contingency, in addition to the slop and drain volume, is desired. The volume in the heads is neglected for simplicity. The liquid hold-up required, A_{L1} in cubic metres (cubic feet), is calculated as follows:

The slop and drain volume of 1,89 m^3 (500 U.S gal, 66,8 ft^3) occupies a bottom segment as follows:

In SI units:

$$A_{L1} = 1{,}89 \left(\frac{1}{L} \right) \tag{41}$$

In USC units:

$$A_{L1} = 66{,}8 \left(\frac{1}{L} \right) \tag{42}$$

A total of 3,9 kg/s (31 000 lb/h) of condensed liquids with a density of 496,6 kg/m^3 (31 lb/ft^3) accumulated for 30 min occupies a cross-sectional segment [see 7.3.2.1.2 a)] as follows:

In SI units:

$$A_{L2} = \frac{3{,}9}{496{,}6} \times (30 \times 60) \left(\frac{1}{L} \right) \tag{43}$$

In USC units:

$$A_{L2} = \frac{31\,000}{31} \times \frac{30}{60} \left(\frac{1}{L} \right) \tag{44}$$

The cross-sectional area remaining for the vapour flow is expressed as follows:

$$A_v = A_t - (A_{L1} + A_{L2}) \tag{45}$$

The vertical depths of the liquid and vapour spaces are determined using standard geometry and the total drum diameter, h_t, is calculated using Equation (46):

$$h_t = h_{L1} + h_{L2} + h_v \tag{46}$$

where

h_{L1} is the depth of slops and drains;

$(h_{L1} + h_{L2})$ is the depth of all liquid accumulation;

h_v is the remaining vertical space for the vapour flow.

The adequacy of the vapour space is verified by determining the liquid dropout time, θ, using Equation (47):

In SI units:

$$\theta = \left(\frac{h_v}{100}\right)\left(\frac{1}{u_c}\right) \tag{47}$$

In USC units:

$$\theta = \left(\frac{h_v}{12}\right)\left(\frac{1}{u_c}\right) \tag{48}$$

where

θ is the liquid dropout time, expressed in seconds;

h_v is the vertical drop available for liquid dropout, expressed in centimetres (inches);

u_c is the dropout velocity, expressed in metres per second (feet per second).

The velocity of N vapour passes, based on one vapour pass, is determined from Equations (49) and (50) for a volume flow rate of 7,34 m^3/s (260 ft^3/s):

In SI units:

$$u_v = \left(\frac{7{,}34}{N}\right)\left(\frac{1}{A_v}\right) \tag{49}$$

In USC units:

$$u_v = \left(\frac{260}{N}\right)\left(\frac{1}{A_v}\right) \tag{50}$$

where

A_v is the cross-sectional area, expressed in square metres (square feet);

N is the number of vapour passes;

u_v is the vapour velocity, expressed in metres per second (feet per second).

The drum length required, L_{min}, is determined as follows:

$$L_{min} = u_v \cdot \theta \cdot N \tag{51}$$

L_{min} shall be less than or equal to the above assumed cylindrical drum length, L; otherwise, the calculation shall be repeated with a newly assumed cylindrical drum length.

Tables 15 and 16 summarize the preceding calculations for one pass for horizontal drums with various inside diameters to determine the most economical drum size. Drum diameters in 15 cm (6 in) increments are assumed, in accordance with standard head sizes.

Table 15 — Optimizing the size of a horizontal knockout drum (SI units)

Trial No.	D_1 [a]	L [b]	Cross-sectional area m²				Vertical depth of liquid and vapour spaces cm				θ [c]	u_v [d]	L_{min} [e]
			A_t	A_{L1}	A_{L2}	A_v	h_{L1}	$h_{L1}+h_{L2}$	h_v	h_t			
1	2,44	5,79	4,67	0,33	2,45	1,89	30	140	104	244	1,45	3,9	5,6
2	2,29	6,25	4,10	0,30	2,27	1,53	29	137	91	229	1,28	4,8	6,2
3	2,13	6,86	3,57	0,28	2,07	1,22	28	133	81	213	1,13	6,0	6,7
4	1,98	7,62	3,08	0,25	1,86	0,98	27	128	70	198	0,98	7,5	7,4

NOTE 1 The data in this table are in accordance with the example given in text for one pass vapour flow.

NOTE 2 The values in this table are rounded-off conversions of the values in Table 16.

[a] D_1 is the assumed drum inside diameter, expressed in metres.

[b] L is the assumed drum cylindrical length, expressed in metres.

[c] θ is the liquid dropout time, expressed in seconds.

[d] u_v is the vapour velocity, expressed in metres per second.

[e] L_{min} is the required drum length, expressed in metres.

It can be concluded from Table 15 that

— all of the drum sizes above fulfill the design requirements;

— the most suitable drum size should be selected according to the design pressure, material requirements and corrosion allowance as well as layout, transportation and other considerations;

— the choice of two-pass flow, as shown in Figure 17, is optional.

Table 16 — Optimizing the size of a horizontal knockout drum (USC units)

Trial No.	D_1 [a]	L [b]	Cross-sectional area ft^2				Vertical depth of liquid and vapour spaces in				θ [c]	u_v [d]	L_{min} [e]
			A_t	A_{L1}	A_{L2}	A_v	h_{L1}	$h_{L1}+h_{L2}$	h_v	h_t			
1	8,0	19,0	50,26	3,52	26,32	20,43	11,75	55,0	41,0	96	1,45	12,76	18,5
2	7,5	20,5	44,17	3,26	24,39	16,52	11,43	54,0	36,0	90	1,28	15,78	20,2
3	7,0	22,5	38,49	2,97	22,22	13,29	11,00	52,3	31,7	84	1,13	19,62	22,1
4	6,5	25,0	33,18	2,67	20,00	10,51	10,5	50,4	27,6	78	0,98	24,82	24,3

NOTE The data in this table are in accordance with the example given in the text for one-pass vapour flow.

[a] D_1 is the assumed drum inside diameter, expressed in feet.

[b] L is the assumed drum cylindrical length, expressed in feet.

[c] θ is the liquid dropout time, expressed in seconds.

[d] u_v is the vapour velocity, expressed in feet per second.

[e] L_{min} is the required drum length, expressed in feet.

It can be concluded from Table 16 that

— all of the drum sizes above fulfil the design requirements;

— the most suitable drum size should be selected according to the design pressure, material requirements and corrosion allowance as well as layout, transportation and other considerations; and

— the choice of two-pass flow, as shown in Figure 17, is optional.

If a vertical vessel is considered, the vapour velocity is equal to the dropout velocity, which is 0,71 m/s (2,35 ft/s). The volume flow rate is 7,34 m^3/s (260 ft^3/s). The required cross-sectional area, A_{cs}, of the drum, in square metres (square feet) is determined as follows:

In SI units:

$$A_{cs} = \frac{7,34}{0,71} = 10,3 \text{ m}^2 \quad (52)$$

In USC units:

$$A_{cs} = \frac{260}{2,35} = 110,6 \text{ ft}^2 \quad (53)$$

The drum diameter, D, is determined as follows:

In SI units:

$$D = \sqrt{10,3 \times \frac{4}{\pi}} = 3,6 \text{ m} \quad (54)$$

In USC units:

$$D = \sqrt{110,6 \times \frac{4}{\pi}} = 11,9 \text{ ft} \quad (55)$$

7.3.2.1.3 During flaring situations, the liquid level in all flare knockout drums should be monitored. High-level alarms should be installed to alert the operators of abnormal knockout conditions. Therefore, these alarms should be set at a relatively low level so there is the required hold-up time between the alarm point and the high liquid level for normal facility shutdown. Redundant level transmitters may be considered if high alarm reliability is needed. Minimum levels in knockout drums should be maintained to ensure sufficient free volume is available in the event of a flaring situation. Level transmitters shall be designed for operation at the minimum design temperature of the knockout drum.

7.3.2.1.4 Check level instrumentation, including any alarms, low level pump shutdown instruments, etc. in accordance with vendor recommendations during planned maintenance intervals. The drum level should be checked in the field on a regular basis: field versus any remote level indication. Since knockout drums and liquid seals are typically considered a "dirty" service, routine cleaning of gauge glasses can be required.

There is a potential for the presence of pyrophoric iron in flare-system components such as liquid-seal drums and, in particular, knockout drums. When any vessel entry is being planned, the potential for the presence of this material should be included in the considerations for safe entry.

7.3.2.2 Liquid-seal drums

7.3.2.2.1 A liquid-seal drum can have many uses as set forth in 6.4.3.6.1. A liquid seal is sometimes used to prevent air entry into the flare-header system. Following a hot release, cooling and/or condensation of vapour at low-flow or no-flow conditions can form a vacuum condition in the flare header. Under such conditions, air can be drawn into the flare system through the flare tip. The rate of contraction accelerates dramatically if the cooling leads to condensation of components of the contained gas.

Factors to consider when assessing this potential are

a) the potential contraction of the contained gas when cooled,

b) the volume of the flare system,

c) the vacuum rating of the flare system, and

d) the anticipated cooling rate of the flare system, which can be affected by insulation or sudden cooling from wind and rain.

To prevent air entry, it is necessary that the seal dip-leg height and the density and amount of seal liquid within the drum be sufficient to prevent the seal from being broken as a result of the vacuum formed in the flare header. The physical dip-leg height is measured from the top opening of the seal head or end piece (e.g. the top of the V-notches on the end of the pipe) to the bottom of the horizontal section of the flare header piping immediately upstream of the inlet leg. The relative elevations of the flare header and other equipment and other factors can limit the vacuum sealing capability. If it is necessary to have the liquid seal inlet some height above the flare-header elevation, then the flare header shall be sloped to avoid low points. The seal drum should be designed to provide the volume of liquid (without credit for make-up liquid) to fill the vertical seal leg up to the specified vacuum. It is important that the purchaser states this performance requirement on the data sheet.

If possible, a minimum height of 3 m (10 ft) should be considered. A tank-blanketing regulator, pressure regulator and/or pressure switch/transmitter that dumps extra purge gas into the system in the event of vacuum can also be considered in addition to or in lieu of the water seal.

The flare header pressure at which gas begins passing through the seal can vary depending upon the purpose of the water seal (i.e., prevent air infiltration, act as a flame arrester, act as a staging device or provide back pressure to a flare gas recovery system). The gas pressure at the start of flow through the liquid seal can vary from 50 mm (2 in H_2O) to 3 050 mm (120 in H_2O) or more. Typical seal depths are equivalent to a gauge pressure of 13,8 kPa to 34,5 kPa (2 psi to 5 psi) for staged flares or 6,9 kPa to 13,8 kPa (1 psi to 2 psi) where a flare-gas recovery system is used. For general applications, a seal depth of 150 mm (6 in) is common. In normal operation, a gas flow that exceeds the lower-stage capacity or capacity of the flare gas recovery unit causes waste gas to start flowing through the seal.

A properly designed and operated liquid seal should allow gas to pass through the seal with minimal surging in gas flow and/or upstream gas pressure. The design of the liquid seal internals and the sizing of the vessel can have a significant impact on the ability of the seal to meet this performance objective. For example, a common design for the end of a dip-leg pipe uses “V” notches cut into the end of the pipe wall. The design is less effective than the proprietary designs developed by vendors of liquid seals. These proprietary designs that use alternative design guidelines can offer economic or operational advantages. The purchaser should determine the applicability of such designs to the purchaser’s system and situation. Additional discussion of liquid seals and examples of alternate seal head designs can be found in Reference [114].

Liquid seals typically use water as the seal medium, however, other fluids are possible. Fluid selection requires consideration of freeze protection in cold climates, hydrocarbon/water separation, implications of carryover, compatibility with the relief stream, cost, availability and disposal. In facilities that have cryogenic products released into the flare header, consideration should be given to the effect of the cold material on the seal medium. Water seals are not recommended if there is a risk of obstructing the flare system due to an ice plug. Alternate sealing fluids such as a glycol/water mixture or other means to prevent freezing can be required.

Examples of liquid-seal drums are provided in Figures D.1 and D.2. They act as a final or secondary knockout drum and, as such, should be designed based on the same principles as a vertical settling drum. When appropriate, liquid-seal drums should include anti-sloshing baffles, which act to dampen any pressure fluctuations created by the liquid movement in the drum (see Reference [115]). An oil-skimming device should also be included in the design, which allows the removal of any hydrocarbon liquid that condenses as it passes through the liquid seal. The skimming device can be connected to a free-flowing drain or loop seal. The mechanical depth of the loop seal should give consideration to the maximum vessel pressure under any operating condition, as well as the minimum specific gravity of the liquid in the loop seal.

7.3.2.2.2 Consideration should be given to providing a continuous flow of seal fluid (typical for water seals), which allows for the continuous skimming of hydrocarbons as well as maintaining liquid level. Proper liquid-seal drum operation is dependent upon maintaining the design liquid level in the seal. Routine surveillance and hydrocarbon skimming, if applicable, are required to ensure proper seal operation. Pulsing combustion noise at the flare tip is a problem sometimes encountered relating to seal drums. The cause is liquid sloshing in the seal drum and can be corrected by installing stiling or perforated baffles in the seal drum.

7.3.2.2.3 Like many components of a flare system, liquid-seal drums are generally in a position where they can be maintained only on a full plant shutdown. However, normal inspection and operational maintenance of liquid-seal drum ancillaries should be carried out by the operators as part of their rounds. Check level instrumentation and gauges including any alarms, pump out switches, etc. in accordance with vendor recommendations. Also look for any leaks or signs of corrosion, especially on connections and flanges. Leaks should be reported and corrected, since a leak can allow gas to escape from the flare or air to enter the flare at low flow rates and draft conditions from high-temperature or low-density flare gas.

The drum level should be checked in the field on a regular basis: field versus any remote level indication. Since liquid seals are typically considered a “dirty” service and tend to act as a scrubber removing any solids or liquids, routine cleaning of gauge glasses can be required. Dirt build-up can also plug the drain or overflow; these should be checked periodically to determine that they are still functional.

Depending on the flare waste gases, sampling of the seal liquid (normally water) is recommended for a new flare-seal drum (or a drum in which the service has been changed) during the initial few months or even the first year of operation. Liquid samples from the seal drum should be checked for levels of sulfur, chlorinated wastes or other components that can damage the seal drum and internals. Check with the vendor for specific limitations. Proper sampling procedures should be followed to prevent the leakage of flare gas or air entry into the flare. If any problem substances are found, then consider a continuous water flush or treatment programme on the water seal to protect it and the stack in accordance with vendor recommendations.

There is also a potential for the presence of pyrophoric iron in the liquid-seal drum. When any vessel entry is being planned, the potential for presence of this material should be included in the considerations for safe entry.

During plant shutdown, inspect the internals of the liquid-seal drum (such as bubble plate) for fouling or plugging, remove any build-up of dirt or foreign matter and check for corrosion. Report any major corrosion in order that these sections can be repaired or replaced, as needed.

CAUTION — Only open the seal-drum-access manway after the complete flare system has been shut down, blinded and purged of all hydrocarbons, gases and vapours. Check oxygen level and follow plant safety procedures before entry (confined space). Inspection approximately once every two years or less is recommended if liquid-seal pressure drop increases or pulsations occur.

WARNING — Possible gas leaks from the liquid-seal drum can be caused by the following.

a) excessive back pressure from fouling, freezing or plugging of the flare equipment (arrestor, burner or flare stack), which can displace water from the overflow and/or drain u-trap thus allowing gas to escape. Pressures downstream of seal should be monitored to ensure pressures are not exceeded.

b) loss of liquid in water seal and trap, which shall always be filled with liquid to contain the flare gas.

7.3.2.2.4 The area for the gas flow above the liquid level should be at least three times the inlet pipe cross-sectional area to prevent surges of gas flow to the flare. As described in 7.3.2.2.1, the flare-header pressure at which gas begins passing through the drum can vary. The submergence depth is set by this pressure with due consideration for the specific gravity of the seal fluid [see Equations (56) and (57)]. This back pressure should not exceed the maximum back pressure allowable in the vent header (see 7.3.1.3). This back pressure sets the maximum distance, h [see Equations (56) and (57)], that the inlet pipe is submerged.

In SI units:

$$h = \frac{102p}{\rho} \tag{56}$$

In USC units:

$$h = \frac{144p}{\rho} \tag{57}$$

where

h is the maximum distance the inlet pipe is submerged, expressed in metres (feet);

p is the maximum header exit absolute pressure, expressed in kilopascals (pounds per square inch);

ρ is the sealing liquid density, expressed in kilograms per cubic metre (pounds per cubic foot).

The area for the gas above the liquid surface should equal at least that of a circle with a diameter, D, that is equal to $2d$, where d is the diameter of the inlet gas pipe. This can be derived assuming a vertical vessel that has an internal area equal to $(\pi D^2)/4$ and an inlet pipe with an area equal to $(\pi d^2)/4$. The annular area is $(\pi/4)(D^2 - d^2)$. Since the suggested ratio is 1:3, then:

$$(D^2 - d^2) = 3d^2 \qquad \therefore D^2 = 4d^2 \qquad \therefore D = 2d \tag{58}$$

The height of the vapour space in a vertical seal drum should be approximately 0,5 to 1,0 times the diameter, D, to provide disengaging space for entrained seal liquid. A minimum dimension of 1 m (3 ft) is suggested.

In some situations, special considerations can affect the size of a seal drum. One such occurrence is a large flow of hot vapour into the vent header. The vacuum created when this vapour cools can pull sufficient liquid into the header to break the seal, thus allowing air to be drawn into the flare system. To prevent this occurrence, the inlet line should be constructed to form a vacuum leg. The total vertical height of the inlet leg at the seal drum is determined by the maximum vacuum expected. The volume of liquid in the inlet line at the maximum vacuum should be obtained from the seal drum. This requirement can necessitate an increase in the size of the drum.

7.3.2.3 Sizing a quench drum

The sizing criteria for quench drums depend so closely on the design of the drum internals, the liquid loading, the amount of condensation and other features specific to the particular installation that no generally meaningful sizing rules can be established. A common criterion is to reduce the temperature of the stream so that the exit liquid and vapour do not exceed the range of 66 °C to 93 °C (150 °F to 200 °F) and, typically, to assume that no more than 40 % to 50 % of the liquid fed is vaporized. Scheiman's articles [116], [117] cover the sizing criteria for one type of internals frequently used in this service.

7.3.2.4 Design details

A convenient way to state the detailed design requirements for a seal drum and a knockout drum is to use data sheets from API Std 537.

Design details that can be applicable to knockout drums and seal drums include the following.

a) Anti-swirl or anti-vortex baffles should be used on the liquid outlet lines.

b) Internally extended liquid outlet nozzles should be used so that sediment settles out in the drums, not in low spots in the lines.

c) Antifreeze, siphon-type drains should be used for normal manual drains if a freezing problem exists.

d) Provisions should be made for water leg or boot and water removal if three-phase separation is expected.

e) Handholes [DN 100 to DN 200 (NPS 4 to NPS 8) nozzles] should be present on the bottom of the drum to permit thorough cleaning. These nozzles should have DN 40 (NPS 1-1/2) or DN 50 (NPS 2) valves in the blind flange to permit complete draining of the vessels before opening.

f) Allowance should be made for blinding, venting, purging (steaming) and preparing the vessel for entry where manways are provided.

g) Provisions should be made for heating the contents of the vessel if cold weather, auto-refrigeration, viscous or congealing liquids can introduce problems. If internal coils are needed, consideration should be given to coil drainage. The coils should have a generous corrosion allowance and adequate support to prevent mechanical failure. Because one side of the vessel shell cannot be inspected, heating jackets that use the vessel shell as one wall should be avoided.

h) Most knockout drums and seal drums are operating at relatively low pressures. To ensure sound construction, a minimum design gauge pressure of 345 kPa (50 psi) is suggested for knockout drums in subsonic flare or other low-pressure applications. A vessel with a design gauge pressure of 345 kPa (50 psi) should not rupture if a deflagration occurs. Stoichiometric hydrocarbon-air mixtures can produce peak explosion pressures on the order of seven to eight times the absolute operating pressure. Most subsonic-flare seal drums operate in the range of gauge pressure from 0 kPa to 34 kPa (0 psi to 5 psi).

i) In designing vessel nozzles, attachments, supports and internals, one should consider shock loadings that result from thermal effects, slugs of liquid or gas expansion.

j) Try cocks for liquid-level detection can be desirable in addition to or instead of level gauges.

k) Facilities to provide for continuous removal or intermittent manual skimming of hydrocarbons that can accumulate should be considered. Constant skimming by means of continuous addition of seal liquid and overflow to drain can be used. Provisions for periodically raising the level of the seal liquid to force lighter fluid out through a skimmer connection are permissible. The designer is cautioned to review the proposed system to ensure that lighter material cannot build up to the point at which a false (non-design) sealing effect is established.

l) Instrumentation components should be the simplest and most rugged available and should be easily maintained (externally mounted and valved). The use of seals instead of valves and of valves instead of traps is preferred, primarily because of the nature of the materials handled and the conditions under which it is necessary that these components operate. On-off valves with large flow areas are frequently preferred to small-passage throttling valves.

m) Provisions for establishing and maintaining an adequate seal level are recommended.

n) If corrosion can occur at the seal fluid/vapour interface, an adequate corrosion allowance should be used. Such corrosion can occur even in hydrocarbon systems that use water as the seal fluid or in areas where water can collect at low points in the system.

In addition to these common details, some details are specific to the various types of equipment. Knockout drums can be of the horizontal or vertical type; and they should be provided with a pump or draining facilities and instrumentation to remove the accumulated liquids to a tank, sewer or other location. The actual type of disposal used depends on the characteristics and hazards associated with the liquids removed. The design of liquid-removal facilities for a knockout drum depends on the size of the vessel and the extent or probability of liquid occurring in the system.

In the simplest system, the vessel might have only a manually operated drain valve and a liquid-level sight glass for reference. A liquid-removal pump is frequently used on knockout drums. Knockout-drum transfer pumps are sized for a minimum net positive suction-head requirement. Their specification should also consider the maximum liquid temperature that can be encountered.

A high-level alarm, a manual starting switch and an automatic shutoff switch to the pump motor are generally provided. More elaborate arrangements can also have high- and low-level alarms and level controls that operate a motorized drain valve or a liquid-removal pump. Where a drain valve is used, the on-off type is more common; however, a throttling type may be employed. The high-liquid level in the drum is limited so that the cross-sectional area of vapour passage is not reduced. The low-liquid limit is established to prevent vapour from entering the liquid removal system.

The seal drum should be located between the stack and the other header drums and as close to the flare stack as is practical. A variation of a seal drum is often incorporated into the base of the flare stack where the flare line enters the stack. The configuration of a seal drum may consist of a vessel partially filled with a sealing liquid (e.g. water).

The problem of surging in seal drums can be minimized by the use of slots or V notches on the end of the dip pipe so that increasing flow area is provided as the gas flow increases, utilizing a principle similar to that involved in the design of a bubble cap. Occasionally, it can be desirable to increase the size of the inlet line inside the drum to reduce gas velocity and allow enough circumference for the slots. The desired sealing level may also be maintained by means of an automatic controller operating on the liquid supply line. Low- and high-level alarms are sometimes used for warning in case the liquid is not maintained within the desired levels. An adequately sized drain line with shutoff valves should be provided for removing the liquid.

The quenching fluid in a quench drum or tower may be water, gas oil or another suitable liquid. The quench liquid collects in the bottom of the drum for subsequent removal. Any increase in inlet-line size should be accompanied by a corresponding change in the vessel size to maintain the 3:1 ratio (see 7.3.2.2.4). The requirement for quenching may be monitored by a temperature- or flow-switch in the relief-discharge header. The liquid in the bottom of the drum or tower may be automatically controlled. Removed liquid may be cooled and recycled, dispatched to sewers or sent to equipment for recovery of condensed-vapour components. Alarms may be provided to signal operators in the event that design liquid levels are exceeded. Heating equipment should be provided in the liquid-collection zone to keep the system operative in situations where low ambient temperatures can be a factor.

7.3.3 Flares

7.3.3.1 General

Additional guidance is provided in API Std 537, which addresses mechanical design, operation and maintenance of flare equipment. API Std 537 also provides data sheets for exchanging both process and mechanical design information. Note that flares are considered pollution-abatement equipment and are usually subject to environmental regulations and permitting requirements for their use and location.

7.3.3.2 Sizing

7.3.3.2.1 Factors governing the sizing of flares are covered in 7.3.3.2.2 through 7.3.3.2.6. General considerations involved in the calculation of these requirements are discussed in Clause 6. Examples covering the full design of a flare stack are given in Annex C. Note that flare-diameter calculations are based on a basic flare. Most commercial flares have flame retainers that restrict flow area by 2 % to 10 %, which should be accounted for in the flare and header sizing.

7.3.3.2.2 Flare-stack diameter is generally sized on a velocity basis, although pressure drop should be checked. It can be desirable to permit a velocity of up to 0,5 Mach for a peak, short-term, infrequent flow, with 0,2 Mach maintained for the more normal and possibly more frequent conditions for low-pressure flares. This depends on the following:

a) volume ratio of maximum conceivable flare flow to anticipated average flare flow;

b) probable timing, frequency and duration of those flows;

c) design criteria established for the project to stabilize flare burning.

However, sonic-velocity operation can be appropriate for high-pressure flares. Smokeless flares should be sized for the conditions under which they are to operate smokelessly. Equations (27) or (28) can be used to calculate the Mach number (see 7.3.1.3.3). Velocity limitations imposed by regulatory agencies (see Reference [118]) might not apply to flares in emergency-relief service.

