Chapter 12

Depressurization, Blowdown and Venting

Hydrocarbon processing facilities pose severe risks with respect to fire, explosions and vessel ruptures. Among the prime methods to prevent and limit the loss potential from such incidents are the provisions of hydrocarbon inventory isolation and removal system. These systems are commonly referred to in the petroleum industry as ESD (emergency shutdown) and depressuring or blowdown. Although most standards and practices acknowledge the need for depressuring capabilities the exact determination of their requirement is not wholly defined. NFPA fire codes and standards rarely mention the subject.

Objective

Typically hydrocarbon process vessels are provided with a pressure safety valve (PSV), to relieve internal vessel pressure that develops above its designed working pressure. The purpose of the PSV is to protect the vessel from rupturing due to overpressure generated from process conditions or exposure to fire heat loads that generate additional vaporization pressures inside the vessel. The engineering calculation behind this application assumes that the process vessel steel strength is unaffected by direct fire exposure causing the increase in pressure. If the vessel is kept at or near it's design temperature this can be assumed the case, however when steel is exposed to a high temperature from a hydrocarbon fire it's capability to contain normal operating pressure deteriorates rapidly sometimes within a few minutes. Since the strength of the material is rapidly deteriorating during this process, regardless of the vessel internal pressure. A rupture of a vessel can easily occur below the operating pressure of the vessel, within minutes of the vessel being exposed to a major heat source.

Pressure safety valves (PSV's) are typically sized to activate at 121% of the working pressure for fire conditions and 110% for the working pressure for