Pressure drops as large as 14 kPa (2 psi) have been satisfactorily used at the flare tip. Modern conventional flare tips with proper flame stabilization can operate well above this level. Most flare vendors also have a line of special-duty high-pressure flares that can operate at gauge pressures around 700 kPa (100 psi) or higher. This general class of flare designs is recognized by API Std 537 and usually operates smokelessly without steam or air assistance. Too low a tip velocity can cause heat and corrosion damage. The burning of the gases becomes quite slow, and the flame is greatly influenced by the wind. The low-pressure area on the downwind side of the stack can cause the burning gases to be drawn down along the stack for 3 m (10 ft) or more. Mechanical details of flare burners relating to stack downwash effect can be found in API Std 537.

7.3.3.2.3 The flare-stack height is generally based on the radiant-heat intensity generated by the flame. Equation (24) in 6.4.2.3.3 applies. The recommended levels of radiation intensity, K, are given in Table 9.

The quality of combustion affects the radiation characteristics. Use of the fraction of heat radiated, F, based on the US Bureau of Mines data given in Table 10, is considered to result in a reasonable but conservative stack height.

7.3.3.2.4 Another factor to be considered is the effect of wind in tilting the flame, thus varying the distance from the centre of the flame, which is considered to be the origin of the total radiant-heat release, with respect to the plant location under consideration. A generalized curve for approximating the effect of wind is given in Figure 9.

7.3.3.2.5 Where there is concern about the resulting atmospheric dispersion if the flare were to be extinguished, the information referred to in 6.3.1 and in Gifford's article [76] may be used to calculate the probable concentration at the point in question.

7.3.3.2.6 A complete description of an enclosed ground flare can be found in API Std 537, where an entire chapter is devoted to this type of flare. Options exist for bottom or side firing, staged or unstaged control and steam-, air-, pressure- or non-assisted burners. This type of flare system is often relatively complex and may involve a number of independent burner systems. Accounting for the various interactions between burners, fences, stack, smoke suppression equipment, piping and wind/weather requires considerable experience and can involve detailed flow modelling before a workable design can be achieved.

7.3.3.3 Design details

7.3.3.3.1 Smoke-free operation of flares can be achieved by various methods, including steam injection, injection of high-pressure waste gas, forced draft air, operation of flares as a premix burner or distribution of the flow through many small burners. A common type of smokeless flare involves steam injection. For more detailed information on the theory of smokeless burning with steam injection, refer to 6.4.3.2.

The amount of steam required for smokeless burning depends on the vapour flow rate to be burned and the detailed composition of the mixture. Key parameters involving smokeless combustion include percentage of unsaturates, percentage of inerts and the mixture relative molecular mass. Certain specific compounds require special consideration by the vendor. Examples include ethylene, butadiene, acetylene and ethylene oxide. Data sheets from API Std 537 provide a convenient means for specifying the composition.

The flare may be designed for various degrees of smokelessness. Many state and federal regulations state the smokeless requirement in the form “No operator shall allow the flare emissions to exceed 20 % opacity for more than 5 min in any consecutive 2-h period.” This type of regulation is usually the basis for designing flares to achieve Ringleman 1 (20 % opacity) performance. Other applications can require Ringleman 0 (zero opacity) for regulatory or community-relations reasons. It is necessary for the user to understand the local regulatory requirements that govern smokeless requirements.

Table 11 may be used to estimate steam requirements as a function of composition. For mixtures, the estimate of steam requirements can be proportioned based on the specific mass fraction of combustible component in the mixture (i.e., exclude inerts). In any event, if a proprietary smokeless flare is purchased, the manufacturer should be consulted about the minimum necessary steam rate.

7.3.3.3.2 One of the most common methods of preventing propagation of flame into the flare system as a result of the entry of air is to install a seal drum as described in 7.3.2.2. The continuous introduction of purge gas at an adequate rate can also be used to reduce the possibility of flashback with or without a seal drum. Flame arresters are seldom used in a flare system for flashback protection because they are subject to plugging. The hazards associated with an obstruction in a flare system are of such a serious nature that flame arresters could be recommended only if vapours are non-corrosive, dry, free from any liquid that can congeal and free from any substances that can cause obstruction. Although these situations are rarely encountered with flare systems, an additional disadvantage to a flame arrester that is pertinent to flare systems is that during the cooling that follows a warm discharge, air can be drawn back through the flame arrester into the system. See Reference[121] for additional precautions associated with flame arresters.

Studies [122], [123], [124] have shown that hydrocarbon-air mixtures can be diluted with inerts to be below the lower flammable limit. Alternatively, the hydrocarbon-air mixture can be enriched with combustible gas so it is above the upper flammable limit and too rich to burn (see NFPA 69 [35] for details).

7.3.3.3.3 Air infiltration down the flare stack from wind or density effects can be mitigated by use of purge gas. If purge gas is required, the user shall assure the reliability of its supply. The amount of purge gas required can be reduced by the use of a purge-reduction seal (e.g. buoyancy seal or velocity seal; see 6.4.3.6.2).

For lighter-than-air purge gases, Equations (59) and (60) can be used to determine Q, he purge gas rate, expressed in normal cubic metres per hour (standard cubic feet per hour) for continuous purge requirements in open flares without the effect of buoyancy seal or velocity seal [121], [122].

In SI units:

$$Q = 190,8D^{3,46}\,\frac{1}{y}\ln\left(\frac{20,9}{\mathrm{O_2}}\right)\left(\sum_{i}^{n} C_{\mathrm{i}}^{0,65}\cdot K_{\mathrm{i}}\right) \tag{59}$$

In USC units:

$$Q = 0,070\,68D^{3,46}\,\frac{1}{y}\ln\left(\frac{20,9}{\mathrm{O_2}}\right)\left[\sum_{i}^{n} C_{\mathrm{i}}^{0,65} K_{\mathrm{i}}\right] \tag{60}$$

where

D is the flare stack diameter, expressed in metres (inches);

y is the column depth at which the oxygen concentration (O_2) is predicted, expressed in metres (feet);

O_2 is the oxygen volume fraction, expressed as a percentage;

C_{i} is the volume fraction of component i, expressed as a percentage;

K_{i} is a constant for component i.

The following are typical values for K_{i}:

- hydrogen: $K = +$ 5,783;
- helium: $K = +$ 5,078;
- methane: $K = +$ 2,328;
- nitrogen: $K = +$ 1,067 (no wind);
- nitrogen: $K = +$ 1,707 [with a wind speed of approximately 7 m/s (15 miles per hour)];
- ethane: $K = -$ 1,067;
- propane: $K = -$ 2,651;
- CO_2: $K = -$ 2,651;
- C_{4+}: $K = -$ 6,586.

NOTE Steam or other condensables are not suitable purge gases.

Equations (59) and (60) can be simplified to Equations (61) and (62) using the standard criteria of limiting the oxygen volume fraction to 6 % at a distance of 7,62 m (25 ft) down the flare stack (except that lower oxygen concentrations should be used for certain compounds such as hydrogen):

In SI units:

$$Q = 31,25\,D^{3,46}\cdot K \tag{61}$$

In USC units:

$$Q = 0,003\,528\,3\,D^{3,46}\cdot K \tag{62}$$

where

Q is the purge gas rate, expressed in normal cubic metres per hour (standard cubic feet per hour);

D is the flare stack diameter, expressed in metres (inches);

K is a constant (see above).

Based on recent test data involving natural gas production facility flares [124], a significant reduction in purge rates as predicted by Equations (59), (60), (61) and (62) may be taken under certain constraints. The user is cautioned not to extrapolate outside the bounds and conditions under which the tests were conducted.

If the gas in the stack (e.g. hydrogen or methane) is lighter than air, the pressure in the bottom of the stack can be lower than atmospheric, even with some outflow from the top of the stack. This condition creates a situation in which any flange leaks, open drains/vents or other openings in the flare header draws air into the flare header, resulting in potential for an internal explosion. For this reason, the users are cautioned to maintain the integrity of their equipment and follow proper safety precautions when opening an active header.

7.3.3.3.4 Note that purge rates higher than those given by Equations (59), (60), (61) and (62) may be required

a) to establish an initial, air-free condition during start-up;

b) in transient conditions associated with a passing rainstorm cooling down header exposed to the sun;

c) after venting a hot condensable release into the flare header;

d) after venting a stream containing significant amounts of compounds that are easily detonated or have unusually wide flammability limits.

Gases or vapours with unusually high burning velocities, such as hydrogen and acetylene, should be evaluated for the possibility of flashback (see 6.4.3.6.1 and 6.4.3.6.2).

7.3.3.3.5 Once the required quantity of purge gas has been established, the injection rate should be controlled by a fixed orifice, rotometer or other device that ensures the supply remains constant and is not subject to instrument malfunction or maladjustment. Consult the vendor to determine purge rates to prevent burning inside the flare tip.

7.3.3.3.6 To ensure ignition of flare gases, continuous pilots with a means of remote ignition are recommended for all flares. Some regulations can require the presence of a continuous pilot flame to be proven by thermocouple or equivalent means. The most commonly used type of igniter is the flame-front propagation type, which uses a spark from a remote location to ignite a flammable mixture. Pilot-igniter controls are located near the base of elevated flares and at least 30 m (100 ft) from ground flares (see 6.4.3.6.4). An extensive discussion of pilots and ignition systems can be found in API Std 537.

7.3.3.3.7 The fuel-gas supply to the pilots and igniters should be highly reliable. Since normal plant fuel sources can be upset or lost, it is desirable to provide a backup system connected to the most reliable alternative fuel source, with a provision for automatic cut-in on low pressure. The use of a waste gas with low-energy content or with unusual burning characteristics should be avoided. Parallel instrumentation for pressure reduction is frequently justifiable. The flare fuel system should be carefully checked to ensure that hydrates cannot present a problem. Because of small lines, long exposed runs, large vertical rises up the stack and pressure reductions, use of a liquid knockout pot or scrubber after the last pressure reduction is frequently warranted. If at all feasible in terms of distance, relative location and cost, a low-pressure alarm should be installed on the fuel supply after the last regulator or control valve so that operators are warned of any loss of fuel to the pilots.

7.3.4 Vent stacks

7.3.4.1 Sizing

Where the atmospheric vent handles combustible vapours, the outlet from the vent should be elevated approximately 3 m (10 ft) above any adjacent equipment, building, chimney or other structure (see 6.3 for additional discussion). Provisions should be made for drainage of each vent pipe so that liquid cannot accumulate in the vent.

The size of a vent stack is determined by the available pressure drop and by any minimum velocity required to prevent hazardous conditions due to combustible or toxic material at grade or working levels. Calculation methods applicable to a vent stack that discharges hazardous materials are given in Clause 6. Normally, a size is selected that results in a high discharge velocity; e.g. a velocity of 150 m/s (500 ft/s) provides excellent dispersion. The size should be checked to ensure that sonic flow is not established or, if it is, that allowance has been made for the pressure discontinuity at the discharge end in calculating pressure drop.

A sample calculation follows.

For this calculation, the following conditions should be assumed: The maximum relief rate, $\dot{m}$, is 31,5 kg/s (250 000 lb/h). The relative molecular mass of the vapour, M, is 44. The absolute temperature of the vapour just inside the exit point from the vent stack, T, is 361 K (650 °R). The exit velocity, v, is 150 m/s (500 ft/s). The absolute pressure of the vapour just inside the exit point from the vent stack, p, is 101 kPa (14,7 psi). The gas constant, R, is 8,3 in SI units (10,7 in USC units). The density, ρ, is then calculated as in Equations (63) and (64):

In SI units:

$$\rho = \frac{M \cdot p}{R \cdot T} = \frac{44 \times 101}{8{,}3 \times 361} = 1{,}48 \text{ kg/m}^3 \tag{63}$$

In USC units:

$$\rho = \frac{M \cdot p}{R \cdot T} = \frac{44 \times 14{,}7}{10{,}7 \times 650} = 0{,}1 \text{ lb/ft}^3 \tag{64}$$

The tip area, A_{T}, is determined by Equations (65) and (66):

In SI units:

$$A_{\text{T}} = \frac{\dot{m}}{\rho \cdot v} = \frac{31{,}5}{1{,}48 \times 150} = 0{,}14 \text{ m}^2 \tag{65}$$

In USC units:

$$A_{\text{T}} = \frac{\dot{m}}{3\,600 \rho \cdot v} = \frac{250\,000}{3\,600 \times 0{,}1 \times 500} = 1{,}39 \text{ ft}^2 \tag{66}$$

Thus, the pipe diameter should be approximately DN 400 (NPS 16).

7.3.4.2 Design details

Once the vent stack has been sized in accordance with the recommendations in 7.3.4.1 and the height has been established in accordance with the principles in 6.3, design is primarily a structural problem. If the vent stack is in a location remote from other facilities, the use of a guyed stack is usually as satisfactory as, and more economical than, providing a structure to support the stack. Vent stacks are frequently located in a process area that contains equipment connected to the stack. The stack can often be supported from a fractionating tower, chimney or other tall structure in the unit. Such an arrangement provides for economical discharge at a safe elevation.

The height of the vent stack is selected so that the concentration of vapour at a point of interest is well below the lower flammable limit of the vapour. Flammability consideration can be satisfied with 0,1 times to 0,5 times the lower flammable limit. Toxicity consideration can require much lower concentrations on certain applications and is, therefore, the controlling factor. Ambient dispersion calculations can be performed to confirm that concentrations at potential personnel locations are at acceptable exposure levels. The radiant heat intensity for vent stacks should also be checked in the event that a relieving vapour should ignite. This is done by the same means used for flare stacks, and the same limits apply for radiant heat intensity. Radiant heat levels sometimes take precedence over dispersion in determining stack height.

In every vent stack installation, careful consideration should be given to two potential problems:

a) accumulation of liquid in lines that terminate at the vent stack;

b) accidental ignition of the released vapour by lightning.

Accumulation of liquid in lines to the vent stack can result from leakage into the system of high-relative-molecular-mass vapours that condense at ambient temperature. If appreciable quantities of liquid collect, they will subsequently be discharged to the atmosphere when vapours are released into the system.

To avoid liquid accumulation, pockets should be prevented from occurring in the lines and the system should be sloped to a low-point drain. These drains can be installed to function automatically by using a properly designed seal. The height of the seal should provide a head equivalent to at least 1,75 times the back pressure under the maximum relief load to avoid release of vapour through the seal. As an alternative to a sealed drain, a small disengaging drum may be installed at the base of the vent stack. This type of installation is recommended where significant quantities of liquid can occur.

The possibility that vapours from the vent stack can be accidentally ignited by lightning or other sources usually makes a remote-controlled snuffing-steam connection desirable on the vent stack. This is especially true in locations where the incidence of lightning is high or where access to the point of discharge is difficult with conventional fire-extinguishing equipment. It is frequently impractical to size the steam-supply line for a rate that is sufficient to extinguish a fire under maximum venting conditions. However, steam is still essential, since, in most cases, vent fires occur when the only flow to the system consists of leakage or minor venting. Furthermore, unless steam is supplied, if ignition occurs when venting at or near the maximum design load, the fire will likely continue to burn when the cause of overpressure is corrected, with an accompanying reduction in venting.

7.3.4.3 Noise

7.3.4.3.1 The noise level at 30 m (100 ft) from the point of discharge to the atmosphere can be approximated by Equation (67):

$$L_{30(100)} = L + 10\,\lg\left(0,5\,q_{\mathrm{m}} \cdot c^2\right) \tag{67}$$

where

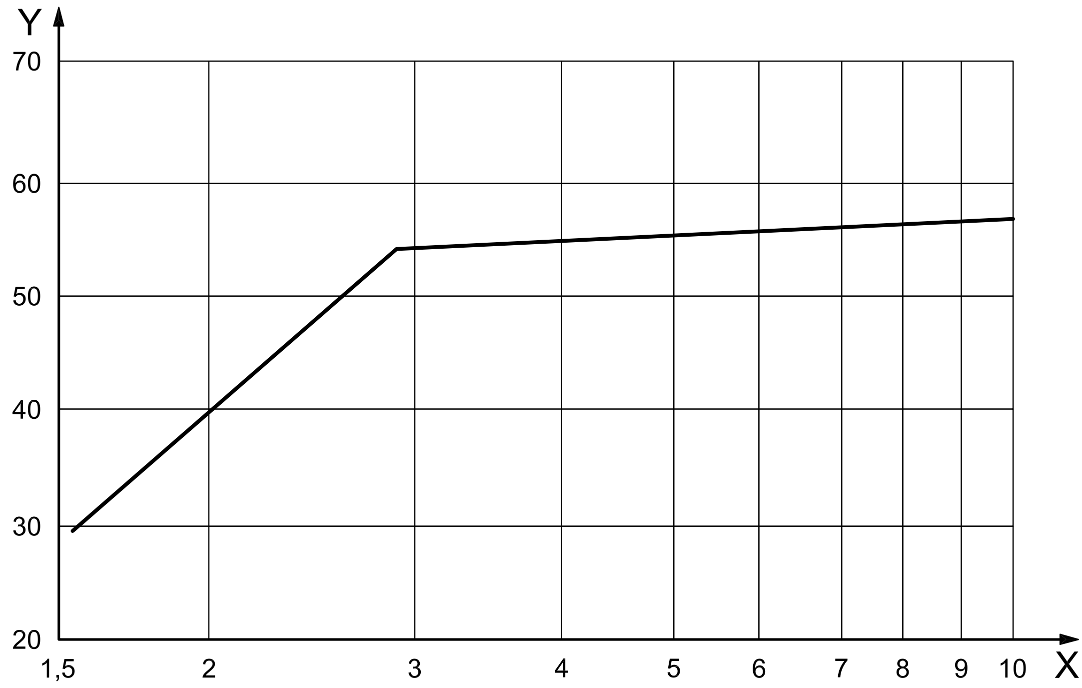
$L_{30(100)}$ is the noise level at 30 m (100 ft) from the point of discharge, expressed in decibels;

L is the noise level from Figure 18, expressed in decibels;

q_{m} is the mass flow through the valve, expressed in kilograms per second (pounds per second);

c is the speed of sound in the gas at the valve, expressed in metres per second (feet per second).

Figure 18 illustrates the noise intensity measured as the sound pressure level at 30 m (100 ft) from the stack tip versus the pressure ratio across the pressure-relief valve.



Key

X pressure ratio, PR

Y sound pressure level, $L_{30(100)}$, decibels

NOTE PR is the pressure ratio and is defined as the absolute static pressure upstream from the restriction (e.g. pressure-relief valve nozzle) divided by the absolute pressure downstream of the restriction while relieving. In some cases, critical flow can occur not only in the pressure-relief valve nozzle but also at the discharge-pipe outlet to atmosphere. In this case, the noise level is additive (logarithmic). In the case of the discharge pipe, the pressure ratio is the absolute pressure within the pipe at the outlet divided by atmospheric pressure.

Figure 18 — Sound pressure level at 30 m (100 ft) from the stack tip

Equations (68) and (69) show how to calculate the speed of sound, c.

In SI units:

$$c = 91{,}2\left(\frac{k \cdot T}{M}\right)^{0{,}5} \text{ m/s} \tag{68}$$

In USC units:

$$c = 223\left(\frac{k \cdot T}{M}\right)^{0{,}5} \text{ ft/s} \tag{69}$$

where

k is the ratio of the specific heats in the gas;

M is the relative molecular mass of the gas;

T is the gas temperature, expressed in kelvin (degrees Rankin).

7.3.4.3.2 The noise level at 30 m from the point of discharge to the atmosphere can be calculated in SI units as follows.

a) Calculate ($0{,}5\ q_\mathrm{m} \cdot c^2$) in watts.

b) Calculate $10 \times \lg(0{,}5\ q_\mathrm{m} \cdot c^2)$.

c) In Figure 18, find the value of PR on the abscissa and read the corresponding ordinate.

d) Add items b) and c) to obtain $L_{30(100)}$, which is the average sound pressure level at 30 m (100 ft), expressed in decibels.

Assume the following:

— $q_\mathrm{m} = 14{,}6$ kg/s;

— $k = 1{,}4$;

— $M = 29$;

— $T = 311$ K;

— $PR = 48/16 = 3$;

— $c = 91{,}2 \times \left(\dfrac{1{,}4 \times 311}{29} \right)^{0{,}5} = 353\ \text{m/s}$.

The results are as follows, referring to the numbered list items:

a) $(0{,}5\ q_\mathrm{m} \cdot c^2) = (0{,}5)(14{,}6)(353)^2 = 910\ 000$;

b) $10 \times \lg(0{,}5\ q_\mathrm{m} \cdot c^2) = 60$;

c) from Figure 18, the ordinate corresponding to $PR = 3$ is 54;

d) $L_{30(100)} = 54 + 60 = 114$ dB.

7.3.4.3.3 The noise level at 30 m from the point of discharge to the atmosphere can be calculated in SI units as follows:

a) Calculate ($0{,}5\ q_\mathrm{m} \cdot c^2$) in watts as follows: Divide the mass flow (pounds per second) by 32,2 foot-pounds per pound-force - square-second to obtain q_m. Multiply ($0{,}5\ q_\mathrm{m} \cdot c^2$) feet times pound-force per second by 1,36.

b) Calculate $10 \times \lg(0{,}5\ q_\mathrm{m} \cdot c^2)$.

c) In Figure 18, find the value of PR on the abscissa and read the corresponding ordinate.

d) Add items b) and c) to obtain $L_{30(100)}$, which is the average sound pressure level at 30 m (100 ft), expressed in decibels.

Assume the following:

— $q_\mathrm{m} = 32{,}2$ lb/s = 1 ft·lbf/s;

— $k = 1{,}4$;

— $M = 29$;

— $T = 560$ °R;

— $PR = 48/16 = 3$;

— $c = 223 \times \left(\frac{1{,}4 \times 560}{29} \right)^{0{,}5} = 1159 \text{ ft/s}$.

The results in USC units are as follows, referring to the list items above:

a) $(0{,}5\ q_m \cdot c^2) = (0{,}5)(1)(1\ 159)^2(1{,}36) = 910\ 000$;

b) $10 \times \lg(0{,}5\ q_m \cdot c^2) = 60$;

c) From Figure 18, the ordinate corresponding to $PR = 3$ is 54;

d) $L_{30(100)} = 54 + 60 = 114$ dB.

The above calculations are based on spherical spreading of the sound. If distances much larger than the height of the vent aboveground are of concern, add 3 dB to the calculated result to correct for hemispherical diffusion.

By applying Equations (70) and (71), the noise level can be adjusted for distances that differ from the 30 m (100 ft) reference boundary:

In SI units:

$$L_p = L_{30} - [20 \lg(r/30)] \tag{70}$$

In USC units:

$$L_p = L_{100} - [20 \lg(r/100)] \tag{71}$$

where

L_p is the sound pressure level at distance r, expressed in decibels;

$L_{30(100)}$ is the sound pressure level at 30 m (100 ft), expressed in decibels;

r is the distance from the sound source (stack tip), expressed in metres (feet).

For distances greater than 305 m (1 000 ft), some credit may be taken for molecular noise absorption. If pressure-relief valves prove to be excessively noisy during operation, the sound can be deadened by the application of insulation around the valve body and the downstream pipe up to approximately five pipe diameters from the valve.

7.4 Flare gas recovery systems

7.4.1 General

Environmental and economic considerations have resulted in the use of flare-gas recovery systems to capture and compress flare gases for other uses. Many times the recovered flare gas is treated and routed to the fuel-gas system. Depending upon flare gas composition, the recovered gas can have other uses.

7.4.2 Safety considerations

7.4.2.1 Path to flare

Flare systems are used for both normal process releases and emergency releases. Emergency streams, such as those from pressure-relief valves, depressuring systems, etc., shall always have flow paths to the flare available at all times. The design of flare-gas recovery systems shall not compromise this path. Several methods of accomplishing this are described in 7.4.3.3.

7.4.2.2 Back flow

Because flare-gas recovery systems usually involve compressors that take their suction directly from the flare header, the potential for back flow of air from the flare into the compressors at low flare-gas loads shall be considered. Typically, oxygen content of the flare gas stream should be measured and provisions shall be made to shut down the flare-gas compressors if potentially dangerous conditions exist.

7.4.2.3 Flare gas characteristics

Flare gases can have widely varying compositions that shall be evaluated during specification of recovery systems. The potential for materials that are not compatible with the flare-gas-treating systems or ultimate destinations shall be determined. For example, streams containing acid gases typically are routed directly to the flare, thereby bypassing the recovery system. Highly inert streams can also be incompatible with recovery systems.

7.4.3 Design considerations

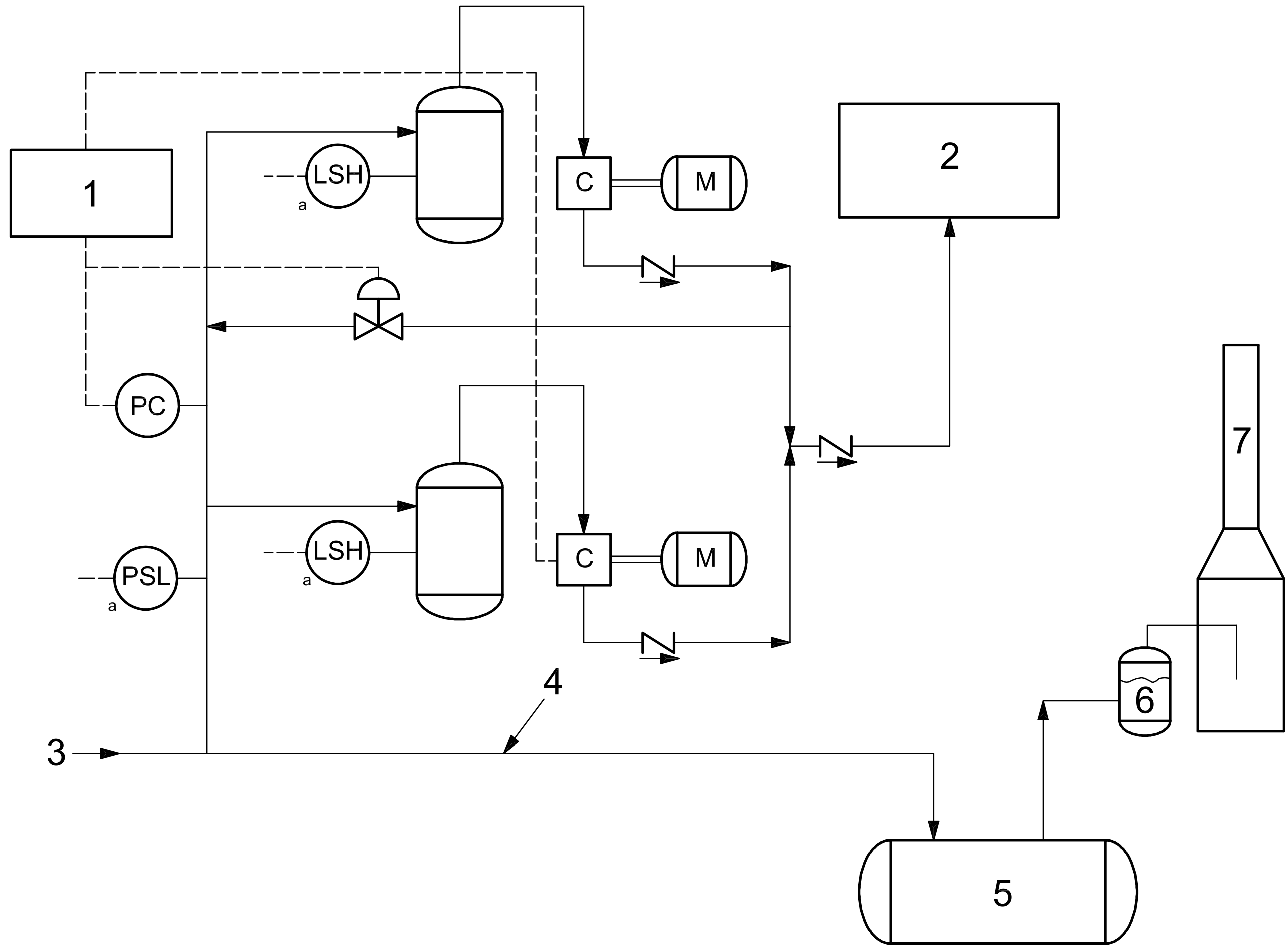
7.4.3.1 Sizing

Figure 19 shows a conceptual design for a flare-gas recovery system. Typically, the system consists of one or more reciprocating compressors whose suction is directly connected to the flare header. The compressed gas is usually routed to some type of treating system appropriate for the gas composition, then to fuel-gas or processing systems.

Flare-gas recovery systems are seldom sized for emergency flare loads. Usually, economics dictate that capacity be provided for some normal flare rate, above which gas is flared. Flare loads vary widely over time, and the normal rate can represent some average flare load, or a frequently encountered maximum load. Actual loads on these systems vary widely and they shall be designed to operate over a wide range of dynamically changing loads. Flare-gas recovery systems often are installed to comply with local regulatory limits on flare operation and, therefore, shall be sized to conform to any such limits.

7.4.3.2 Location

Typically, flare-gas recovery systems are located on the main flare header downstream of all unit header tie-ins and at a point where header pressure does not vary substantially with load. Locations upstream of process unit tie-ins should be carefully considered because of the potential for back flow and high oxygen concentrations. Limited downstream tie-ins for material not suitable for recovery can be required.



Key

1 compressor load control
2 flare gas treating
3 from process unit flare KO drums
4 flare header
5 flare KO drum (if used)
6 water seal
7 flare

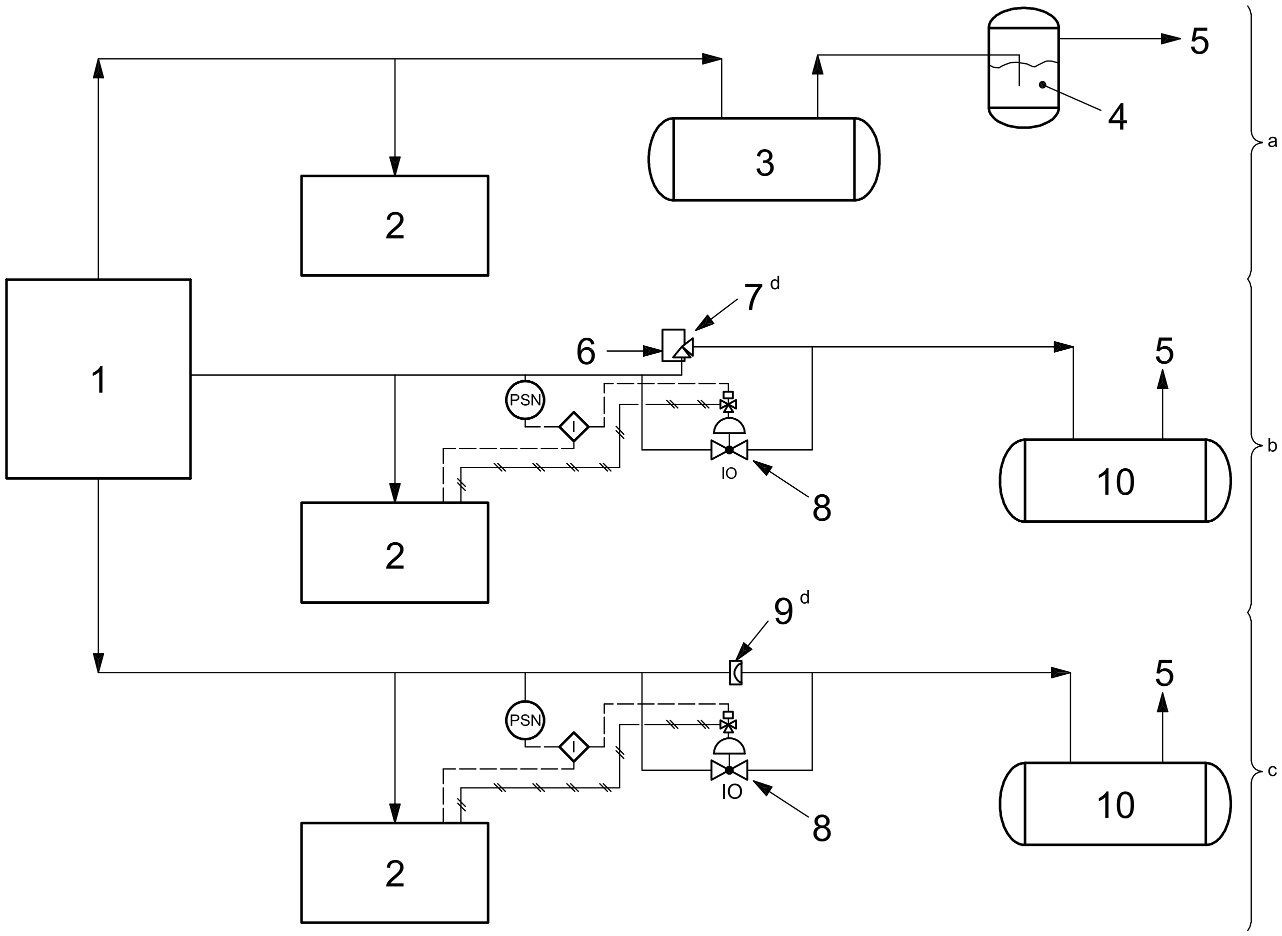
[a] Compressor shutdown.

Figure 19 —Typical flare gas recovery system

7.4.3.3 Flare tie-in

7.4.3.3.1 As discussed in 7.4.2, a major consideration in flare-recovery-system design is preservation of a path to the flare for emergency releases. The flare gas recovery system shall be designed as a side stream from the flare header. Main flare flow should not be through any compressor knockout or suction piping. The tie-in line to the flare gas recovery system should come off the top of the flare line to minimize the possibility of liquid entrance.

Some method of ensuring a positive pressure on the flare gas recovery system shall also be provided. Figure 20 shows some methods of doing this while preserving a reliable open path to the flare.



Key

1 from process units/flare knockout drums
2 flare-gas recovery
3 flare KO drum
4 seal pot
5 to flare
6 fuel-gas purge
7 pilot-operated relief valve
8 open valve on high pressure or compressor shutdown/unload
9 rupture disk or other non-reclosing pressure-relief device

[a] Preferred system: Water seal.
[b] Alternative system 1.
[c] Alternate system 2.
[d] See 7.4.3.3.3.

Figure 20 — Flare gas recovery inlet pressure

7.4.3.3.2 The most positive and preferred method for preventing air ingress is the installation of a water-seal vessel between the flare knockout drum and the flare itself. The seal provides a relatively constant, low back pressure on the flare header and provides a narrow, but usually adequate, control range for the flare-gas-recovery control system. The water seal should be designed to function over the pressure for which the flare-gas-recovery system is designed to operate. At higher release rates, flare gas flows through the seal and out the flare. Design provisions shall be made to maintain the seal level, prevent high flare rates from carrying the seal water up the flare and prevent seal freeze-up. See Figure D.1 for a typical seal-drum design.

7.4.3.3.3 If process requirements are such that the narrow operating ranges afforded by water seals cannot be accepted, an alternate method is to use a fail-open control valve to regulate the suction pressure of the flare-gas-recovery system. A positive path to the flare is provided by installing a low-pressure, high-capacity pilot-operated pressure-relief valve around the control valve. The sensing line for the pressure-relief valve pilot shall be provided with a clean gas purge and a backflow preventer.

The sizes of the control and pressure-relief valves can become quite large. The flare-header system shall also be studied to verify that the back pressure imposed by the pressure-relief device (assuming the control valve is closed) at full header load does not induce unacceptable back pressures on devices releasing into the headers at the processing units.

An alternative to the use of a pressure-relief valve is the installation of non-reclosing devices such as rupture disks or breaking-pin devices. These installations shall also be carefully reviewed to ensure that the devices operate when required to do so, at as low a pressure as possible, and that they do not cause unacceptable back pressure. An analysis should be performed to assure that these requirements are met.

7.4.3.3.4 If it is necessary to use a control valve in the flare line to regulate flare-gas-recovery-system suction pressure, the control valve should be of a fail-open design and be interlocked to go fully open upon a higher-than-normal header pressure, high-oxygen content or when the compressors are unloaded or shutdown. These interlocks are not a substitute for a positive path around the control valve, as described in 7.4.3.3.3.

7.4.3.4 Back flow protection

Provisions shall be made to prevent back flow of air from the flare into the flare-gas-recovery system. All compressors should be equipped with highly reliable low-suction-pressure shutdown controls. Consideration should also be given to installation of additional instrumentation in the section of header between the flare and the compressor suction take-off to detect reverse flow and automatically shut down the flare gas recovery system.

7.4.3.5 Flare-gas-recovery controls

7.4.3.5.1 Flare-gas-recovery systems operate over wide ranges, usually within very narrow suction pressure bands. A typical system can operate over a suction-pressure range of 0,5 kPa to 1,2 kPa (2 in H_2O to 5 in H_2O) to 2,5 kPa to 3 kPa (10 in H_2O to 12 in H_2O). The flare-gas-recovery compressors should be equipped with several stages of unloaders and a compressor-recycle valve. Suction pressure is maintained by pressure control of a recycle valve, with additional loading and unloading of the compressors when limits of valve opening or closing or suction pressure are reached. Usually, the controls are set up to sequentially load and unload the compressors.

7.4.3.5.2 The possibility of significant liquid in flare systems is usually quite high. Liquid-knockout vessels should be provided for the compressors with automatic shut down of the compressors on high suction-drum levels. Other mechanical protection systems can also be required for the compressors. These systems can either shut down or just unload the compressors. Refer to ISO 13707 [1] for guidance on compressor protection.

NOTE For the purpose of this provision, API Std 618 [11] is equivalent to ISO 13707.

Annex A
(informative)

Determination of fire relief requirements

A.1 Background

The problem of estimating fire-relief requirements for storage tanks was first recognized in 1928 when the NFPA requested API to recommend that a table of minimum emergency-relief loads for a series of tank capacities be included in the NFPA Suggested Ordinance Regulating the Use, Handling, Storage, and Sale of Flammable Liquids and the Products Thereof.

It was later recognized that tank capacities did not provide the best basis for estimating the amount of vapour to be handled. Since the heat was absorbed almost entirely by radiation, the area exposed, not the volume of the tank contents, seemed to be the important factor. Many of the tanks were large and were never expected to be entirely surrounded by fire; the assumption was, therefore, made that the larger the area of the container, the less the likelihood that the tank would be fully exposed to radiation. In other words, the larger the surface area of the tank shells, the lower the average unit heat absorption rate from a fire.

By 1948, several different equations [125] were in general use, prompting the API Subcommittee on Pressure Relieving Systems to develop an equation for determining the heat absorbed from open fires using the test data available at the time. The resultant equation has remained in general use since its publication in 1955 [126], and its development is documented in a paper presented by F. J. Heller in 1983 [127].

Table A.1 contains data from ten fire tests and one actual fire. These data result from tests in which means were provided to measure the total heat absorbed by a vessel by (a) computing the heat required to bring the liquid contents to the boiling range and (b) measuring the amount of liquid contents evaporated in a given time. The unit heat absorption rates in Table A.1 are average rates on the wetted surface.

Table A.1 — Comparison of heat absorption rates in fire tests

Test	Source	Type of exposure	Fuel	Vessel capacity		Total area		Wetted area		Total heat input		Temperature of surface		Heat input per unit wetted area		Ref. [b]
				m^3	(BBL) [a]	m^2	(ft^2)	m^2	(ft^2)	kW	(Btu/h)	°C	(°F)	kW/m^2	($Btu/h{\cdot}ft^2$)	
1	Hottel, average of 36 tests	6-inch thick metal stack	Gasoline	Conning tower		27	(296)	11	(123)	1 100	(3 760 000)	—	—	96	(30 500)	[144]
2	Hottel, average of 13 tests	6-inch thick metal stack	Gasoline	Conning tower		27	(296)	11	(123)	630	(2 139 000)	—	—	55	(17 400)	[145]
3	Standard Oil Company of California	Heating water in drum	Naphtha	0,41	(2,6)	—	—	2,4	(26)	120	(416 000)	—	—	50	(16 000)	[146]
4	Standard Oil Company of California	Heating water in tank	Naphtha	5,2	(33)	19	(206)	9,8	(105)	990	(3 370 000)	21 to 100	(70 to 212)	100	(32 000)	[146]
5	Underwriters Laboratories, Inc.	Water flowing over plate	Gasoline	—	—	2,2	(24)	2,2	(24)	230	(780 000)	24	(76)	100	(32 500)	[147]
6	Rubber Reserve Corporation test No. 17	Heating water in tank	Gasoline	18,9	(119)	37	(568)	37	(400)	2 700	(9 280 000)	150	(300)	73	(23 200)	[148]
7	Rubber Reserve Corporation test No. 17	Generating steam in tank	Gasoline	31,6	(199)	53	(568)	37	(400)	2 500	(8 400 000)	—	—	66	(21 000)	[148]
8	Rubber Reserve Corporation test No. 17	Water flowing in 3/4-inch standard pipe	Gasoline	—	—	0,84	(9,0)	0,8	(9,0)	80	(274 000)	—	—	96	(30 400)	[148]
9	API project test No. 1	Heating water in tank	Kerosene	0,14	(0,88)	1,5	(16,2)	0,6	(6,1)	28	(95 800)	150	(300)	50	(15 700)	[149]
10	API project test No. 2	Heating water in tank	Kerosene	0,14	(0,88)	1,5	(16,2)	0,6	(6,1)	30	(102 500)	160	(320)	53	(16 800)	[149]
11 [c]	Report to API on 38-ft. butane sphere	Plant fire	Butane	800	(5 000)	400	(4 363)	400	(4 363)	6 900	(23 560 000)	—	—	17	(5 400)	[150]

a BBL = barrels.

b Bibliographic reference number.

c This represents an actual fire.

Examinations of detailed reports on these tests indicate that the setup for tests 4, 5 and 8 was arranged to provide continuous and complete flame envelopment of the small vessels; under these conditions, maximum average heat input rates of 96 kW/m^2 to 100 kW/m^2 (30 400 Btu/h·ft^2 to 32 500 Btu/h·ft^2) were realized. The environmental conditions set up for tests 1, 3, 6, 7, 9 and 10 allowed the flame to be subjected to air currents and wind. All other factors were conducive to maintaining maximum heat input, a condition that is not likely to exist in a refinery. Under these conditions, the maximum average heat-input rates varied greatly. Test 2 differed from Test 1 in that drainage away from the equipment was provided. The maximum heat input rate is reduced by 60 % when drainage is provided; this fact was incorporated into the development of Equations (6) and (7). Test 11 gives an indication of the effect of a large area on average heat input during an actual fire.

The test reports mentioned that in some cases the tests were delayed until the arrival of a calm day so that the wind would not blow the flames away from the vessel. Copious supplies of fuel were available. In most cases, the fuel was maintained by dikes in a pool beneath the vessel and was not allowed to drain away as it normally would. In the Rubber Reserve Corporation tests, a 5 cm (2 in) gasoline line, running full, was required to keep the fuel supplied during the test. Without these special adverse conditions, the maximum heat absorption values obtained by these tests are extremely unlikely to occur in an actual refinery fire.

A.2 Nature of an open fire

The nature of an open fire of flammable fluid, as related to test data, is important. This kind of fire differs from the fire in the firebox of a boiler or heater, where the fuel and air are mixed by means other than the convection currents caused by the heated gases. The flame accordingly have a core of flammable vapour, either unmixed with air or insufficiently mixed to burn. Combustion occurs on the exterior envelope of this core. Because the actual combustion zone is on the rich side, a considerable amount of black smoke is generated. This envelope of soot can serve to mask much of the flame.

Hot gases from the combustion rise and the air that supports the combustion flows in at the bottom. The flame mass is quite turbulent; as masses of the burning vapour tumble and billow, the smoky mantle is displaced and the bright flame can be seen intermittently. This flame is not a blazing white, as it is in a furnace; it is red or orange, indicating a lower temperature than that of a furnace flame.

Flames of this type tend to rise because of their temperature; however, they can also be blown aside by the wind and can be blown so far from a vessel that the heating effect on the vessel is small.

A.3 Data on latent heat of vaporization of hydrocarbons

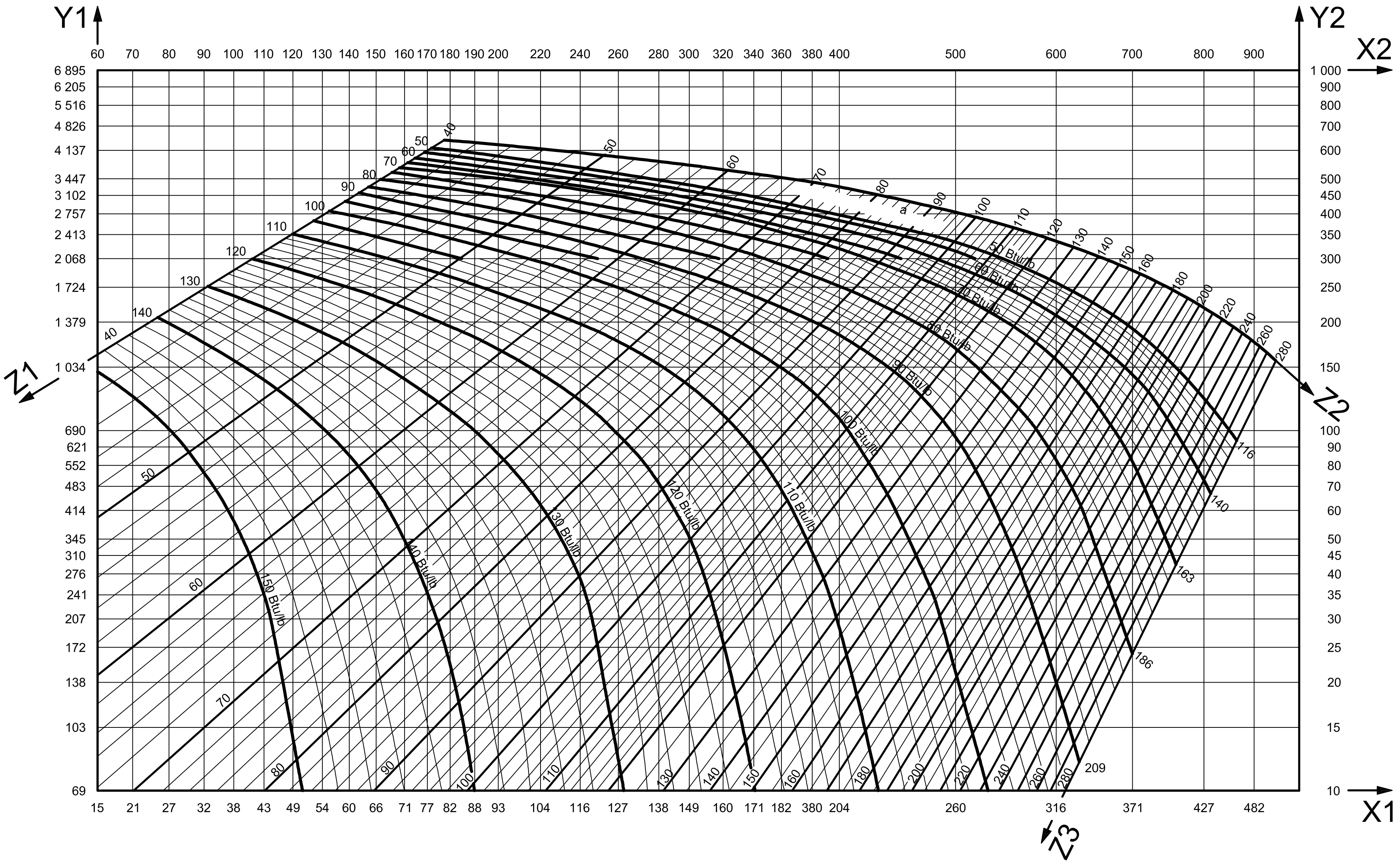
Different hydrocarbon liquids have different latent heats of vaporization even though hydrocarbons as a group behave similarly to one another. The latent heat of vaporization of a pure single-component liquid decreases as the temperature at vaporization increases and the becomes zero at the critical temperature and pressure for that liquid.

Figure A.1 shows the vapour pressures and latent heats of the pure, single-component paraffin hydrocarbon liquids. This chart is directly applicable to such liquids and applies as an approximation to paraffin-hydrocarbon mixtures composed of two components whose relative molecular masses vary no more than by the difference between propane and butane or butane and pentane.

The chart can also be applicable to isomer hydrocarbons, aromatic or cyclic compounds, or paraffin-hydrocarbon mixtures of components that have slightly different relative molecular masses. The equilibrium temperature should be calculated. Using the relationship for the calculated temperature versus vapour pressure, one can obtain the latent heat from Figure A.1. The relative-molecular-mass relationship as shown by the chart is not to be used in such cases; the relative molecular mass of the vapour should be determined from the vapour-liquid equilibrium calculation.

For cases that involve mixtures of components that have a wide boiling range or significantly different relative molecular masses, a rigorous series of equilibrium calculations can be required to estimate vapour generation rates, as discussed in 5.15.3.2.

Other recognized sources [128] of latent heat data or methods of calculating latent heat of vaporization should be used where Figure A.1 does not apply.



Key

X1 temperature, expressed in degrees Celsius X2 temperature, expressed in degrees Fahrenheit

Y1 vapour pressure, expressed in kPa (absolute) Y2 vapour pressure pounds per square inch (absolute)

Z1 latent heat of vaporization, expressed in British thermal units per pound Z2 latent heat of vaporization, expressed in kilojoules per kilogram Z3 relative molecular mass

a Pseudo-critical — do not use.

Figure A.1 — Vapour pressure and heat of vaporization of pure, single-component paraffin hydrocarbon liquids

Annex B
(informative)

Special system design considerations

B.1 Single pressure-relief device protecting several components in a process system

In some situations, a single pressure-relief device can be desirable to protect several equipment components in a process system. For this system to be adequately designed, the following four criteria should be satisfied.

a) No means can exist for blocking any of the equipment components being protected from the installation of the single pressure-relief device unless closure of these valves is positively controlled (e.g. see API RP 520-II:2003, Section-4 or EN 764-7).

b) The set pressure of the first pressure-relief device to be actuated should be at or below the lowest design pressure or MAWP of any equipment component being protected in the system.

c) The accumulated pressure when one or more pressure-relief devices are discharging shall not exceed the maximum allowable pressure in accordance with the pressure-design code.

d) The operating pressure in any equipment component shall not exceed that allowed by the pressure-design code when the pressure-relief device protecting the system is not discharging.

B.2 Description of a typical process system

A typical process system that can be provided with only one pressure-relief device is a hydrotreater-reactor recycle-gas loop. Such a system can contain the following main equipment components:

a) recycle gas compressor;

b) feed/effluent heat exchanger;

c) fired heater;

d) reactor;

e) effluent condenser;

f) separator drum;

g) interconnecting piping;

h) piping for liquid feed, product, and purge gas.

Figure B.1 is a schematic for the typical hydrotreating process system indicated in the preceding list of components.

B.3 Procedure to calculate the design pressure or MAWP of equipment components

If the procedure outlined in the following list is followed, the design pressure or MAWP of any equipment component in the system will never be exceeded unless the pressure in the system actuates the pressure-relief device. These steps should be taken.

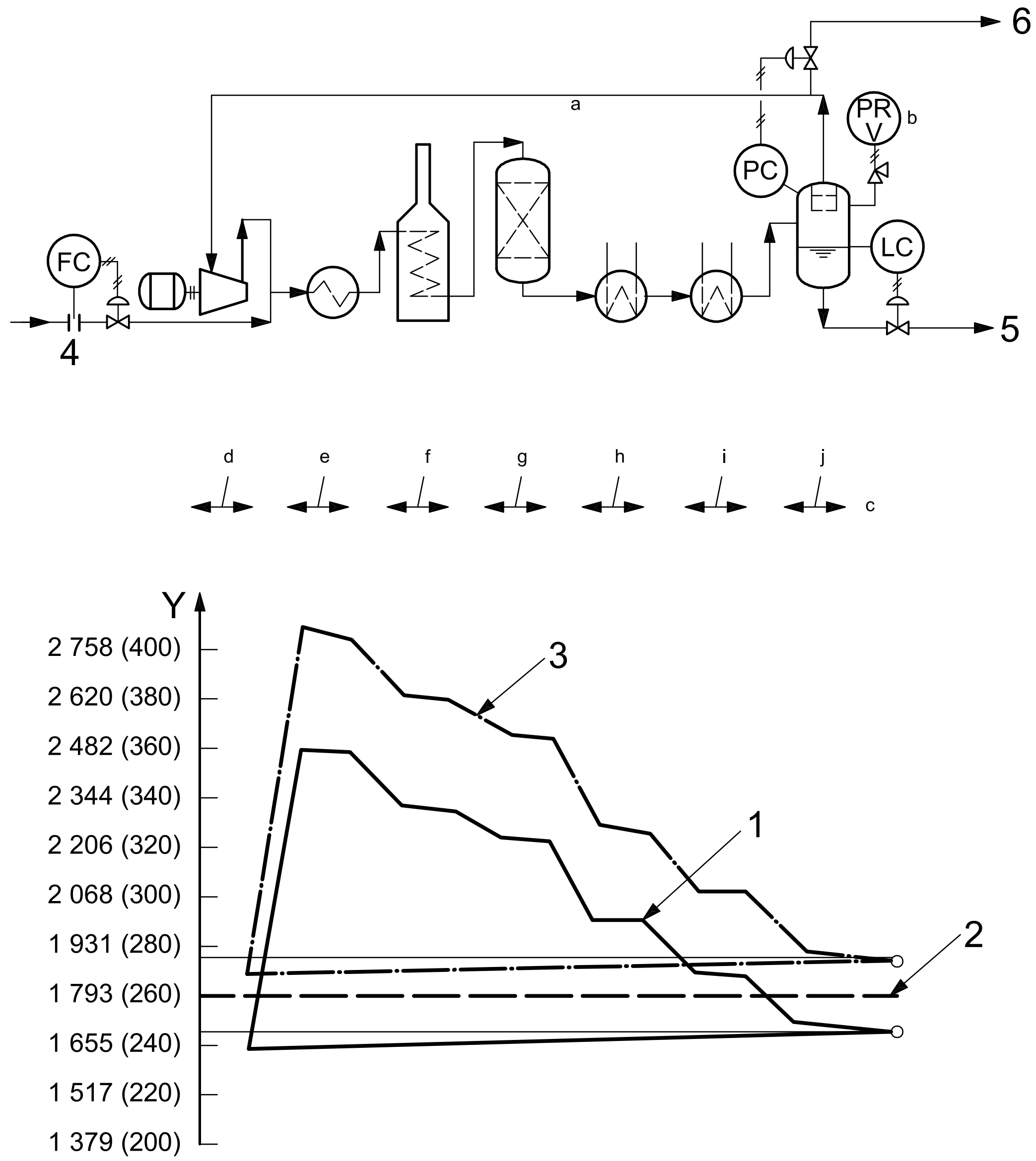
a) The pressure profile should be developed for the processing conditions that result in the maximum pressure drop (normal end-of-run conditions with fouled equipment).

b) The settling-out pressure that develops when the compressor stops during the maximum pressure drop case should be calculated. The separator drum should be assumed to be operating at normal operating pressure before compressor stoppage and the purge gas line should be assumed to be closed to conserve gas.

c) The minimum design pressure or MAWP of the separator drum should be calculated as 1,05 times the settling-out pressure. This provides an adequate differential between the operating pressure and set pressure of the pressure-relief device for a compressor shutdown contingency.

d) The pressure profile should be developed for the system with the pressure of the separator drum at the set pressure of the pressure-relief device. Assuming equal volumetric gas flow, the pressure gradient is proportional to the change in absolute pressure.

NOTE The minimum design pressure or MAWP of each equipment component is the inlet pressure for each equipment component as determined in list item d).



Key

Y pressure, expressed in kilopascals (pounds per square inch gauge)

1 pressure profile for normal operation based on end-of-run (EOR) process condition [see B.3, list item a)]

2 settling-out pressure when system pressure equalizes after compressor stops during normal EOR operation [see B.3, list item b)]

3 pressure profile when system is operating with the pressure of the HP separator drum at the set pressure of the pressure-relief device [see B.3, list item d)]

4 feed

5 liquid product

6 gas bleed-off

[a] Normally set at 1 689 kPa (245 psig).

[b] Set at 1 896 kPa (275 psig).

[c] Typical minimum design pressures for equipment components in the system.

[d] Compressor: DP = 2 799 kPa (406 psi) min.

[e] Feed/Product exchange: DP = 2 779 kPa (403 psi) min.

[f] Furnace: DP = 2 613 kPa (379 psi) min.

[g] Reactor: DP = 2 517 kPa (365 psi) min.

[h] Feed/product exchange: DP = 2 234 kPa (324 psi) min.

[i] Product condensorr: DP = 2 062 kPa (299 psi) min.

[j] H.P.separator: DP = 1 896 kPa (275 psi) min.

Figure B.1 — Typical flow scheme of a system involving a single pressure-relief device serving components in a process system with typical pressure profiles

Annex C
(informative)

Sample calculations for sizing a subsonic flare stack

C.1 General

This annex presents examples of the two methods used to size subsonic flare stacks based on the effects of radiation. The first method covered is the simple approach presented in Clause 6; the second is the more specific approach using Brzustowski's and Sommer's method [94]. The height and location of the flare stack should be considered, based on gas dispersion if the flame is extinguished (see 6.3).

C.2 Example 1 — Sizing a flare stack using the simple approach

C.2.1 Basic data

In this example, the material flowing is hydrocarbon vapours. The mass flow rate, q_{m}, is 45 360 kg/h (100 000 lb/h). The average relative molecular mass of the vapours, M, is 46,1. The flowing temperature, T, is 422 K (760 °R). The compressibility factor, Z, is 1,0. The heat of combustion is 50 000 kJ/kg (21 500 Btu/lb). The absolute pressure within the flare tip while flaring, p_2, is 101,3 kPa (14,7 psi). The design wind velocity (u_∞) is 32,2 km/h (8,9 m/s) [20 mph (29,3 ft/s)].

C.2.2 Calculation of flare diameter

The Mach number is determined from Equation (27) or (28) from 7.3.1.3.3:

In SI units:

$$Ma_2 = 3{,}23 \times 10^{-5} \left(\frac{q_{\mathrm{m}}}{p_2 \cdot d^2} \right) \left(\frac{Z \cdot T}{M} \right)^{0{,}5} \qquad (27)$$

In USC units:

$$Ma_2 = 1{,}702 \times 10^{-5} \left(\frac{q_{\mathrm{m}}}{p_2 \cdot d^2} \right) \left(\frac{Z \cdot T}{M} \right)^{0{,}5} \qquad (28)$$

For $Ma_2 = 0{,}2$, the flare diameter is calculated as follows:

In SI units:

$$0{,}2 = 3{,}23 \times 10^{-5} \left(\frac{45\,360}{101{,}3 d^2} \right) \sqrt{\frac{1 \times 422}{46{,}1}}$$

$$d^2 = 0{,}219$$

$$d = 0{,}468 \text{ m (inside diameter)}$$

In USC units:

$$0{,}2 = 1{,}702 \times 10^{-5} \left(\frac{100\ 000}{14{,}7 d^2} \right) \sqrt{\frac{1 \times 760}{46{,}1}}$$

$$d^2 = 2{,}35$$

$d = 1{,}53$ ft (inside diameter)

For $Ma = 0{,}5$, the flare diameter is calculated as follows:

In SI units:

$$d^2 = 0{,}088$$

$d = 0{,}296$ m (inside diameter)

In USC units:

$$d^2 = 0{,}94$$

$d = 0{,}97$ ft (inside diameter)

C.2.3 Calculation of flame length

The heat liberated, Q, is calculated as follows (see Figures 7 and 8):

In SI units:

$Q = (45\ 360$ kg/h$) \times (5\text{x}10^4$ kJ/kg$) \times (1$ h/3600 s$)$

$= 6{,}3 \times 10^5$ kW

In USC units:

$Q = (100\ 000) \times (21\ 500)$

$= 2{,}15 \times 10^9$ Btu/h

From Figures 7 and 8, flame length, L, is 50 m (170 ft). See Figure C.1.

C.2.4 Simple calculation of flame distortion caused by wind velocity

The vapour volume flow rate, q_{vap}, is determined as follows:

In SI units:

$$q_{vap} = \left(\frac{45\ 360}{3\ 600} \right) \times \left(\frac{22{,}4}{46{,}1} \right) \times \left(\frac{422}{273} \right) = 9{,}46 \text{ m}^3\text{/s (actual)}$$

In USC units:

$$q_{vap} = \left(\frac{100\ 000}{3\ 600} \right) \times \left(\frac{379{,}1}{46{,}1} \right) \times \left(\frac{760}{520} \right) = 333{,}9 \text{ ft}^3\text{/s (actual)}$$

The flame distortion caused by wind velocity (see Figure 9) can be represented by Equation (C.1):

$$\frac{u_\infty}{u_\mathrm{j}} \qquad \text{(C.1)}$$

where

u_∞ is the wind velocity;

u_j is the flare tip velocity.

The flare tip exit velocity, u_j, can be determined from Equation (C.2) (see C.3.3 for another method of calculating u_j):

$$u_\mathrm{j} = \frac{q}{\pi d^2 / 4} \qquad \text{(C.2)}$$

For $Ma = 0{,}2$:

In SI units:

$$u_\mathrm{j} = \frac{9{,}46}{\pi \times 0{,}468^2 / 4} = 55 \text{ m/s}$$

In USC units:

$$u_j = \frac{333{,}9}{\pi \times 1{,}53^2 / 4} = 181 \text{ ft/s}$$

For $Ma = 0{,}5$:

In SI units:

$$u_j = \frac{9{,}46}{\pi \times 0{,}296^2 / 4} = 137 \text{ m/s}$$

In USC units:

$$u_j = \frac{333{,}9}{\pi \times 0{,}97^2 / 4} = 452 \text{ ft/s}$$

At $Ma = 0{,}2$:

In SI units:

$$\frac{u_\infty}{u_\mathrm{j}} = \frac{8{,}94}{55} = 0{,}162$$

From Figure 9: $\sum \frac{\Delta y}{L} = 0{,}36$

From Figure 9: $\sum \frac{\Delta x}{L} = 0{,}85$

$$\sum \Delta y = 0{,}36 \times 50 = 18 \text{ m}$$

$$\sum \Delta x = 0{,}85 \times 50 = 42{,}5 \text{ m}$$

In USC units:

$$\frac{u_\infty}{u_j} = \frac{29{,}3}{181} = 0{,}162$$

From Figure 9: $\sum \frac{\Delta y}{L} = 0{,}36$

From Figure 9: $\sum \frac{\Delta x}{L} = 0{,}85$

$$\sum \Delta y = 0{,}36 \times 170 = 61 \text{ ft}$$

$$\sum \Delta x = 0{,}85 \times 170 = 144 \text{ ft}$$

At $Ma = 0{,}5$:

In SI units:

$$\frac{u_\infty}{u_j} = \frac{8{,}94}{137} = 0{,}065$$

From Figure 9: $\sum \frac{\Delta y}{L} = 0{,}55$

From Figure 9: $\sum \frac{\Delta x}{L} = 0{,}68$

$$\sum \Delta y = 0{,}55 \times 50 = 27{,}5 \text{ m}$$

$$\sum \Delta x = 0{,}68 \times 50 = 34{,}0 \text{ m}$$

In USC units:

$$\frac{u_\infty}{u_j} = \frac{29{,}3}{452} = 0{,}065$$

From Figure 9: $\sum \frac{\Delta y}{L} = 0{,}55$

From Figure 9: $\sum \frac{\Delta x}{L} = 0{,}68$

$$\sum \Delta y = 0{,}55 \times 170 = 93 \text{ ft}$$

$$\sum \Delta x = 0{,}68 \times 170 = 115 \text{ ft}$$

C.2.5 Calculation of required flare stack height

For the basis of the calculations used in C.2.5, see 6.4.2.3. See Figure C.1 for dimensional references.

The design basis for these calculations is as follows.

The fraction of heat radiated, F, is 0,3. The heat liberated (see C.2.3), Q, is $6{,}3 \times 10^5$ kW ($2{,}15 \times 10^9$ Btu/h). Assume it is necessary for the flare stack design to limit the maximum allowable radiation, K, at 45,7 m (150 ft) from the flare stack to 6,3 kW/m^2 (2 000 Btu/h·ft^2).

In Equation (24) in 6.4.2.3.3, the value of τ should be assumed to be 1,0. The distance from the flame centre to the grade-level boundary (that is, the object being considered), D, is then calculated according to Equation (24):

$$D = \sqrt{\frac{\tau \cdot F \cdot Q}{4\pi \cdot K}} \tag{24}$$

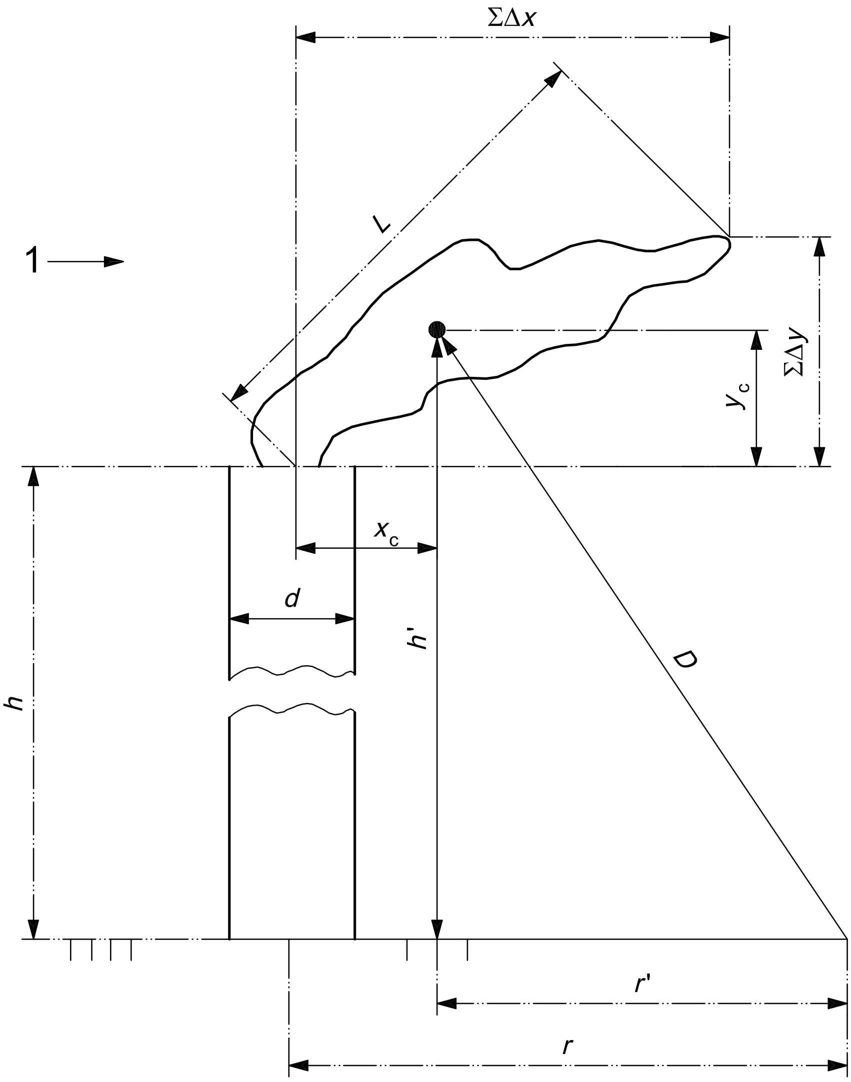
In SI units:

$$D = \sqrt{\frac{1{,}0 \times 0{,}3 \times 6{,}3 \times 10^5}{4\,\pi \times 6{,}3}} = 48{,}9 \text{ m}$$

In USC units:

$$D = \sqrt{\frac{1{,}0 \times 0{,}3 \times 2{,}15 \times 10^9}{4\,\pi \times 2000}} = 160 \text{ ft}$$

The physical arrangement shown in Figure C.1 is the basis of the remaining calculations in C.2.5.



Key

1 wind direction

Figure C.1 — Dimensional references for sizing a flare stack

At $Ma = 0{,}2$, the flare stack height, h, is calculated as follows:

$$h' = h + (0{,}5 \sum \Delta y) \tag{C.3}$$

$$r' = r - (0{,}5 \sum \Delta x) \tag{C.4}$$

In SI units:

From C.2.4, at $Ma = 0{,}2$: $\sum \Delta y = 18\,\text{m}$

From C.2.4, at $Ma = 0{,}2$: $\sum \Delta x = 42{,}5\,\text{m}$

Based on a grade level distance from the flare, r, of 45,7 m:

$r' = 45{,}7 - (0{,}5 \times 42{,}5) = 24{,}4$ m

and:

$D^2 = r'^2 + h'^2$

$48{,}9^2 = 24{,}4^2 + h'^2$

$h' = 42{,}3$ m

$h = 2{,}3 - (0{,}5 \times 18) = 33{,}3$ m

In USC units:

From C.2.4, at $Ma = 0{,}2$: $\sum \Delta y = 61\,\text{ft}$

From C.2.4, at $Ma = 0{,}2$: $\sum \Delta x = 144$ ft

Based on a grade level distance from the flare, r, of 150 ft:

$r' = 50 - (0{,}5 \times 144) = 78$ ft

and:

$D^2 = r'^2 + h'^2$

$160^2 = 78^2 + h'^2$

$h' = 140$ ft

$h = 140 - (0{,}5 \times 61) = 110$ ft

At $Ma = 0{,}5$, h is calculated as follows:

In SI units:

From C.2.4, at $Ma = 0{,}5$: $\sum \Delta y = 27{,}5\,\text{m}$

From C.2.4, at $Ma = 0{,}5$: $\sum \Delta x = 34{,}0\,\text{m}$

Based on a grade level distance from the flare, r, of 45,7 m:

$r' = 45{,}7 - (0{,}5 \times 34) = 28{,}7$ m

and

$D^2 = r'^2 + h'^2$

$48{,}9^2 = 28{,}7^2 + h'^2$

$h' = 39{,}6$ m

$h = 39{,}6 - (0{,}5 \times 27{,}5) = 25{,}9$ m

In USC units:

From C.2.4, at $Ma = 0{,}5$: $\sum \Delta y = 93$ ft

From C.2.4, at $Ma = 0{,}5$: $\sum \Delta x = 115$ ft

Based on a grade level distance from the flare, r, of 150 ft:

$r' = 150 - (0{,}5 \times 115) = 92{,}5$ ft

and:

$D^2 = r'^2 + h'^2$

$160^2 = 92{,}5^2 + h'^2$

$h' = 131$ ft

$h = 131 - (0{,}5 \times 93) = 85$ ft

C.3 Example 2 — Sizing a flare using Brzustowski's and Sommer's approach

C.3.1 Basic data

In this example, based on Brzustowski's and Sommer's method [94], the material flowing is hydrocarbon vapours. The flow rate, q_m, is 126 kg/s (1 000 000 lb/h). The relative molecular mass of the flare gas, M_j, is 46,1, and the relative molecular mass of air, M_∞, is 29. The normal average wind speed, u_∞, is 32,2 km/h (8,9 m/s) [20 mph (29,3 ft/s)]. The velocity of the flare gas at the flare tip, u_j, is measured in m/s (ft/s). The inside diameter of the flare tip, d_j, is measured in metres (feet). The absolute pressure within the flare tip while flaring, p_j, is 108 kPa (15,7 psi). The average relative humidity, R_H, is 50 %. The heat of combustion is 50 000 kJ/kg (21 500 Btu/lb). The compressibility factor, Z, is 1,0. The lower-explosive-limit concentration of the flare gas in air, C_L, measured as a volume fraction, is 0,021 (see C.3.6.1). The absolute temperature of the flare gas, T_j, is 422 K (760 °R), and temperature of the air, T_∞, is 289 K (520 °R).

The fraction by which the flame radiation is reduced when transmitted through the atmosphere is indicated by τ. The fraction of heat radiated is indicated by F. The heat release, Q, is measured in kW (Btu/h), and the allowable radiation intensity, K, is measured in kW/m^2 (Btu/h·ft^2).

C.3.2 Calculation of flare diameter

The Mach number is determined as follows (see 7.3.1.3.3):

In SI units:

$$Ma_2 = 3{,}23 \times 10^{-5} \left(\frac{q_m}{p_2 \cdot d^2} \right) \left(\frac{Z \cdot T}{M} \right)^{0{,}5} \tag{27}$$

In USC units:

$$Ma_2 = 1{,}702 \times 10^{-5} \left(\frac{q_m}{p_2 \cdot d^2} \right) \left(\frac{Z \cdot T}{M} \right)^{0{,}5} \tag{28}$$

For $Ma = 0{,}5$, the flare diameter is calculated as follows:

In SI units:

$$0{,}5 = 3{,}23 \times 10^{-5}\left(\frac{453\,600}{108 d_{\mathrm{j}}^{2}}\right)\left(\frac{1 \times 422}{46{,}1}\right)^{0{,}5}$$

$d_{\mathrm{j}}^{2} = 0{,}82$

$d_{\mathrm{j}} = 0{,}91$ m

In USC units:

$$0{,}5 = 1{,}702 \times 10^{-5}\left(\frac{1\,000\,000}{15{,}7 d_{j}^{2}}\right)\left(\frac{1 \times 760)}{46{,}1}\right)^{0{,}5}$$

$d_{\mathrm{j}}^{2} = {,}80$

$d_{\mathrm{j}} = 2{,}97$ ft

C.3.3 Location of flame centre

The tip exit velocity, u_{j}, is calculated as follows:

In SI units:

$$\text{isothermal sonic velocity} = 91{,}2\ (T_{\mathrm{j}}/M_{\mathrm{j}})^{0{,}5} \tag{C.5}$$

$$= 91{,}2 \times (422/46{,}1)^{0{,}5} = 276 \text{ m/s}$$

u_{j} = jet Mach number times sonic velocity

$= (0{,}5) \times (276) = 138$ m/s

In USC units:

$$\text{isothermal sonic velocity} = 223\ (T_{\mathrm{j}}/M_{\mathrm{j}})^{0{,}5} \tag{C.6}$$

$$= 223 \times (760/46{,}1)^{0{,}5} = 905 \text{ ft/s}$$

u_{j} = jet Mach number times sonic velocity

$= (0{,}5) \times (905) = 453$ ft/s

The lower-explosive-limit concentration parameter for the flare gas, $\overline{C_L}$, is calculated as follows:

$$\overline{C_L} = C_{\mathrm{L}}\left(\frac{u_{\mathrm{j}}}{u_{\infty}}\right)\left(\frac{M_{\mathrm{j}}}{M_{\infty}}\right) \tag{C.7}$$

In SI units:

$$\overline{C_L} = 0{,}021 \times \left(\frac{138}{8{,}9}\right) \times \left(\frac{46{,}1}{29}\right) = 0{,}517$$

In USC units:

$$\overline{C_L} = 0{,}021 \times \left(\frac{453}{29{,}3}\right) \times \left(\frac{46{,}1}{29}\right) = 0{,}516$$

The parameter for jet thrust and wind thrust, ($d_j \cdot R$), is calculated as follows (see C.3.6.2):

$$d_j \cdot R = d_j \left(\frac{u_j}{u_\infty}\right) \left(\frac{T_\infty \cdot M_j}{T_j}\right)^{0{,}5} \tag{C.8}$$

In SI units:

$d_j \cdot R = 0{,}91 \times 138/8{,}9 \times (289 \times 46{,}1/422)^{0{,}5} = 79{,}3$

In USC units:

$d_j \cdot R = 2{,}97 \times 453/29{,}3 \times (520 \times 46{,}1/760)^{0{,}5} = 258$

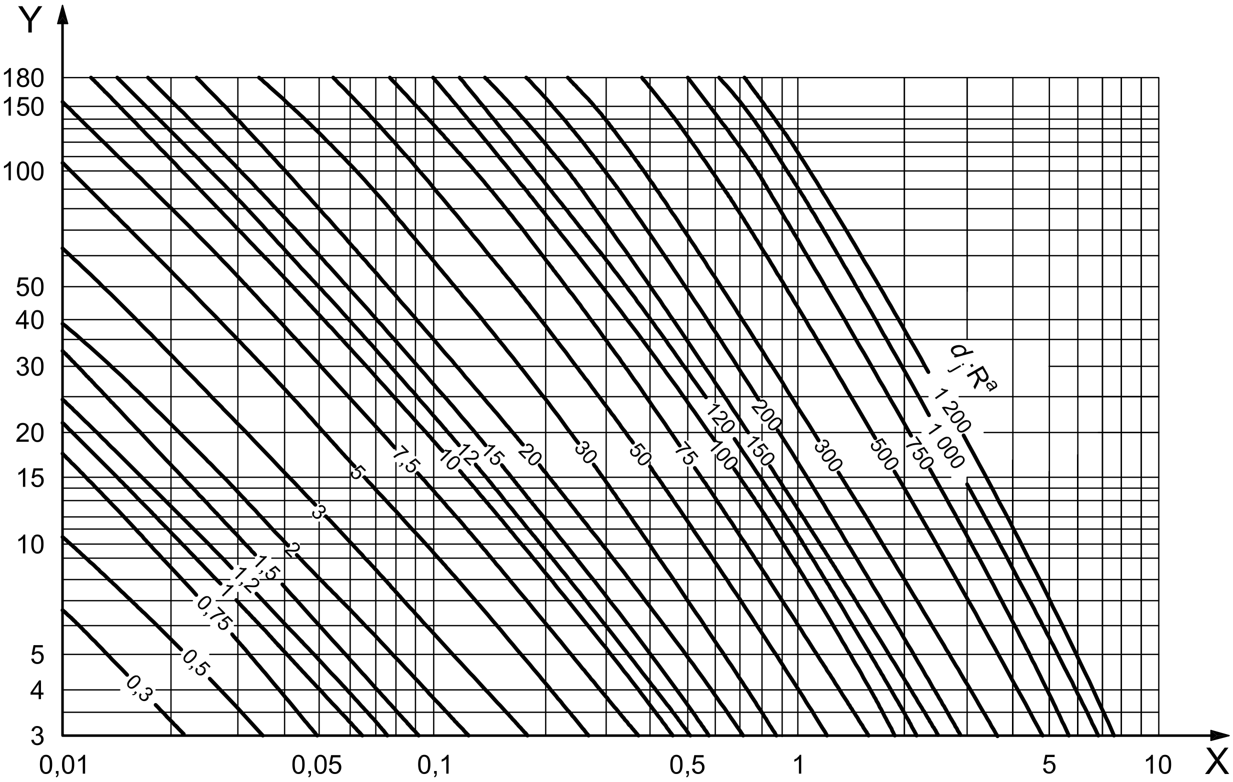
The horizontal and vertical distances from the flare tip to the flame centre, x_c and y_c, respectively, are determined as follows:

From Figure C.2, at $\overline{C_L} = 0{,}517$ and $(d_j \cdot R) = 79{,}3$: $x_c = 18$ m

From Figure C.3, at $\overline{C_L} = 0{,}516$ and $(d_j \cdot R) = 258$: $x_c = 58$ ft

From Figure C.4, at $\overline{C_L} = 0{,}517$ and $(d_j \cdot R) = 79{,}3$: $y_c = 30$ m

From Figure C.5, at $\overline{C_L} = 0{,}516$ and $(d_j \cdot R) = 258$: $y_c = 100$ ft



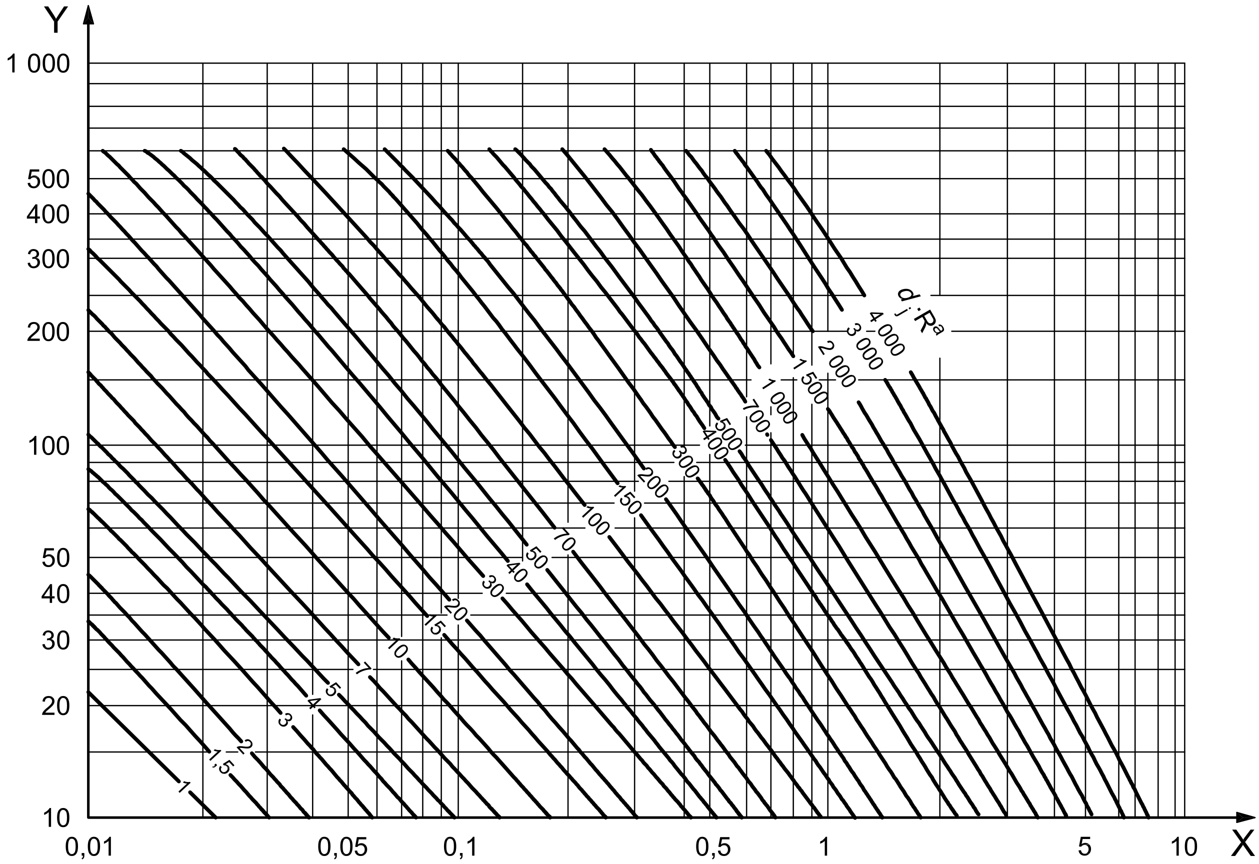
Key

X $\overline{C_L}$, the lower-explosive-limit concentration parameter for the flare gas, see Equation (C.7)

Y x_c, horizontal distance from the stack to flame centre, expressed in metres

[a] $(d_j \cdot R)$ is the parameter for jet thrust and wind thrust, see Equation (C.8).

Figure C.2 — Flame centre for flares and ignited vents — Horizontal distance, x_c (SI units)



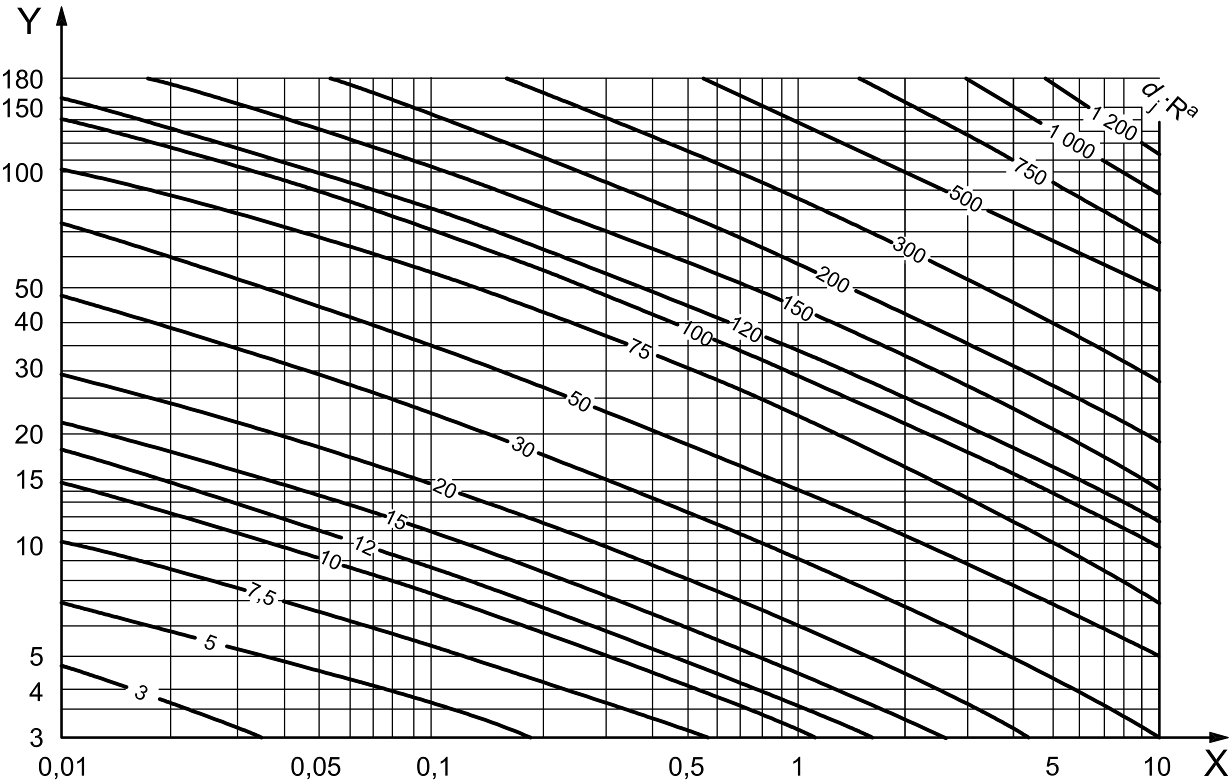
Key

X $\overline{C_L}$, the lower-explosive-limit concentration parameter for the flare gas, see Equation (C.7)

Y x_c, horizontal distance from the stack to flame centre, expressed in feet

[a] $(d_j \cdot R)$ is the parameter for jet thrust and wind thrust, see Equation (C.8).

Figure C.3 — Flame centre for flares and ignited vents — Horizontal distance, x_c (USC units)



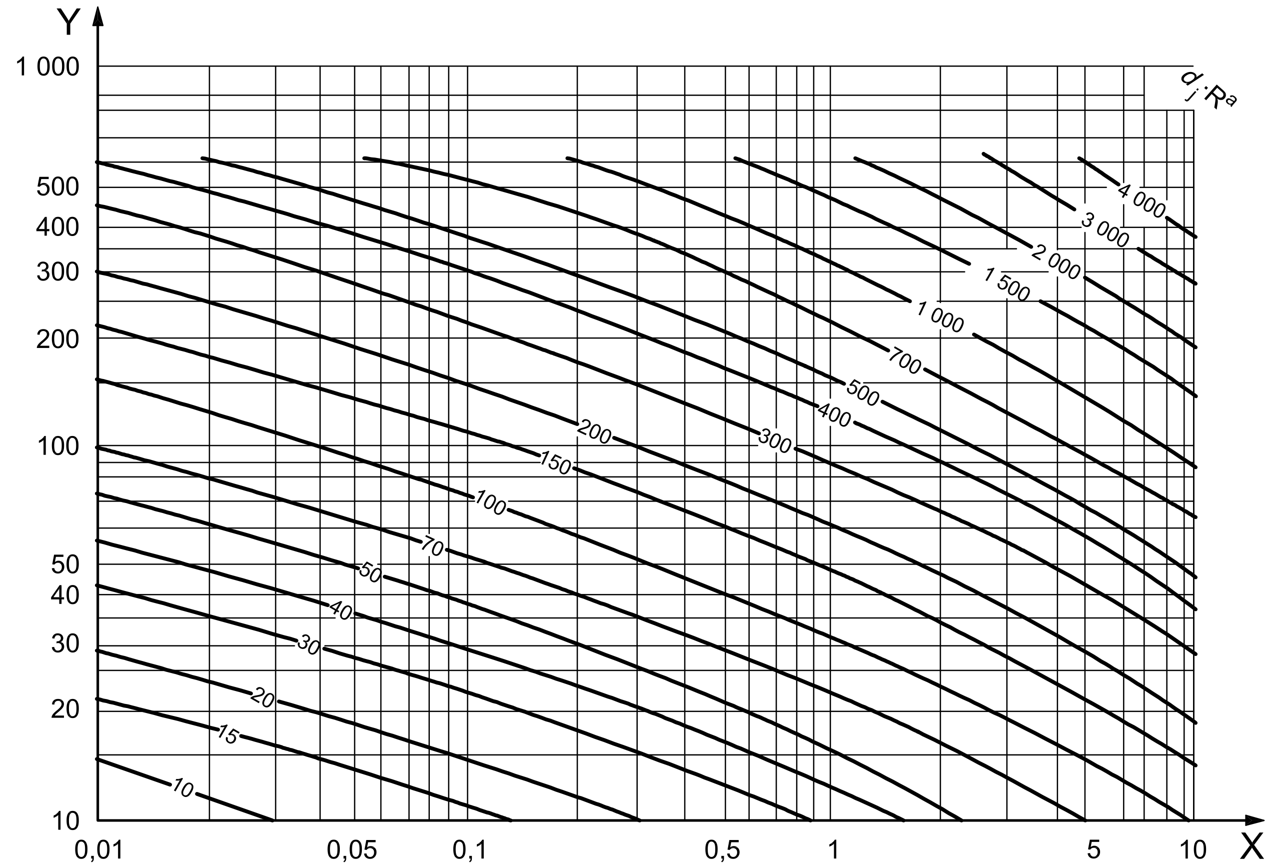
Key

X $\overline{C_L}$, the lower-explosive-limit concentration parameter for the flare gas, see Equation (C.7)

Y y_c, vertical distance from the stack to flame centre, expressed in metres

a ($d_j \cdot R$) is the parameter for jet thrust and wind thrust, see Equation (C.8).

Figure C.4 — Flame centre for flares and ignited vents — Vertical distance, y_c (SI units)



Key

X $\overline{C_L}$, the lower-explosive-limit concentration parameter for the flare gas, see Equation (C.7)

Y y_c, vertical distance from the stack to flame centre, expressed in feet

[a] ($d_j \cdot R$) is the parameter for jet thrust and wind thrust, see Equation (C.8).

Figure C.5 — Flame centre for flares and ignited vents — Vertical distance, y_c (USC units)

C.3.4 Calculation of the distance from the flame centre to the object or point being considered

The design basis for this calculation is as follows: The fraction of heat radiated, F, is 0,3. The heat liberated (see C.2.3), Q, is $6,3 \times 10^6$ kW ($2,15 \times 10^{10}$ Btu/h). Say the flare stack design must limit the maximum allowable radiation (see 6.4.2.3), K, is 9,5 kW/m^2 (3 000 Btu/h·ft^2).

In Equation (24), the value of τ should be assumed to be 1,0 (see C.3.6.3 and C.3.6.4). The distance from the flame centre to the object or point being considered (that is, the distance to the limit of the radiant heat intensity, such as grade level, an equipment platform, or a plant boundary), D, is then calculated as follows:

$$D = \sqrt{\frac{\tau \cdot F \cdot Q}{4\pi \cdot K}} \quad (24)$$

In SI units:

$$D = \sqrt{\frac{1,0 \times 0,3 \times 6,3 \times 10^6}{4\pi \times 9,5}} = 126 \text{ m}$$

In USC units:

$$D = \sqrt{\frac{1{,}0 \times 0{,}3 \times 2{,}15 \times 10^{10}}{4\pi \times 3\ 000}} = 413 \text{ ft}$$

Thus, at a maximum allowable K of 9,5 kW/m^2 (3 000 Btu/h·ft^2), D = 126 m (413 ft) from the flame centre.

Similarly, if the maximum allowable K is 6,3 kW/m^2 (2 000 Btu/h·ft^2), D = 154 m (507 ft) from the flame centre.

C.3.5 Determination of flare stack height

The limiting height of the flare stack depends on the design criteria selected and the facilities near the flare. At grade level, directly under the flame centre, with K up to 9,5 kW/m^2 (3 000 Btu/h·ft^2), the minimum flare stack height, h, is determined as follows:

In SI units:

$$h = D - y_{\text{c}}$$

$$= 126 - 30 = 96 \text{ m}$$

In USC units:

$$h = D - y_{\text{c}}$$

$$= 413 - 100 = 313 \text{ ft}$$

At grade level, at a radius, r, of 45,7 m (150 ft) from the base of the flare stack, with K limited to 6,3 kW/m^2 (2 000 Btu/h·ft^2) and following the general arrangement shown in Figure C.1, h is determined as follows:

In SI units:

$$h' = h + y_{\text{c}}$$

$$r' = r - x_{\text{c}}$$

$$D^2 = r'^2 + h'^2 = 154^2$$

$$D^2 = (r - x_{\text{c}})^2 + (h + y_{\text{c}})^2$$

$$(h + 30)^2 = 154^2 - (45{,}7 - 18)^2 = 22\ 949$$

$$h = 151 - 30 = 121 \text{ m}$$

In USC units:

$$h' = h + y_{\text{c}}$$

$$r' = r - x_{\text{c}}$$

$$D^2 = r'^2 + h'^2 = 507^2$$

$$D^2 = (r - x_{\text{c}})^2 + (h - y_{\text{c}})^2$$

$$(h + 100)^2 = 507^2 - (150 - 58)^2 = 248\ 585$$

$$h = 499 - 100 = 399 \text{ ft}$$

C.3.6 Explanatory notes

C.3.6.1 Lower explosive limits for pure components may be obtained from AGA XK0101 [7] or from NFPA HAZ01 [37]. The lower explosive limits of mixtures may be calculated using Le Chatelier's Rule as follows:

$$C_L = \left[\left(\frac{y_1}{C_{L1}} \right) + \left(\frac{y_2}{C_{L2}} \right) + \ldots + \left(\frac{y_n}{C_{Ln}} \right) \right]^{-1} \tag{C.9}$$

where

C_{L1}, C_{L2},…, C_{Ln} is the lower explosive concentration (i.e. lower flammable limit) of the component 1, 2,..n in air;

$y_1, y_2, \ldots, y_n$ is the mole fraction (or volume fraction) of the component 1, 2,..n in the mixture.

C.3.6.2 The graphs in Figures C.2, C.3, C.4 and C.5 are based on two independent variables, C_L and a modified form of ($d_j \cdot R$). The variable ($d_j \cdot R$) was modified (from that proposed in the Brzustowski and Sommer article [94]) to include gas and air temperatures and relative molecular masses instead of densities. The ideal gas law was assumed. Some adjustments were made in the graph curves over the C_L range from 0,5 to 1,5 to smooth out discontinuities. No significant difference, compared with hand calculated results, is introduced with the data smoothing. See the original article for the details of the hand calculation procedure.

C.3.6.3 Brzustowski and Sommer recommend the use of the fraction of heat intensity transmitted, τ, to correct the radiation impact. The following is quoted from the original article [94]):

In the case of flares, atmospheric absorption attenuates K by about 10 % to 20 % over distances of 150 m (500 ft). The empirical Equations (C.10) and (C.11) are obtained by cross-plotting absorptivities calculated from the Hottel charts. It is strictly applicable only when a luminous, hydrocarbon flame is radiating at 1 227 °C (2 240 °F), the dry bulb ambient temperature is 27 °C (80 °F), the relative humidity is more than 10 %, and the distance from the flame is between 30 m and 150 m (100 ft and 500 ft); however, the equation can be used to estimate the order of magnitude of τ under a wider range of conditions.

In SI units:

$$\tau = 0{,}79 \left(\frac{100}{R_H} \right)^{1/16} \left(\frac{30}{D} \right)^{1/16} \tag{C.10}$$

In USC units:

$$\tau = 0{,}79 \left(\frac{100}{R_H} \right)^{1/16} \left(\frac{100}{D} \right)^{1/16} \tag{C.11}$$

where

τ is the fraction of K transmitted through the atmosphere;

R_H is the relative humidity, expressed as a percentage;

D is the distance from the flame to the illuminated area, expressed in metres (feet).

Equations (C.10) and (C.11) should prove adequate for most flare gases, except H_2 and H_2S which burn with little or no luminous radiation. If the anticipated design conditions are very different from those under which Equations (C.1) and (C.2) were derived, the designer should revert to the Hottel charts.

C.3.6.4 Where steam injection is used at a rate of about 0,3 kg (0,7 lb) of steam per kilogram (pound) of flare gas, then the fraction of heat radiated, F, is decreased by 20 %.

Annex D
(informative)

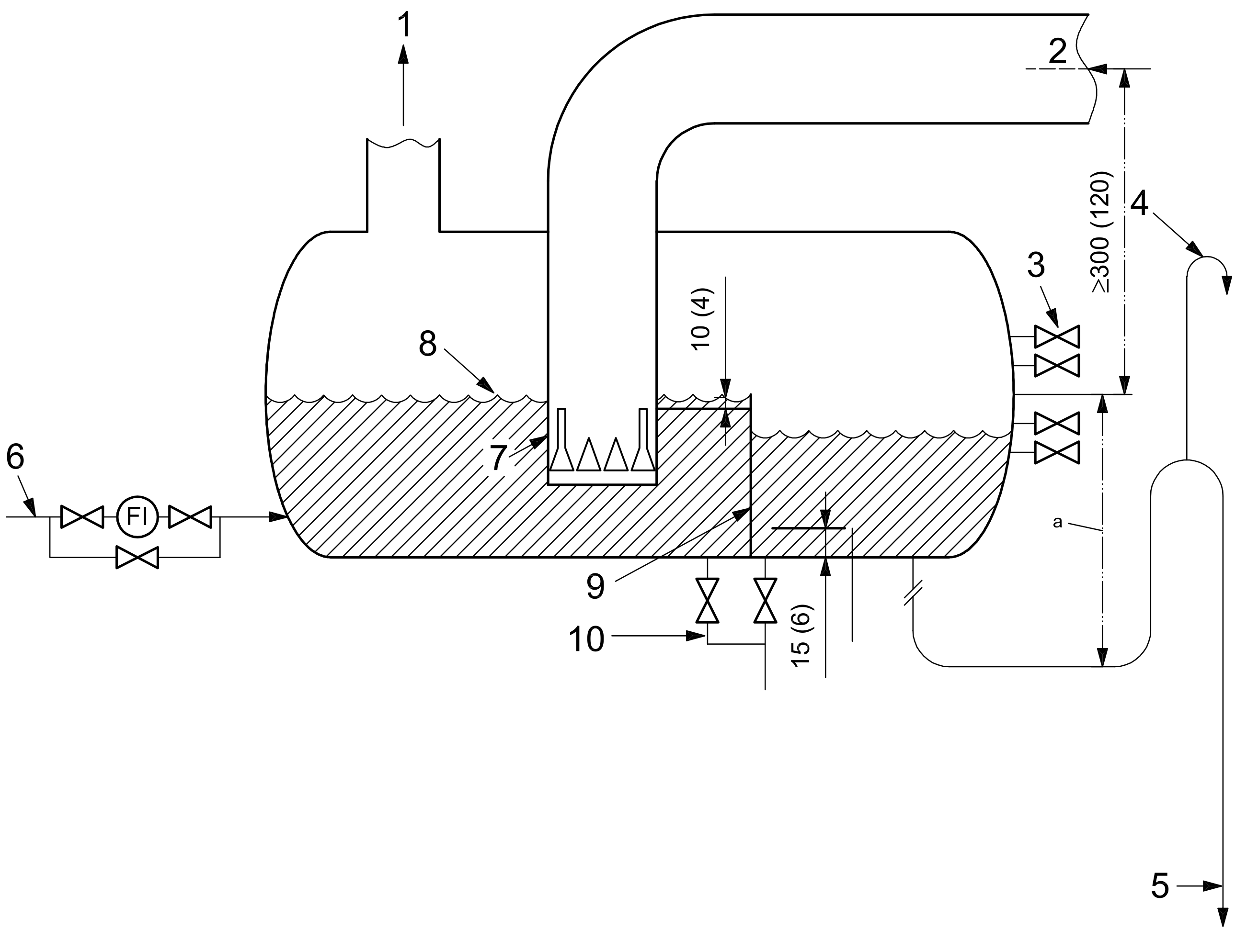
Typical details and sketches

Figure D.1 shows a typical horizontal flare seal drum.

Figure D.2 shows a quench drum.

Figure D.3 shows a typical flare installation.

Dimensions in centimetres (inches)



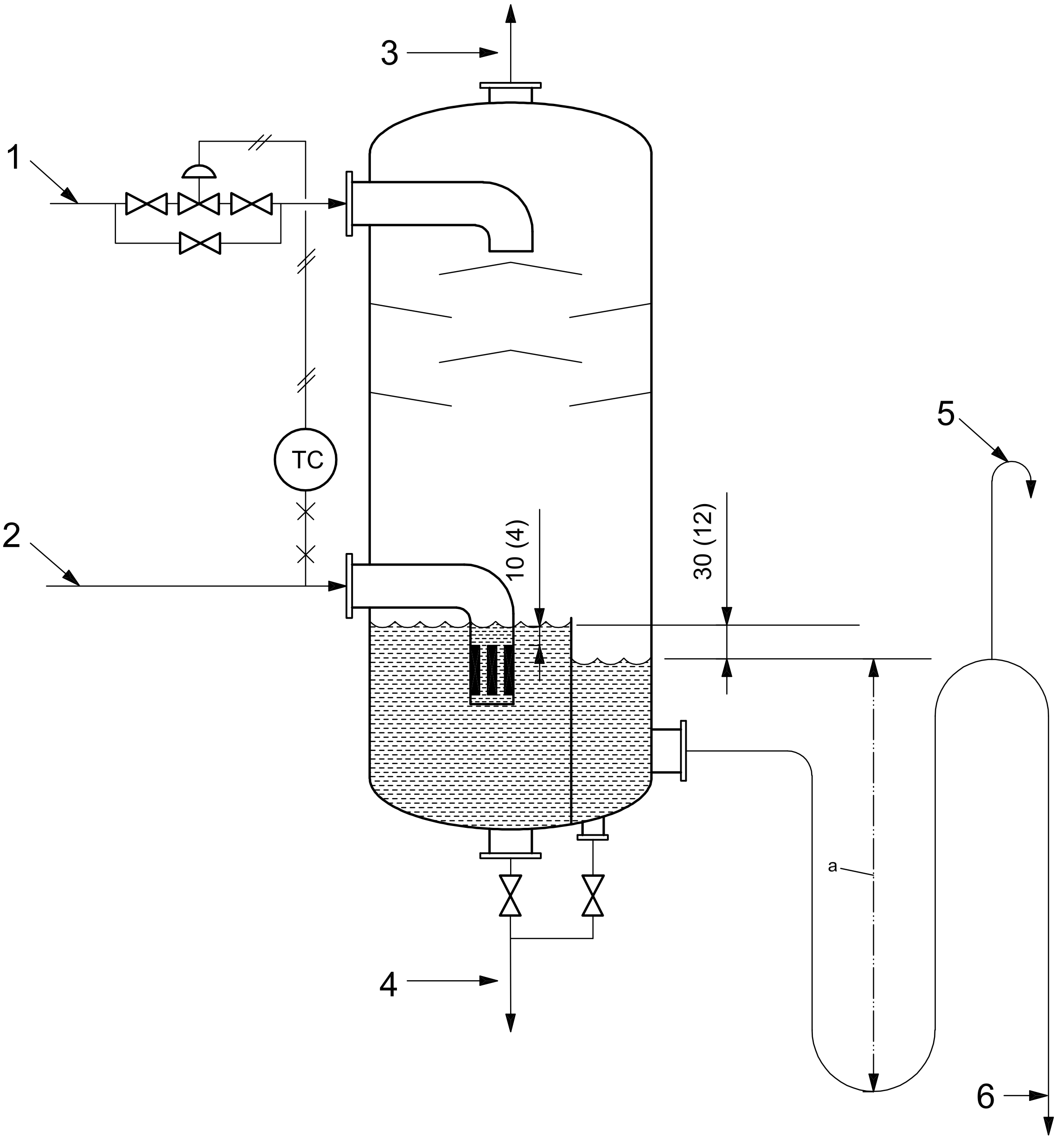
Key

1 to flare
2 flare header
3 Try cocks for checking for hydrocarbons
4 vent
5 to sewer
6 water supply
7 submerged weir welded on end of flare line
8 water level
9 baffle
10 drain

[a] The sewer seal should be designed for a minimum of 175 % of the drum's maximum operating pressure.

Figure D.1 — Typical horizontal flare seal drum

Dimensions in centimetres (inches)

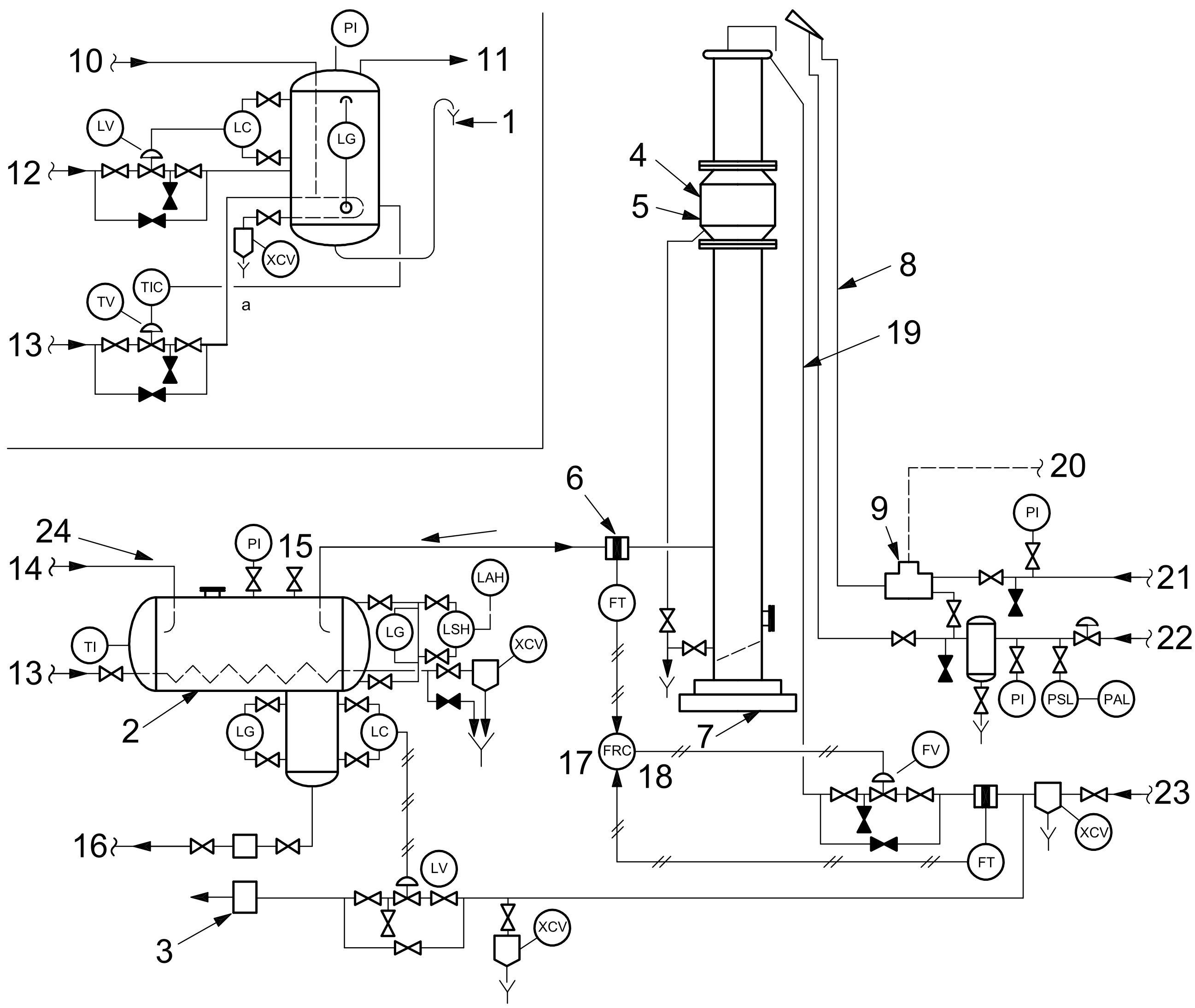


Key

1 cooling water
2 hydrocarbon
3 vent to atmosphere or flare header
4 drain
5 vent
6 water and condensed hydrocarbon out to sewer

[a] The sewer seal should be designed for a minimum of 175 % of the drum's maximum operating pressure; see 7.3.2.2. for details.

Figure D.2 — Quench drum



Key

1 oily water sewer (to sour water system if large quantities of H_2S are flared continuously)
2 knockout drum
3 steam-driven pump and electrically-driven spare
4 molecular seal
5 purge gas
6 flow-measuring element
7 flare stack
8 igniter line
9 flame-front generator
10 from knockout drum
11 to flame stack
12 water
13 steam
14 from relief or vent header system
15 vent
16 to oil recovery facilities or slop
17 panel-mounted
18 ratio
19 steam to nozzle manifold for smokeless burning
20 power supply for spark ignition
21 air supply
22 fuel gas to pilots and ignition
23 steam for smokeless burning
24 slope towards drum

[a] Insert shows alternative sealing method (water seal).

Figure D.3 — Typical flare installation

Figure D.3 represents an operable system arrangement and its components. The arrangement of the system varies with the performance required. Correspondingly, the selection of types and quantities of components, as well as their applications, should match the needs of the particular plant and its specifications.

Annex E
(informative)

High integrity protection systems (HIPS)

E.1 Introduction

Traditional methods of pressure relief employ a mechanical device such as a relief valve for reducing the likelihood of overpressure of vessels and piping systems. A different approach to overpressure protection is the use of an instrumented system. High-integrity protection systems (HIPS) typically involve an arrangement of instruments, final control elements (e.g. valves, switches, etc.), and logic solvers configured in a manner designed to avoid overpressure incidents by removing the source of overpressure or by reducing the probability of an overpressure contingency to such a low level that it is no longer considered to be a credible case.

With appropriate levels of redundancy, a HIPS can be designed to achieve a level of availability equal to or greater than a mechanical relief device. However, the application of HIPS requires a number of special procedures within the design process to ensure an adequately safe HIPS design and it requires particular attention during its operational life such as maintenance, testing and inspection. For these reasons, the decision to implement a HIPS on a given project should be made with a great deal of caution and careful consideration. Note that it can be necessary for the required overpressure protection availability to be higher than that provided by a single mechanical relief device.

This annex provides a discussion of the elements of a HIPS, the applicable codes and standards associated with HIPS and the procedures which should be followed when implementing HIPS.

E.2 Background

E.2.1 Elements of a HIPS

A HIPS includes field instruments (e.g. sensors), logic solving devices (e.g. Safety System Logic Solver, relays, etc.), final control elements, power supply and inspection, testing, and maintenance procedures. The boundaries of a HIPS incorporate all aspects from the sensor to the final element.

E.2.2 Application of HIPS

There are five principal uses of a HIPS:

a) to eliminate a particular overpressure scenario from the design basis;

b) to eliminate the need for a particular relief device;

c) to provide system overpressure protection where a relief device is ineffective;

d) to reduce the probability that several relief devices will have to operate simultaneously, thereby allowing for a reduction in the size of the disposal system;

e) to reduce the demand rate on a relief device consequently reducing the risk.

There is a large amount of overlap between these categories of HIPS applications; a particular application of HIPS can pertain to more than one of the above categories.

One of the chief benefits of HIPS is cost-effectiveness. By eliminating the need for costly upgrades to an existing relief/flare system or by reducing the size of a new relief/flare system, a large amount of capital savings can be realized. In other cases, by demonstrating the need for smaller or fewer pressure-relief devices, less dramatic but still potentially significant cost savings can be realized. Moreover, HIPS can be designed to achieve a higher level of availability/reliability than a mechanical relief device, by using components designed to have very low failure-to-danger rates and which are designed to primarily fail safe by incorporating appropriate levels of redundant instrumentation and by ensuring that the HIPS is inspected and tested on a regular basis. Thus, a HIPS can be used as a risk-reduction measure for particularly high-risk process units (e.g. those which involve acutely toxic materials). In some cases, a HIPS can be used in concert with a relief device (where the relief device is generally a “backup” to the HIPS) to achieve especially high levels of protection. Note, however, that the ongoing cost of ownership of a HIPS should be taken into account. This includes costs of routine testing of the HIPS versus routine pressure-relief device maintenance. This ongoing cost rises disproportionately with the SIL level (see 3.66, 3.67 and E.3.3.2).

HIPS availability addresses only the operation of the HIPS itself. Careful analysis should also be made of the response of the process to the operation of the HIPS. For example, successful operation of a HIPS system to shut off fuel to a fired heater does not eliminate all heat flux to the heater tubes, since there is residual heat contained in the furnace-wall refractory. Further, when a pressure-relief device fails such that it opens during normal operating conditions, the effects on the process can vary from an operational nuisance to a major event. When a HIPS operates inadvertently, it can result in a major shutdown and thus incur the hazards associated with the shutdown and subsequent restart.

E.3 Relevant regulations and industry standards

E.3.1 General

The user should obtain the latest edition of the documents referenced in this annex and review local jurisdictional applicability.

E.3.2 ASME Section VIII, Code Case 2211-1

In 1995/1996, Code Case 2211 was initially approved, with a modified version (2211-1) approved in 1999. Within this Code Case, ASME adopted the opinion that, “a pressure vessel may be provided with overpressure protection by system design in lieu of a mechanical relief device,” under the following conditions, paraphrased from Reference [129].

a) The vessel is not exclusively in air, water or steam service unless these services are critical to preventing the release of fluids that can result in safety or environmental hazards.

b) The decision to provide a vessel with overpressure protection by system design is the responsibility of the user. The manufacturer is responsible only for verifying that the user has specified overpressure protection by system design and for listing this code case on the data report.

c) The user shall ensure that the MAWP of the vessel is greater than the highest pressure that can reasonably be expected to be achieved by the system. The user shall conduct a detailed analysis of all credible overpressure scenarios. This analysis shall utilize an organized, systematic process safety analysis approach such as a hazards and operability (HAZOP) review, a failure modes, effects and criticality analysis (FMECA), fault tree analysis, event tree analysis, what-if analysis or other similar methodology.

d) The analysis described in (c) shall be conducted by an engineer(s) experienced in the applicable analysis methodology. The results of the analysis (including a qualitative or quantitative evaluation of reliability) shall be documented and signed by the individual in charge of the operation of the vessel. The documentation shall be made available to all authorities having jurisdiction at the site where the vessel will be installed. The user is cautioned that prior jurisdictional acceptance can be required.

e) The code case number shall be shown on the manufacturer's data report and it shall be noted that prior jurisdictional acceptance may be required.

E.3.3 ISA S84.01

E.3.3.1 General discussion

ISA S84.01 [31] is intended for those who are involved with safety instrumented systems (SIS) in the areas of design, manufacturing, selection, application, installation, commissioning, testing, operation, maintenance and documentation. The main body of the standard presents normative specific requirements.

NOTE ISA S84.01, while recognized by OSHA as good practice, is not in itself mandatory in the U.S. The annexes present non-mandatory, but informative, technical information, guidance and examples that are useful in such applications.

ISA S84.01 defines a safety instrumented system as a "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. Other terms commonly used include emergency shutdown system (ESD, ESS), safety shutdown system (SSD), and safety interlock system."

HIPS also fit the definition of a safety instrumented system. Accordingly, the philosophy and procedures set forth in ISA S84.01 or other equivalent "good engineering practices" are appropriate for use in the application of HIPS systems for the process industries. ISA S84.01 was released in final form in February 1996. It was adopted as an ANSI standard in March 1997. ISA TR84.02 [32] was issued in 2002 to supplement ISA S84.01; ISA TR84.02 discusses a number of methods for quantifying SIS availability and validating the SIL of proposed designs.

E.3.3.2 Requirements

ISA S84.01 sets forth is a performance standard for the design and life-cycle ownership of a SIS, beginning in the research and development stage and continuing all the way through decommissioning. The discussion in E.3.3.2 focuses on those aspects that should be accounted for during the process design stage of an engineering project where the standard is recognized.

The general steps to be followed in using a SIS to protect against a process hazard can be broken down as follows.

a) Perform a process-hazards analysis (PHA) to identify the hazards to be addressed.

b) Apply non-SIS protection layers first to eliminate identified hazards or reduce the associated risk.

c) Determine if a SIS is needed.

d) Define a target safety-integrity level (SIL) or an actual target-availability value for the SIS based on the perceived risk associated with the identified hazards (taking into account non-SIS protection layers).

ISA S84.01 does not specify the method to be used for the PHA, nor does its scope include list items a) through c) above. While a user is free to select from a number of recognized methods, layer-of-protection analysis (LOPA) is effective in assessing the SIL required of a HIPS versus the availability of pressure-relief devices. CCPS publishes typical risk-reduction factors for various layers of protection, including pressure-relief devices and SISs [132].

The safety-integrity level (SIL-1, SIL-2 or SIL-3) defines the level of performance (availability required of a SIS expressed in the probability of failure upon demand). The intended level of risk reduction is used to determine the SIL actually required. A higher SIL indicates a more robust system and hence a higher level of availability. The performance requirements for each SIL are given in Table E.1.

Verification that the designed SIS actually meets this performance requirement can be performed quantitatively using one of the methods described in ISA TR84.02 or other recognized technique. This analysis is performed in the design stage by a safety/reliability specialist.

E.3.3.3 Application

The process for applying the safety-integrity level requirements of ISA S84.01 to a HIPS application consists of the following general steps.

a) Select appropriate SIL or actual availability value.

b) Design SIS to meet target SIL or availability value.

c) Perform availability calculations to verify system integrity.

d) Specify testing intervals and procedures required to maintain the SIL of the HIPS.

e) Apply management of change processes to any modifications.

There is no requirement in ISA S84.01 to perform reliability calculations where spurious trips can be hazardous. It is suggested to consider the probability of a spurious trip and its effect on the system. The operation of a HIPS typically results in a major shutdown of a unit, obviously requiring a subsequent start-up. It is recognized within Industry that the start-up phase of a unit is the one during which there is a higher possibility of an incident involving loss of containment and or personal injury. In addition, unnecessary shutdowns adversely affects productivity. Thus, it is necessary to ensure that the system configuration reflects a low potential for such spurious trips. Neither the SIS nor the availability values necessarily reflect a high reliability (low frequency of spurious trips). As noted in ISA S84.01, the value and purpose of redundant features, such as 2-out-of-3 voting architecture, in system designs can be more to increase the reliability of the SIS than to increase the availability.

E.3.4 IEC 61508 and IEC 61511

IEC 61508 [3] is the International Standard that addresses the general requirements for identification and implementation of SISs. IEC 61508 is a broad-scope document intended to cover a wide range of industries. IEC 61511 (all parts) [4] was issued in 2003. These documents address IEC 61508 requirements as applicable to the process industries and, with some differences, are equivalent to ISA S84.01. It is likely that once IEC 61511 is adopted, ISA will adopt this standard as a replacement to ISA S84.01. The user is cautioned to verify which standard is effect at the time of the design of HIPS. The chief differences between IEC and ISA standards are summarized as follows.

a) IEC 61508 includes a provision for a fourth safety integrity level, SIL-4, which requires a system’s minimum availability of 99,99 %.

b) IEC 61508 discusses and specifies requirements related to external (non-instrumented) risk-reduction facilities.

c) IEC 61508 requires the use of the ISO 9000 series of quality systems or equivalent, whereas ISA S84.01 does not.

d) IEC 61508 requires the use of a “safety plan,” whereas ISA S84.01 requires documentation consistent with the US regulation 29CFR1910.119 [139].

e) IEC 61508 addresses a number of management-system issues that are outside the scope of ISA S84.01. For US applications, these management-system issues are covered extensively by 29CFR1910.119.

f) IEC 61508 uses a number of terms and abbreviations that are slightly different from those found in ISA S84.01.

g) IEC 61508 has additional hardware and software requirements that becomes more demanding with increasing SIL requirement.

The user should ensure that component-failure rates applied in availability calculations accurately reflect the type and model of the component installed in the HIPS and the intended service. The component failure rates are key to calculating HIPS availability and setting the required testing intervals. While using one's own company's and service-specific data is preferable, there are a number of publicly available sources listing component-failure rates. The selected data should best approximate the industry, equipment and service planned for or installed in the HIPS and should be based on sufficient populations of unrevealed failure events to be statistically significant.

E.3.5 Comparison of various SIL standards

Table E.1 compares the various SIL standards.

Table E.1 — SIL versus availability

Safety integrity level (SIL)			SIS performance requirements	
ISA S84.01	**IEC 61508**	**DIN V 19250** [140] **(TUV class)**	**Safety availability required** %	**Average probability of failure on demand** PFD_{avg}
1	1	1 and 2	90,00 to 99,00	10^{-1} to 10^{-2}
2	2	3 and 4	99,00 to 99,90	10^{-2} to 10^{-3}
3	3	5 and 6	99,90 to 99,99	10^{-3} to 10^{-4}
	4	7	$>$ 99,99	$< 10^{-4}$
—	—	8	$>$ 99,999	$< 10^{-5}$

E.4 Procedures for applying HIPS

E.4.1 General

The use of HIPS for any particular application has both advantages and disadvantages. Thus, for a given case, it is necessary to weigh the risk versus the benefit, and make a well-considered, informed decision as to whether HIPS is the best option.

E.4.2 Safety integrity level assignment or availability value

In accordance with ISA S84.01, a necessary step in safety-instrumented system design is to set a safety integrity level (SIL) or availability value target for system design. The system is assigned as a SIL-1, SIL-2 or SIL-3 system, with SIL-3 being the most robust and most reliable and SIL-1 being the least. Associated with each SIL is a minimum performance requirement, i.e., a minimum of 90 % availability for SIL-1, a minimum of 99 % availability for SIL-2, and a minimum of 99,9 % availability for a SIL-3 system. The determination of target SIL for a given system is dependent upon the risk associated with the hazard that the system is protecting against, i.e. the likelihood of the initiating and contributing events, the magnitude of the consequences and the credit that can be taken for other safeguards. The SIL assignment should be performed by a multi-disciplinary team.

The acceptability criterion for HIPS performance is expressed in terms of the SIL level, which corresponds to a level of system availability (i.e., the probability that the system will work properly when needed). Each case should be examined individually to determine the appropriate response. The selected SIL for a given system is dependent upon a number of factors, including the following:

a) likelihood of placing a demand on the HIPS in the first place (i.e., the likelihood of getting a high-high pressure situation that requires proper action from the HIPS in order to prevent a negative consequence);

b) consequences of a failure of the HIPS, given that a demand has been placed on it;

c) risk tolerance of the user;

d) requirements from local jurisdictional authorities.

In the large majority of cases for HIPS, the result of the hazard analysis is either a SIL-2 system (requiring a minimum of 99 % availability) or a SIL-3 system (requiring a minimum of 99,9 % availability).

E.4.3 Conceptual proposal of HIPS configuration

After a target SIL or availability value has been assigned, a base-case HIPS configuration should be devised, with the intention of arriving at a system configuration that meets the availability requirement associated with the assigned SIL. At this point, a base-case maintenance/testing interval for the individual components of the HIPS should be decided as well. The base-case configuration and test data then serve as the basis for the next step in the work process, the reliability analysis.

E.4.4 HIPS availability analysis

The purpose of the HIPS availability analysis is to evaluate the system performance of the proposed configuration. The availability analysis should utilize standard techniques, such as fault-tree analysis (see ISA TR-84.02). In addition to considering the integrity of the hardware, which is addressed in part by suitable scope and frequency of testing, the potential for human error and other sources of systematic failures throughout the system lifecycle should also be considered. The result produced is compared against the performance requirement associated with the assigned safety integrity level to determine if the proposed system is acceptable. If the proposed system does not meet the performance requirement, then it is necessary to modify the system configuration by the following:

a) using better quality sensors, logic solvers and final control elements with lower unrevealed failure rates;

b) increasing the level of diagnostic coverage, so that fewer of the failure modes are unrevealed and become revealed through the action of the diagnostics facility;

c) using redundant components (i.e., duplicated or even triplicated elements);

d) using diverse components;

e) increasing the planned testing frequency.

Careful consideration should also be given in the system design to the calculated rate of nuisance failures. Nuisance trips can be costly and also increase the risk that operators might circumvent the shutdown systems. The potential for nuisance failures can be reduced by including functions such as voting logic and similar techniques to make the application more robust to spurious failures. These provisions can decrease the required testing interval.

E.5 Test intervals for HIPS

For a typical safety instrumented system, the large majority of instrument failures that it is necessary to consider are the type of failure that is referred to as a dormant fault (or covert or unrevealed failure). A dormant fault occurs when an equipment item that is not under constant demand fails, such that the failure is not immediately detected. Such a failure is detected only when either the item fails on demand during a test or the item fails on an actual demand. Accordingly, the frequency of testing plays an important role in determining the availability of an item to perform its intended function on demand.

Take, for example, the case of a high-pressure switch. The switch can be expected to enter a failed state at a constant rate in units of 1/time. However, since the switch is only required to perform its function in the event of high pressure, this failure is not revealed until there is a demand placed on it, either through a test or a true process demand. It is assumed that if the switch fails on demand, it will be immediately repaired or replaced. Thus, the unavailability of the pressure switch, or the probability that it will fail on demand, is a direct function of the test interval.

NOTE This example is illustrative. In a SIL-2 or SIL-3 application, process-connected switches are seldom sufficiently available to support the required integrity. Process transmitters are generally used because of their lower unrevealed failure rates.

Because of this, it is critical that the instrumentation associated with a HIPS be tested at regular intervals. The availability analysis assumes a test interval for each piece of equipment. In order for the actual HIPS reliability to align with that predicted by the availability calculations, it is necessary that the actual testing frequency (during operation) corresponds with that assumed in the availability calculations performed during the design.

There are two other important aspects of testing that should be considered in setting testing intervals. The first is the capability of the site at which the HIPS is to be installed to carry out such tests. There is little value in specifying a system that requires testing a HIPS every three months where the site only has the resources for annual testing. Secondly, testing as a process contains the potential for introducing faults and spurious shutdowns due to human error. The consequences of such events can be hazardous and every effort should be made to minimize their occurrence. Thus, where possible, the aim should be to design a system that can achieve the desired availability with the minimum of off-line testing. The advent of supervised circuits, built-in diagnostics and the use of built-in redundancy (that facilitates on-line testing of components and circuits) all contribute to minimizing the frequency of off-line testing.

E.6 Documentation

As indicated in ISA S84.01, documentation of HIPS should be consistent with the US regulation 29CFR1910.119. All of the design criteria, calculations, installation details and follow-on maintenance and testing should be properly documented.

E.7 Training

Adequate training should be provided to operating and maintenance personnel to ensure that the integrity of the HIPS system is maintained as designed.

E.8 Additional source material

Additional source material can be found in References [130], [131], [132], [133], [134], [135], [136], [137].

Bibliography

[1] ISO 13707, *Petroleum and natural gas industries — Reciprocating compressors*

[2] ISO 15649, *Petroleum and natural gas industries — Piping*

[3] IEC 61508 (all parts), *Functional safety of electrical/electronic/programmable electronic safety-related systems*

[4] IEC 61511 (all parts), *Functional safety — Safety instrumented systems for the process industry sector*

[5] EN 764-7, *Pressure Equipment — Part 7: Safety systems for unfired pressure vessels* [4)]

[6] prEN 14015-1, *Specification for the design and manufacture of site built, vertical, cylindrical, flat-bottomed, above ground, welded, metallic tanks for the storage of liquids at ambient temperature and above — Part 1: Steel tanks*

[7] AGA XK0101, *Purging Principles and Practice* [5)]

[8] API RP 520-II:2003, *Sizing, Selection, and Installation of Pressure-Relieving Devices in Refineries — Part II: Installation*

[9] API Std 526, *Flanged Steel Pressure Relief Valves*

[10] API Std 537, *Flare Details for General Refinery and Petrochemical Service*

[11] API Std 618, *Reciprocating Compressors for Petroleum, Chemical, and Gas Industry Services*

[12] API Std 620, *Design and Construction of Large, Welded Low-Pressure Storage Tanks*

[13] API Std 650, *Welded Steel Tanks for Oil Storage*

[14] API Std 2000, *Venting Atmospheric and Low-Pressure Storage Tanks: Nonrefrigerated and Refrigerated*

[15] API RP 2003, *Protection Against Ignitions Arising Out of Static, Lightning, and Stray Currents*

[16] API Publ 2030, *Guidelines for Application of Water Spray Systems for Fire Protection in the Petroleum Industry*

[17] API Publ 2216, *Ignition Risk of Hydrocarbon Vapours by Hot Surfaces in Open Air*

[18] API Publ 2218, *Fireproofing Practices in Petroleum and Petrochemical Processing Plants*

[19] API Std 2510, *Design and Construction of Liquified Petroleum Gas Installations (LPG)*

[20] ASME, *Boiler and Pressure Vessel Code, Section VIII, Rules for Construction of Pressure Vessels* [6)]

[21] ASME B31.3, *Process Piping*

4) Comité Européen de Normalisation, 36, rue de Stassart, B-1050 Brussels, Belgium.

5) American Gas Association, 400 N. Capitol St., NW, Suite 450, Washington DC 20001, USA.

6) American Society of Mechanical Engineers, Three Park Avenue, New York, NY 10016-5990, USA.

[22] ASTM A515, *Standard Specification for Pressure Vessel Plates, Carbon Steel, for Intermediate- and Higher-Temperature Service* [7)]

[23] ASTM C533, *Standard Specification for Calcium Silicate Block and Pipe Thermal Insulation*

[24] ASTM C552, *Standard Specification for Cellular Glass Thermal Insulation*

[25] ASTM C553, *Specification for Mineral Fiber Blanket Thermal Insulation for Commercial and Industrial Applications*

[26] ASTM C592, *Standard Specification for Mineral Fiber Blanket Insulation and Blanket-Type Pipe Insulation (Metal-Mesh Covered) (Industrial Type)*

[27] ASTM C610, *Standard Specification for Molded Expanded Perlite Block and Pipe Thermal Insulation*

[28] ASTM C612, *Standard Specification for Mineral Fiber Block and Board Thermal Insulation*

[29] ASTM Data Series DS 11S1, *An evaluation of the elevated temperature, tensile, and creep-rupture properties of wrought carbon steel*, prepared for the Metals Properties Council by G. V. Smith

[30] Scandpower Report 27.207.291/R1, Version 2, *Guidelines for the protection of pressurised systems exposed to fire*, March 31, 2004

[31] ISA S84.01, *Application of Safety Instrumented Systems for the Process Industries* [8)]

[32] ISA TR84.02, *Functional Safety: Safety Instrumented Systems for the Process Industry Sector — Part 2: Guidelines for the Application of ANSI/ISA-84.00.01-2004*, IEC 61511-2 Mod

[33] NFPA 15, *Standard for Water Spray Fixed Systems for Fire Protection* [9)]

[34] NFPA 68, *Guide for Venting of Deflagration*

[35] NFPA 69, *Explosion Prevention Systems*

[36] NFPA 780, *Standard for the Installation of Lightning Protection Systems*

[37] NFPA HAZ01, *Fire-protection guide to hazardous materials*

[38] M. A. GROLMES, J. C. LEUNG, and H. K. FAUSKE, *Reactive Systems Vent Sizing Evaluations* International Symposium on Runaway Reactions, ISBN 0-8169-0460-X, March 7–9, 1989, Cambridge, Massachusetts, pp. 451–476

[39] H. G. FISHER, et al, *Emergency Relief System Design Using DIERS Technology*, ISBN 0-8169-0568-1, American Institute of Chemical Engineers, New York, 1992

[40] H. G. FISHER, *An Overview of Emergency Relief System Design Practice*, Plant Operations Progress, January 1991

[41] J. C. LEUNG, *Overpressure During Emergency Relief Venting in Bubbly and Churn Turbulent Flow*, AIChE Journal, Volume 33, Number 6, June 1987

7) ASTM International, 100 Barr Harbor Drive, PO Box C700, West Conshohocken, PA, 19428-2959 USA.

8) Instrumentation, Systems, and Automation Society, 67 Alexander Drive, Research Triangle Park, NC 27709 USA.

9) National Fire Protection Association, 1 Batterymarch Park, Quincy, MA 02269-9101, USA.

[42] J. C. LEUNG, *Simplified Vent Sizing Equations for Emergency Relief Requirements in Reactors and Storage Vessels*, AIChE Journal, Volume 32, Number 10, October 1986, pp. 1622-1634

[43] Cyril F. PARRY, *Relief Systems Handbook*, ISBN 0-85295-267-8, Institution of Chemical Engineers, Rugby, Warwickshire, U.K., 1992

[44] G. C. KARCHER, *Pressure Changes in Liquid Filled Vessels or Piping Due to Temperature Changes*, ExxonMobil Research and Engineering Mechaniacl Newsletter, EE.84E.76, August 1976

[45] Center for Chemical Process Safety, *Guidelines for Pressure Relief and Effluent Handling Systems*, Section 3A.5, Published by the American Institute of Chemical Engineers, 1998, ISBN 0-8169-0476-6

[46] R. H. PERRY, D. W. GREEN, J. O. Maloney Editors, *Perry's Chemical Engineers' Handbook*, Sixth Edition (ISBN 0-07-049479-7, 1984) and Fifth Edition (ISBN 0-07-049478-9, 1969)

[47] J. A. DEAN (Editor), *Lange's Handbook of Chemistry*, Twelfth Edition, ISBN 0-07-016191-7, 1979

[48] H. R. WHARTON, *Digest of Steels for High-Temperature Service*, Timken Steel, 1946

[49] F. J. HELLER, *Safety Relief Valve Sizing: API Versus CGA Requirements Plus a New Concept for Tank Cars*, 1983 Proceedings—Refining Department, Volume 62, American Petroleum Institute, Washington, D.C., pp. 123–135

[50] J. J. DUGGAN, C. H. GILMOUR, and P. F. FISHER, *Requirements for Relief of Overpressure in Vessels Exposed to Fire*, Transactions of the ASME, 1944, Volume 66, pp. 1–53

[51] I. HEITNER, T. TRAUTMAUIS, and M. MORRISSEY, *Relieving Requirements for Gas Filled Vessels Exposed to Fire*, 1983 Proceedings—Refining Department, Volume 62, American Petroleum Institute, Washington, D.C., pp. 112–122, and Hydrocarbon Processing, November 1983, pp. 263-268

[52] J. O. FRANCIS and W. E. SHACKELTON, *A Calculation of Relieving Requirements in the Critical Region*, 1985 Proceedings — Refining Department, Volume 64, American Petroleum Institute, pp. 179–182

[53] H. G. FISHER and H. S. FORREST, *Protection of Storage Tanks from Two-Phase Flow Due to Fire Exposure*, Process Safety Progress, Volume 14, No. 3, July 1995, pp. 183-199

[54] H. S. FORREST, *Emergency Relief System Design for Fire Exposure with Consideration of Multiphase Flow*, 1995 International Symposium on Runaway Reactions and Pressure Relief Design, ISBN 0-8169-0676-9, Published by American Institute of Chemical Engineers, pp. 604-630

[55] H. MOTT and B. SPARROW, *A Simple Relief Valve Rating/Sizing Method for Fire Relief from Liquid-filled Vessels*, presentation to API RP-521 Subcommittee, September 30, 2002

[56] L. L. SIMPSON, *Fire Exposure of Liquid-Filled Vessels*, paper D29-190-1 presented at the 29th DIERS Users Group Meeting, Las Vegas, NV, April 29-May 1, 2002

[57] H. K. FAUSKE, et al, *Emergency Relief Vent Sizing for Fire Emergencies Involving Liquid Filled Atmospheric Storage Vessels*, Plant Operations Progress, Volume 5, Number 4, October 1986, pp. 205-208

[58] T. Z. HARMATHY and L. W. ALLEN, *Thermal Properties of Selected Masonry Unit Concretes*, Title Number 70-15, ACI Journal, February 1973

[59] Crane Engineering Division Technical Paper 410, *Flow of Fluids Through Valves, Fittings, and Pipe*

[60] J. C. LEUNG, *Size Safety Relief Valves for Flashing Liquids*, Chemical Engineering Progress, February 1992, pp. 70-75

[61] J. C. LEUNG, and F. N. NAZARIO, *Two-Phase Flashing Flow Methods and Comparison*, Journal of Loss Prevention Process Industries, Volume 3, Number 2, 1990, pp. 253-260

[62] H. K. FAUSKE, *Scale-up for Safety Relief of Runaway Reactions* published in Plant/Operations Progress, Volume 3, No. 1, 1984, pp. 7-11

[63] R. E. HENRY and H. K. FAUSKE, *The Two Phase Critical Flow of One Component Mixtures in Nozzles*, Orifices, and Short Tubes published in Journal of Heat Transfer, May 1971, pp. 179-187

[64] D. W. FOWLER, T. R. HERNDON and R. C. WAHRMUND, *An Analysis of Potential Overpressure of Heat Exchanger Shell Due to a Ruptured Tube*, Paper presented at the American Society of Mechanical Engineers Petroleum Division Conference, September 22–25, 1968

[65] V. H. SUMARIA, J. A. ROVNAK, I. HEITNER and R. J. HERBERT, *Model to Predict Transient Consequences of a Heat Exchanger Tube Rupture*, Proceedings—Refining Department, 1976, Volume 55, American Petroleum Institute, Washington, D.C., pp. 631–654

[66] Institute of Petroleum *Guidelines for the Design and Safe Operation of Shell and Tube Heat Exchangers to Withstand the Impact of Tube Failure*, August 2000, ISBN 0-85293-286-3

[67] Institute of Petroleum, *Guidelines for the design and protection of pressure systems to withstand severe fires*, 2003, ISBN 0-85293-279-0

[68] L. L. SIMPSON, *Estimate Two-Phase Flow in Safety Devices*, Chemical Engineering, August 1991, pp. 98-102

[69] M. G. ZABETAKIS, *Flammability Characteristics of Combustible Gases and Vapours*, U.S. Bureau of Mines Bulletin 627, U.S. Department of the Interior, Washington, D.C., 1965

[70] V. O. HOEHNE, R. G. LUCE and L. W. MIGA, *The Effect of Velocity, Temperature, and Gas Molecular Weight on Flammability Limits in Wind-Blown Jets of Hydrocarbon Gases*, Report to the American Petroleum Institute, Battelle Memorial Institute, Columbus Ohio, 1970 and Proc. API Division of Refining, Volume 50, 1970, pages 1057-1081

[71] J. F. TAYLOR, H. L. GRIMMETT and E. C. COMINGS, *Isothermal Free Jets of Air Mixing with Air*, Chemical Engineering Progress, 1951, Volume 47, pp. 175–180

[72] *International Safety Guide for Oil Tankers and Terminals*, International Chamber of Shipping/Oil Companies International Marine Forum, Witherby and Co., London, 4th edition, 1996

[73] D. E. LOUDON, *Requirements for Safe Discharge of Hydrocarbons to Atmosphere* API Proceedings, 1963, Volume 43, American Petroleum Institute, Washington, D.C., pp. 418–433

[74] J. H. BURGOYNE, *Mist and Spray Explosions*, Chemical Engineering Progress, 1957, Volume 53, Number 3, pp. 121–124

[75] D. I. SALETAN, *The Theory Behind Static Electricity Hazards in Process Plants*, Chemical Engineering, 1959, Volume 66, Number 1, pp. 99–102

[76] F. A. GIFFORD, Jr., *Atmospheric Dispersion Calculations Using the Generalized Garrison Plume Model*, Nuclear Safety, December 1960, pp. 56–59

[77] F. G. EICHEL, *Electrostatics* Chemical Engineering, March 13, 1967, pp. 153–167

[78] J. W. YOUNG, et al., *Electrostatic and Gas Blowout*, Canadian Mining Meeting Bulletin Number 63, 1960, p. 456 and Metallurgical Bulletin, Volume 53, No. 581, pages 682-686, September 1960

[79] *Toroidal Ring Prevents Gas Ignition at Vent Stack Outlet*, National Aeronautics and Space Administration Technical Brief 67-10098, National Technical Information Service, Springfield, Virginia, 1967

[80] *Guidelines for Evaluating the Characteristics of Vapour Cloud Explosions, Flash Fires, and BLEVES*, Center for Chemical Process Safety, American Institute of Chemical Engineers, New York, 1994, ISBN 0-8169-0474-x

[81] *American Conference of Governmental Industrial Hygienists' Proposed Threshold Limit Value for Noise*, American Industrial Hygiene Association Journal, November-December 1968, Volume 26, Number 26, pp. 537–540

[82] H. W. HUSA, *How to Compute Safe Purge Rates*, Hydrocarbon Processing and Petroleum Refiner, 1964, Volume 43, Number 5, pp. 179–182

[83] S. H. TAN, *Flare System Design Simplified*, Hydrocarbon Processing and Petroleum Refiner, 1967, Volume 46, Number 1, pp. 172–176

[84] G. R. KENT, *Practical Design of Flare Stacks*, Hydrocarbon Processing and Petroleum Refiner, 1964, Volume 43, Number 8, pp. 121–125

[85] J. D. HAJEK and E. E. LUDWIG, *How to Design Safe Flare Stacks, Part 1, Petro/Chem Engineer*, 1960, Volume 32, Number 6, pp. C31–C38; Part 2, Petro/Chem Engineer, 1960, Volume 32, Number 7, pp. C44–C51

[86] R. SCHWARTZ and M. KELLER, *Flaring in Hostile Environments*, presented at a Seminar on Flare Systems arranged by the Norwegian Society of Chartered Engineers, 1982

[87] R. E. SCHWARTZ and J. W. WHITE, *Predict Radiation From Flares*, Chemical Engineering Progress, July 1997, pp. 42-49

[88] R. E. SCHWARTZ and S. G. KANG, *Effective Design of Emergency Flaring Systems*, Hydrocarbon Engineering, Feb. 1998, pp. 57-62

[89] M. MCDANIEL, *Flare Efficiency Study*, United States EPA Report No.600/2-83-052, July 1983 (PB83-261644)

[90] J. R. ARTHUR, *Some Reactions of Atomic Hydrogen in Flames*, Nature, 1950, Volume 165, pp. 557-558

[91] W. R. SMITH, *Eliminate Air-Polluting Smoke*, Petroleum Processing, June 1954, pp. 887–881

[92] A. M. STOLL and L. C. GREEN, *The Production of Burns by Thermal Radiation of Medium Intensity*, Paper Number 58-A-219, American Society of Mechanical Engineers, New York, 1958

[93] J. K. J. BUETTNER, *Heat Transfer and Safe Exposure Time for Man in Extreme Thermal Environment*, Paper Number 57-SA-20, American Society of Mechanical Engineers, New York, 1957

[94] T. A. BRZUSTOWSKI and E. C. SOMMER, Jr., *Predicting Radiant Heating from Flares*, Proceedings–Division of Refining, 1973, Volume 53, American Petroleum Institute, Washington, D.C., pp. 865–893

[95] M. G. ZABETAKIS and D. S. BURGESS, *Research on the Hazards Associated with the Production and Handling of Liquid Hydrogen*, U.S. Bureau of Mines Rept. Invest. 5707, U.S. Department of the Interior, Washington, D.C., 1961

[96] R. SCHWARTZ and M. KELLER, Environmental Factors Versus Flare Application, Paper Number 13d, presented at the Symposium on Loss Prevention in the Chemical Industry, 83rd National Meeting of the American Institute of Chemical Engineers, Houston, Texas, March 22, 1977

[97] L. D. CLEVELAND, *Oil and Gas Journal*, August 4, 1952

[98] R. D. REED, *Design and Operation of Flare Systems*, Chemical Engineering Progress, June 1968, Volume 64, Number 6

[99] EPA-600/2-86-080, *Evaluation of the Efficiency of Industrial Flares: H_2S Gas Mixtures and Pilot Assisted Flares*, Environmental Protection Agency, Washington, D.C., September 1986

[100] *Prevention of Brittle Fracture of Pressure Vessels*, 1st edition, ANSI/API Publication 920-1992, American Petroleum Institute, Washington, D.C., 1990

[101] C. BUXTON, *Modulating PRV Performance Data*, Presentation to the Pressure Relieving Systems Committee at the Fall 2003 API Refining Meeting, September 17, 2003

[102] O. COX, Jr., and M. L. WEIRICK, *Sizing Safety Valve Inlet Lines*, Chemical Engineering Progress, November, 1980

[103] B. A. VAN BOSKIRK, *Sensitivity of Relief Valves to Inlet and Outlet Line Lengths*, Chemical Engineering, August, 1982

[104] J. P. BRILL and H. D. BEGGS, *Two-Phase Flow in Pipes*, Published by the University of Tulsa, Oklahoma, USA, Sixth Edition April 1994

[105] H. Y. MAK, *New Method Speeds Pressure-Relief Manifold Design*, Oil and Gas Journal, November 20, 1978

[106] P. KANDELL, *Program Sizes Pipe and Flare Manifolds for Compressible Flow*, Chemical Engineering, June 29, 1981

[107] C. E. LAPPLE, *Isothermal and Adiabatic Flow of Compressible Fluids*, Transactions of the American Institute of Chemical Engineers, 1943, Volume 39, pp. 385–432

[108] B. M. COULTER, *Compressible Flow Manual*, Fluid Research Publishing Company, Melbourne, Florida

[109] D. M. KIRKPATRICK, *Simpler Sizing of Gas Piping*, Hydrocarbon Processing, December, 1969

[110] J. E. HUFF and K. R. SHAW, *Measurement of Flow Resistance of Rupture Disc Devices*, Plant/Operations Progress, July, 1992

[111] J. C. LEUNG and M. A. GROLMES, *The Discharge of Two-Phase Flashing Flow in a Horizontal Duct*, AIChE Journal, Volume 33, No. 3, March 1987, pp. 524-527

[112] *Guidelines for Pressure Relief and Effluent Handling Systems*, American Institute of Chemical Engineers, Center for Chemical Process Safety, 1998, ISBN 0-8169-0476-6

[113] R. H. PERRY, C. H. CHILTON, and S. D. KIRKPATRICK, editors, *Chemical Engineers Handbook*, 5th edition, McGraw-Hill, New York, 1972, pp. 5–61

[114] C.E. BAUKAL, Jr., Editor and R. E. SCHWARTZ, Associate Editor, *The John Zink Combustion Handbook*, CRC Press, Boca Raton, FL, 2001, ISBN 0-8439-2337-1

[115] J. G. SEEBOLD, *Pulsating Combustion in Elevated Flares Caused by Seal Drum Sloshing*, NOISE CONTROL ENGINEERING July-August 1974, pp.14-17

[116] A. D. SCHEIMAN, *How to Size Shower Deck Baffled Towers Quicker: Part 1—Tower Diameter*, Petro-Chem Engineer, 1965, Volume 37, Number 3, pp. 28–33

[117] A. D. SCHEIMAN, *How to Size Shower Deck Baffled Towers Quicker: Part 2—Tower Tangent Length*, Petro/Chem Engineer, 1965, Volume 37, Number 4, pp. 75, 78–79

[118] US EPA 40CFR60.18, *General Control Device Requirements*, Code of Federal Regulations, Government Printing Office, Washington, D.C

[119] Health and Safety Executive, Offshore Technology Report, 2002/023, *Testing and analysis of relief device opening times*

[120] J. LENCLUD and J. VENARD, *Single and two-phase discharge from a pressurized vessel*, Rev Gén Therm (1996) 35 503-516

[121] W. B. HOWARD, *Use Precaution in the Selection, Installation, and Operation of Flame Arresters*, Chemical Engineering Progress April 1992, Paper RP92-01

[122] H. W. HUSA, *How to Compute Safe Purge Rates*, Hydrocarbon Processing and Petroleum Refiner, 1964, Volume 43, Number 5, pp. 179–182

[123] H. W. HUSA, *Purging Requirements of Large Diameter Stacks*, paper presented at the Fire/Safety Engineering Subcommittee of the American Petroleum Institute, Fall 1977 Meeting, September 13-15, 1977, San Francisco, CA

[124] G. BUSS and P. TROMANS, *Reduced Purge Rates in Natural Gas Vent Stacks*, paper presented at the Major Hazards On- and Offshore Conference, UMIST, Manchester, October, 24-26, 1995, IChemE Symposium Series No. 139, pp. 595-607, 1995

[125] N. E. SYLVANDER and D. L. KATZ, *Investigation of Pressure Relieving Systems*, Engineering Research Bulletin Number 31, University of Michigan, Ann Arbor, April 1948

[126] American Petroleum Institute, *Recommended Practice for the Design and Construction of Pressure Relieving Systems for Process Equipment and Pressure Storage in Refineries (tentative)*, First Edition, September 1955

[127] F. J. HELLER, *Safety Relief Valve Sizing: API Versus CGA Requirements Plus a New Concept for Tank Cars*, 1983 Proceedings—Refining Department, Volume 62, American Petroleum Institute, Washington, D.C., pp. 123–135

[128] API Publication 999 (English Edition), *Technical Data Book—Petroleum Refining*, American Petroleum Institute, Washington, D.C.

[129] ASME Code Case 2211, *Pressure Vessels with Overpressure Protection by System Design*, Section VIII, Divisions 1 and 2, American Society of Mechanical Engineers, August 1996

[130] Center for Chemical Process Safety (CCPS), *Guidelines for Engineering Design for Process Safety*, ISBN 0-8169-0565-7, 1993

[131] Center for Chemical Process Safety (CCPS), *Guidelines for Safe Automation of Chemical Processes*, American Institute of Chemical Engineers, New York, NY, 1993, ISBN 0-8169-0554-1

[132] Center for Chemical Process Safety (CCPS), *Layer of Protection Analysis: A simplified Process Risk Assessment*, Published by American Institute of Chemical Engineers, New York, New York, 2001, ISBN 0-8169-0811-7

[133] K. A. KIMBERLY A. and A. E. SUMMERS, *Are Your Instrumented Safety Systems Up to Standard?*, Chemical Engineering Progress, November 1998

[134] D. F. MONTAGUE and J. J. ROONEY, *Trade-off Risks: Addressing the Tension between Reliability and Safety*, International Conference and Workshop on Reliability and Risk Management, San Antonio, TX 1998

[135] M. D. MOOSEMILLER and W. H. BROWN, *Finding an Appropriate Level of Safeguards*, International Conference and Workshop on Risk Analysis in Process Safety, Atlanta, GA, 1997

[136] I. D. PIERCE, *Reliability*, 85 5B/R/1, 1985

[137] J. C. A. WINDHORST, *Overpressure Protection by Means of a Designed System Rather Than Pressure Relief Device*, International Conference and Workshop on Risk Analysis in Process Safety, Atlanta, GA, 1997

[138] R.E. SANDERS, *Chemical Process Safety*, Published by Butterworth-Heinemann, pp. 21-34, pp. 54-57

[139] OSHA 29CFR1910.119, *Process safety management of highly hazardous chemicals*, Code of Federal Regulations, Government Printing Office, Washington, D.C.

[140] DIN V 19250, Control Technology; Fundamental Safety Aspects To Be Considered for Measurement and Control Equipment

[141] E. STANGE et al., Determination of temperatures and flare rates during depressurization and fire, paper presented at 72nd annual Gas Processors Association (GPA) Convention, San Antonio, TX, 1993

[142] API RP 521, *Guide for Pressure Relieving and Depressurizing Systems*

[143] Tube-Turn Catalogue and Engineering Data Book No. 211, Chemetron Corp., Louisville, Kentucky

[144] H.C. HOTTEL, *Private communication to API Subcommittee on Pressure-Relieving Systems*, January 1948

[145] H.C. HOTTEL, *Private communication to API Subcommittee on Pressure-Relieving Systems*, December 1950

[146] F.L. MAKER, *Private communication to API Subcommittee on Pressure-Relieving Systems*, regarding 1925 tests, December 22, 1950

[147] *Opacity of Water to Radiant Heat Energy*, Research Bulletin 3, Underwriters Laboratory, Inc., Northbrook, Illinois, 1938

[148] Safety Memorandum 89, Rubber Reserve Corporation, Washington, DC, May 1944

[149] University of Michigan, Unpublished tests made for API Subcommittee on Pressure-Relieving Systems, June 1947

[150] Anonymous report to API Subcommittee on Pressure-Relieving Systems regarding a fire (not a test), June 1941

The following documents are not directly referenced in this International Standard but provide a useful source of relevant information:

Disposal systems

[151] G. ARMISTEAD, Jr., *Safety in Petroleum Refining and Related Industries*, 2nd edition, John G. Simmonds, New York, 1959

[152] C. H. BOSANQUET, W. F. CAREY and F. M. HALTON, *Journal of the Institution of Mechanical Engineers*, Volume 162, Number 3, (1950), London, p. 355

[153] R. M. FRISTROM, *Chemical and Engineering News*, Volume 41, Number 41 (October 14, 1963), pp. 150-160

[154] R. L. GORDIER, *Studies on Fluid Jets Discharging Normally Into Moving Liquid*, Technical Paper Number 28, Series B. St. Anthony Falls Hyd. Lab., University of Minnesota, 1959

[155] B. KARLOVITZ, *Chemical Engineering Progress*, Volume 61, Number 8, (1965), pp. 56-62

[156] W. R. KEAGY, and A. E. WELLER*, A Study of Freely Expanding Inhomogeneous Jets*, Paper Number 89, Heat Transfer and Fluid Mechanics Institute, University of California, Berkeley, 1949

[157] J. F. KEFFER and W. D. BAINES, *Journal of Fluid Mechanics*, Volume 15, (1963), p. 487

[158] D. E. LOUDON, *API Proceedings*, 1963, Volume 43, American Petroleum Institute, Washington, D.C., pp. 418-433

[159] *Manual on Disposal of Refinery Wastes — Volume on Atmospheric Emissions*, American Petroleum Institute, Washington, D.C., 1977

[160] P. D. MILLER, Jr., H. J. HIBSHMAN and J. R. CONNELL, *API Proceedings*, 1958, Volume 38, American Petroleum Institute, Washington, D.C., pp. 276-281

[161] O.A. PIPKIN and C. M. SLIEPCEVICH, *Industrial and Engineering Chemistry Fundamentals*, Volume 3, Number 2, (1964), pp. 147-154

[162] WOHL, K., N. M, KAPP, and C. GAZLEY, *Proceedings: 3rd Symposium (International) on Combustion and Flame Explosion Phenomena*, Madison, Wisconsin, 1948, Williams and Wilkins, Baltimore, 1949, pp. 3-21

Drums and separators

[163] G. D. KERNS, *Petroleum Refiner*, Volume 39, Number 7 (1960), pp. 168-170

[164] E. J. LAPADULA, *Chemical Engineering*, Volume 70, Number 16, (August 5, 1963), pp. 128-129

[165] C. F. MONTROSS, *Chemical Engineering*, Volume 60, Number 10, (1953), pp. 213-236

[166] F. R. NIEMOYER, *Hydrocarbon Processing and Petroleum Refiner*, Volume 40, Number 6, (1961), pp. 155-156

[167] A. D. SCHEIMAN, *Hydrocarbon Processing and Petroleum Refiner*, Volume 42, Number 10, (1963), pp. 165-168

[168] A. D. SCHEIMAN, *Hydrocarbon Processing and Petroleum Refiner*, Volume 43, Number 5, (1964), pp. 155-160

[169] A. D. SCHEIMAN, *Petro/Chem Engineer*, Volume 37, Number 3, (1965), pp. 28-33

[170] A. D. SCHEIMAN, *Petro/Chem Engineer*, Volume 37, Number 4, (1965), pp. 75, 78-79

[171] V. SMITH, *Oil and Gas Journal*, Volume 57, Number 1, (1959), pp. 97-101

[172] R. N. WATKINS, *Hydrocarbon Processing*, Volume 46, Number 11, (1967), pp. 253-256

Flares and stacks

[173] Anonymous, *Ringelmann Smoke Chart*, (Revision of IC 7718), U.S. Department of the Interior, Bureau of Mines, Information Circular 8333, 1967

[174] M. ASLAM, *How to find.... Minimum Length of Process Piping*, *Hydrocarbon Processing*, Nov. 1970, p. 203-208

[175] K. BANERJEE, N. P. CHEREMISINOFF and P. N. CHEREMISINOFF, *Flare Gas Systems Pocket Handbook*, Gulf Publishing Company, Houston, Texas, 1985, ISBN 0-87201-310-3

[176] E. D. BERGMAN,, *The Design of Vertical Pressure Vessels Subjected to Applied Forces*, ASME Paper 54-A-104, *Transactions of the ASME*, August 1955, p. 863-866

[177] P. P. BIJLAARD, *Stresses from Radial Loads and External Moments in Cylindrical Pressure Vessels*, The Welding Journal

[178] R. G. BLICK,, *How to Calculate Wind-Caused Sway on Slender Columns*, *Petroleum Refiner*, Vol. 37, No. 11, Nov. 1958, pp. 237-240

[179] W. C. BLUHM, *Safe Operation of Refinery Flare Systems*, *API Proceedings*, Vol. 41, 1961, American Petroleum Institute, Washington, D.C.

[180] F. T. BOUDURTHA, *Chemical Engineering*, Volume 65, Number 25, (December 15, 1958), pp. 177-180

[181] F. T. BOUDURTHA, *The Behaviour of Dense Stack Gases*, Paper presented at the meeting of the Air Pollution Control Association, New York, 1955

[182] BS 2742C, *Ringelmann Chart*, British Standards Institution, London

[183] BS 2742M, *Miniature Smoke Chart*, British Standards Institution, London

[184] BS 2742:1969, *Notes on the use of the Ringelmann and Miniature Smoke Charts*, British Standards Institution, London

[185] A. A. BROWN, *How to design Pile-supported Foundations for Flare Stacks*, Hydrocarbon Processing, Dec. 1969, pp. 139-146

[186] G. A. CHAMBERLAIN, *Developments in Design Methods for Predicting Thermal Radiation From Flares*, Chem Eng Res Des, Volume 65, July 1987, pp. 299-309

[187] Chemical Manufacturers Association, *A Report on a Flare Efficiency Study*, Washington, D.C., March 1983

[188] L. D. CLEVELAND, Oil and Gas Journal, August 4, 1952

[189] CONCAWE, *The Calculation of Atmospheric Dispersion from a Stack*, CONCAWE, The Hague, The Netherlands, 1966

[190] E. K. DANIELS, J. R. LUTZ and L. A. CASTLER, *API Proceedings*, Volume 38, 1958, American Petroleum Institute, Washington, D.C., pp. 291-292

[191] C.E. FREESE, *Vibration of Vertical Pressure Vessels*, Journal of Engineering for Industry, pp. 77-91, February 1959

[192] Gas Processors Suppliers Association Data Book, Section 5, 10th Edition, Tulsa, Oklahoma, 1987

[193] J. GRUMER, A. STRASSER, J. M. SINGER and P. M. GUSSEY, *Hydrogen Flare Stack Diffusion Flames, Low Flow Instability, Burning Rates, and Dilution Limits*, U.S. Bureau of Mines Preprint, ERC No. 4030, U.S. Department of the Interior, Washington, D.C.,1968

[194] J. D. HAJEK. and E. E. LUDWIG, *Petro/Chem Engineer*, Volume 32, Number 6, (1960), pp. C31-C38

[195] J. D. HAJEK and E. E. LUDWIG, *Petro/Chem Engineer*, Volume 32, Number 7, 1960, pp. C44-C51

[196] J. R. HANNAMAN and A. J. ETINGEN, *Petroleum Processing*, March 1956, pp. 66-69

[197] E. W. HEWSON, *Transactions of the American Society of Mechanical Engineers*, Volume 77, Number 10, 1955, pp. 1163-1172

[198] V. O. HOEHNE, R. G. LUCE and L. W. MIGA, *The Effect of Velocity, Temperature, and Gas Molecular Weight on Flammability Limits in Wind-Blown Jets of Hydrocarbon Gases*, Report to the American Petroleum Institute. Battelle Memorial Institute, Columbus, Ohio, 1970

[199] H. W. HUSA, *Hydrocarbon Processing and Petroleum Refiner*, Volume 43, Number 5, 1964, pp. 179-182

[200] H. W. HUSA, *How to Compute Safe Purge Rates*, Hydrocarbon Processing and Petroleum Refiner, Vol. 43, No. 5, May 1964

[201] HYMES, BOYDELL & PRESCOTT, *Thermal Radiation: Physiological and Pathological Effects*, IChemE Major Hazards Monograph, ISBN 0-85295-328-3 (1996)

[202] J. H. JOHNSON, *Petroleum Refiner*, Volume 26, Number 6, 1947, p. 91

[203] M. R. KELLER and R. K. NOBLE, *RACT for VOC – A Burning Issue*, *Pollution Engineering*, July 1983

[204] J. KELLER, R. K. NOBLE and M. R. KELLER, *Determining Safe Levels of Thermal Radiation Exposure on Personnel*, presented at the 34th Annual Loss Prevention Symposium, AIChE National Meeting, March 6-8, 2000

[205] G. R. KENT, *Hydrocarbon Processing and Petroleum Refiner*, Volume 43, Number 8, 1964, pp. 121-125

[206] G. R. KENT, *Hydrocarbon Processing*, Volume 47, Number 6, 1968, pp. 119-130

[207] J. F. KUONG, *Find Wind-Caused Bending Moment on Tall Vessels by Nomograph*, *Hydrocarbon Processing*, June 1969, pp. 192-195

[208] S. H. KWON, et al, *Improve Flare Management*, *Hydrocarbon Processing*, July 1997

[209] P. D. MILLER, Jr., H. J. HIBSCHMAN and J. R. CONNELL, *API Proceedings, Volume 38*, American Petroleum Institute, 1958, Washington, D.C., pp. 276–281

[210] M. MCDANIEL, *Flare Efficiency Study*, U.S. Environmental Protection Agency Report EPA-600/2-83-052, July 1983, 133 pp, P83-261644

[211] G. B. MOODY, *Mechanical Design of a Tall Stack*, *Hydrocarbon Processing*, Sept. 1969, pp. 173-178

[212] J. H. POHL, *Combustion Efficiency of Flares*, U.S. Environmental Protection Agency, Research Triangle Park, N.C., 1985

[213] R. D. REED, *Chemical Engineering Progress*, Volume 64, Number 6, June 1968, pp. 53-57

[214] R. D. REED, *Furnace Operations*, 3rd Edition, Gulf Publishing Company, Houston, Texas, 1981, ISBN 0-87201-301-4

[215] R. R. ROMANO, *Control Emissions with Flare Efficiency*, *Hydrocarbon Processing*, Oct. 1983

[216] R. E. SCHWARTZ and S. G. KANG, *Effective Design of Emergency Flaring Systems*, Hydrocarbon Engineering, Feb. 1998

[217] R. E. SCHWARTZ and M. R. KELLER, *Environmental Factors Versus Flare Application*, Paper Number 13d, presented at the Symposium on Loss

[218] R. E. SCHWARTZ and M. R. KELLER, *Flaring in Hostile Environments*, paper presented at a Seminar arranged by the Norwegian Society of Chartered Engineers

[219] R. E. SCHWARTZ and J. W. WHITE, *Predict Radiation from Flares, Chemical Engineering Progress*, July 1997, pp. 42-49

[220] *Prevention in the Chemical Industry*, 83rd National Meeting of the American Institute of Chemical Engineers, Houston, Texas, March 22, 1977

[221] *Session on Refinery Flare Design*, API Proceedings, 1952, Volume 32M, American Petroleum Institute, Washington, D.C., pp. 143-164

[222] R. H. SHERLOCK and E. J. LESHER, *Transactions of the American Society of Mechanical Engineers*, Volume 77, Number 1, 1955, pp. 1-9

[223] W. R. SMITH, *Petroleum Processing*, June 1954, pp. 867-880

[224] W. H. SMOLEN, *Petroleum Processing*, Volume 6, Number 9 1951, pp. 978-982

[225] E. J. STANKIEWICZ, *Combustion*, Volume 26, Number 8, February 1955, pp. 51-55

[226] R. S. STEINBOCK, *Chemical Engineering*, Volume 59, Number 2, 1952, pp. 202-203

[227] R. S. STEINBOCK, *Chemical Engineering*, Volume 59, Number 3, 1952, pp. 144-147

[228] R. S. STEINBOCK, *Chemical Engineering*, Volume 59, Number 4, 1952, pp. 154-155

[229] J. F. STRAITZ III, *Improved Flare Design*, *Hydrocarbon Processing*, Oct. 1994

[230] S. H. TAN, *Hydrocarbon Processing and Petroleum Refiner*, 1967, Volume 46, Number 1, pp. 172-176

[231] U.S. Environmental Protection Agency, *Compilation of Air Pollutant Emissions Factors, Vol. 1, Stationary Point and Area Sources*, AP – 42, 4th Edition, 1985 and Supplement F: Section 11.5, 9/91, ‘Industrial Flares

[232] W. R. WICHMANN, A. G. HOPPER and J. L. MERSHON, *Local Stresses in Spherical and Cylindrical Shells due to External Loadings*, Welding Research Council

[233] A. YOUNESS, *New Approach to Tower Deflection*, *Hydrocarbon Processing*, June 1969, pp. 121-126

Flashing flow in pipes

[234] R. J. ANDERSON and T. W. F. RUSSELL, *Chemical Engineering*, Volume 72, Number 25, December 6, 1965, pp. 139-144

[235] R. J. ANDERSON and T. W. F. RUSSELL, *Chemical Engineering*, Volume 72, Number 26, December 20, 1965, pp. 99-104

[236] R. J. ANDERSON and T. W. F. RUSSELL, *Chemical Engineering*, Volume 73, Number 1, January 3, 1966, pp. 87-90

[237] O. BAKER, *Oil and Gas Journal*, Volume 53, Number 12, July 26, 1954, pp. 185-190, 192, 195

[238] O. BAKER, *Oil and Gas Journal*, Volume 56, Number 45, November 10, 1958, pp. 156-157, 159-161, 163, 165, 167

[239] O. BAKER, *et al.*, *AGA/API Report NX-28*, October 1970

[240] J. A. CHAVEZ, *Oil and Gas Journal*, Volume 57, Number 35, August 24, 1959, pp. 100-102

[241] J. M. CHENOWETH and M. W. MARTIN, *Petroleum Refiner*, Volume 34, Number 10, 1955, pp. 151-155

[242] J. COATES and B. S. PRESSBURG, *Chemical Engineering*, Volume 66, Number 18, September 7, 1959, pp. 153-154

[243] A. E. DUKLER, M. WICKS III, and R. G. CLEVELAND, *American Institute of Chemical Engineers Journal*, Volume 10, Number 1, 1964, pp. 38-51

[244] K. E. FIRST and J. E. HUFF, *Design Charts for Two-Phase Flashing Flow in Emergency Pressure Relief Systems*, Paper presented at AIChE International Symposium on Runaway Reactions, March 7-9, 1989

[245] O. FLANIGAN, *Oil and Gas Journal*, Volume 56, Number 10, (March 10, 1958): pp. 132-133, 136, 140-141

[246] O. FLANIGAN, *Oil and Gas Journal*, Volume 57, Number 49, November 30, 1959, pp. 60-66

[247] J. C. LEUNG, *Size Safety Relief Valves for Flashing Liquids*, Chemical Engineering Progress, February, 1992

[248] J. C. LEUNG and M. A. GROLMES, *The Discharge of Two-Phase Flashing Flow in a Horizontal Duct* AIChE Journal, Volume 33, Number 3, March, 1987

[249] J. C. LEUNG, *A Generalized Correlation for One-Component Homogeneous Equilibrium Flashing Choked Flow*, AIChE Journal, Volume 32, Number 10, October, 1986

[250] R. W. LOCKHART and R. C. MARTINELLI, *Chemical Engineering Progress*, Volume 45, Number 1, 1949, pp. 39-48

[251] P. M. PAIGE, *Chemical Engineering*, Volume 74, Number 17, August 14, 1967, pp. 159-164

[252] S. H. RICHTER, *Proceedings — Refining Department*, 1978, Volume 57, American Petroleum Institute, Washington, D.C., pp. 23-35

[253] N. C. J. ROS, *Journal of Petroleum Technology*, Volume 13, Number 10, 1961, pp. 1037-1049

Flashing flow in valves

[254] H. D. BAUMANN, *The Introduction of a Critical Flow Factor for Valve Sizing*, Preprint No. T 34.3.62-1, Paper presented at the 18th Annual Conference of the Instrument Society of America, Chicago, October 1962

[255] L. R. DRISKELL, *Instrumentation Technology*, Volume 14, Number 6, 1967, pp. CVS1–CVS12

[256] A. J. HANSSEN, *Accurate Valve Sizing for Flashing Fluids*, Technical Bulletin Number 108, Conoflow Corp., Blackwood, N.J., 1964

[257] G. R. KENT, *Control Engineering*, Volume 13, Number 5, 1966, pp. 87-92

[258] G. R. KENT, *Control Engineering*, Volume 13, Number 6, 1966, pp. 69-73

[259] C. W. SHELDON and C. B. SCHUDER, *Instruments and Control Systems*, Volume 38, Number 1, 1965, pp. 134-137

Piping

[260] H. A. ALTORFER, *Oil and Gas Journal*, Volume 53, Number 32, December 13, 1954, pp. 126-131

[261] J. E. CONISON, *Oil and Gas Journal*, Volume 52, Number 44, March 8, 1954, pp. 119-120

[262] J. E. CONISON, *Oil and Gas Journal*, Volume 52, Number 45, March 15, 1954, p. 115

[263] J. E. CONISON, *Oil and Gas Journal*, Volume 52, Number 47, March 29, 1954, p. 123

[264] J. E. CONISON, *Oil and Gas Journal*, Volume 52, Number 48, April 5, 1954, p. 123

[265] L. R. DRISKELL, *Petroleum Refiner*, Volume 39, Number 7, 1960, pp. 127-132

[266] *Flow of Fluids Through Valves, Fittings, and Pipe*, Technical Paper Number 410, Crane Co., New York, 1957, pp. 2–14, 2–15, A-19–A-21

[267] C. E. LAPPLE, *Transactions of the American Institute of Chemical Engineers*, Volume 39, 1943, American Institute of Chemical Engineers, New York, pp. 385-432

[268] F. LIPINSKI, *Oil and Gas Journal*, Volume 56, Number 14, April 14, 1958, p. 128

[269] F. LIPINSKI, *Oil and Gas Journal*, Volume 56, Number 18, May 5, 1958, p. 129

[270] F. LIPINSKI, *Oil and Gas Journal*, Volume 56, Number 20, May 19, 1958, p. 187

[271] R. W. MISSEN, *Chemical* Engineering, Volume 69, Number 22, October 29, 1962, pp. 101-102

[272] C. E. PARISH and D. CORNELL, *Petroleum Refiner*, Volume 34, Number 2, 1955, pp. 121-123

[273] R. W. ROBERTS and D. CORNELL, *Petroleum Refiner*, Volume 34, Number 7, 1955, pp. 141-143

[274] N. STESHKO, *Oil and Gas Journal*, Volume 54, Number 41, February 13, 1956, p. 141

[275] N. STESHKO, *Oil and Gas Journal*, Volume 54, Number 46, March 19, 1956, pp. 265-266

[276] N. STESHKO, *Oil and Gas Journal*, Volume 54, Number 47, March 26, 1956, p. 153

[277] S. H. TAN, *Hydrocarbon Processing and Petroleum Refiner*, Volume 46, Number 10, 1967, pp. 149-154

Piping guides and anchors

[278] J. E. BROCK, Heating, *Piping and Air Conditioning*, Volume 36, Number 1, 1964, pp. 152-155

[279] J. E. BROCK, Heating, *Piping and Air Conditioning*, Volume 39, Number 1, 1967, pp. 144-147

[280] C. S. PARKER, Heating, *Piping and Air Conditioning*, Volume 39, Number 11, 1967, pp. 93-100

Pressure relief valves

[281] C. S. BEARD and F. D. MARTON, *Regulators and Relief Valves*, Instrument Publishing Company, Pittsburgh, 1959, pp. 103-130

[282] J. E. BIGHAM, *Chemical Engineering*, Volume 65, Number 3, February 10, 1958, pp. 133-136

[283] J. E. CONISON,, *Petroleum Refiner*, Volume 34, Number 7, 1955, pp. 137-140

[284] F. J. HELLER, *Oil and Gas Journal*, Volume 52, Number 49, April 12, 1954, p. 159

[285] F. J. HELLER, *Oil and Gas Journal*, Volume 52, Number 52, May 3, 1954, p. 111

[286] F. J. HELLER, *Oil and Gas Journal*, Volume 53, Number 3, May 24, 1954, pp. 263-264

[287] J. C. LEUNG, *Size Safety Relief Valves for Flashing Liquids*, Chemical Engineering Progress, February, 1992

[288] J. C. LEUNG and F. N. NAZARIO, *Two-Phase Flashing Flow Evaluations Based on DIERS, API, and ASME Methodologies*, Paper presented at AIChE Loss Prevention Symposium, April 2-6, 1989

[289] P. A. PULEO, *Petroleum Refiner*, Volume 39, Number 10, 1960, pp. 157-162

[290] C. G. WEBER, *Chemical Engineering*, Volume 62, Number 10, 1955, pp. 170-174

[291] C. G. WEBER, *Relief Valve Selecting and Sizing*, Paper presented at the 7th Southeast Regional Instrument Society of America Conference, Charlotte, North Carolina, April 20, 1961

[292] C. G. WEBER, *Oil and Gas Equipment*, Volume 7, Number 11, September 1961, pp. 12-13

Rupture disks

[293] J. E. BIGHAM, *Chemical Engineering*, Volume 65, Number 7, April 7, 1958, pp. 143-145

[294] W. J. BOYLE, Jr., *Chemical Engineering Progress*, Volume 63, Number 8, 1967, pp. 61-66

[295] J. F. W. BROWN, *Transactions of the Institution of Chemical Engineers*, Volume 36, April 1958, London, pp. 81-86

[296] E. DISS, H. KARMA. and C. JONES, *Chemical Engineering*, Volume 68, Number 19, September 18, 1961, pp. 187-188

[297] J. G. LOWENSTEIN, *Chemical Engineering*, Volume 65, Number 1, January 13, 1958, pp. 157-158

[298] J. A. LUKER and M. J. LEIBSON, *Journal of Chemical and Engineering Data*, Volume 4, Number 2, April 1959, pp. 133-136

[299] G. MONDAY, *Chemical Process Hazards with Special Reference to Plant Design*, Symp. Soc. Number 15, Volume 2, 1963, Institution of Chemical Engineers, London, p. 46

[300] R. L. SOLTER, L. L. FIKE and F. A. HANSEN, *API Proceedings*, Volume 43, 1963, American Petroleum Institute, Washington, D.C., pp. 448-456

[301] T. S. MURPHY, *Chemical and Metallurgical E*ngineering, Volume 51, November 1944, pp. 108-112

[302] T. S. MURPHY, *Chemical and Metallurgical Engineering*, Volume 51, December 1944, pp. 99-104

[303] J. A. WEIL, *Transactions of the American Society of Mechanical Engineers*, Series E: Journal of Applied Mechanics, Volume 81, Number 4, December 1959, pp. 621-624

[304] H. R. WRIGHT, *Transactions of the Institution of Chemical Engineers*, Volume 36, April 1958, London, pp. 69-78

Surge and pressure transients

[305] L. BERGERON, *Water Hammer in Hydraulics and Wave Surges in Electricity*, John Wiley & Sons, New York, 1961

[306] R. R. BURNETT, *Oil and Gas Journal*, Volume 58, Number 18, May 2, 1960, pp. 153-160

[307] C. L. COCCIO, *Combustion*, Volume 38, Number 8, February 1967, pp. 18-22

[308] G. A. LINDBERG, *Pipeline Engineer*, Volume 37, Number 3, 1965, pp. 132-135

[309] W. T. ROULEAU, *Transactions of the American Society of Mechanical Engineers*, Series D: Journal of Basic Engineering, Volume 82, Number 4, December 1960, pp. 912-920

[310] I. M. SARLAT and T. L. WILSON, *Transactions of the American Society of Mechanical Engineers*, Series D: Journal of Basic Engineering, Volume 84, Number 3, September 1962, pp. 363-368

[311] A. H. SHAPIRO, *Dynamics and Thermodynamics of Compressible Fluid Flow*, Volume 2, Chapter 24, Ronald Press, New York, 1953

[312] C. W. SIGNOR, American Society of Heating, *Refrigeration and Air Conditioning Engineers Journal*, Volume 2, Number 4, 1960, p. 69

[313] J. F. WILKINSON, D. V. HOLLIDAY and E. F. BATEY, *Oil and Gas Journal*, Volume 63, Number 2, January 11, 1965, pp. 94-95

Systems

[314] API RP 554, *Process control systems*, American Petroleum Institute, Washington, DC

[315] B. BLOCK, *Chemical Engineering*, Volume 69, Number 2, January 22, 1962, pp. 111-118

[316] W. C. BLUHM, *API Proceedings*, Volume 41, 1961, American Petroleum Institute, Washington, D.C., pp. 169-179

[317] P. P. M. BROWN and D. W. FRANCE, *How to Protect Air Cooled Exchangers from Overpressure*, Hydrocarbon Processing, August 1975, pp. 103-106

[318] R. C. CASE, *Proceedings — Division of Refining*, Volume 50, 1970, American Petroleum Institute, Washington, D.C., pp. 1082-1092

[319] Center for Chemical Process Safety (CCPS), *Guidelines for Engineering Design for Process Safety*, American Institute of Chemical Engineers, New York, NY, 1993

[320] Center for Chemical Process Safety (CCPS), *Guidelines for Safe Automation of Chemical Processes*, American Institute of Chemical Engineers, New York, NY, 1993

[321] J. E. CONISON, *API Proceedings*, Volume 43, 1963, American Petroleum Institute, Washington, D.C., pp. 434-447

[322] W. H. DOYLE, *National Safety Congress Transactions*, Volume 5, pp. 30–34; Volume 19, pp. 25-29, 1964

[323] D. W. FOWLER, T. R. HERNDON and R. C WAHRMUND, *An Analysis of Potential Overpressure of a Heat Exchanger Shell Due to a Ruptured Tube*, Paper presented at the American Society of Mechanical Engineers Joint Annual Petroleum Mechanical Engineering Conference, Dallas, Texas, September 22-26, 1968

[324] *General American Tank Car Manual*, GATX, Chicago,1961

[325] D. L. KATZ and N. E. SYLVANDER, *Engineering Research Bulletin Number 31*, University of Michigan Press, Ann Arbor, 1948

[326] D. E. LOUDON, *API Proceedings*, Volume 43, 1963, American Petroleum Institute, Washington, D.C., pp. 418-433

[327] *Plant and Design Safety*, CEP Reprint, American Institute of Chemical Engineers, New York, 1965

[328] H. F. RASE and M. H. BARROW, *Project Engineering of Process Plants*, Chapter 24, Wiley & Sons, New York, 1957

[329] V. H. SUMARIA, J. A. ROVNAK, I. HEITNER and R. J. HERBERT, *Proceedings — Refining Department*, Volume 55, 1976, American Petroleum Institute, Washington, D.C., pp. 631-654

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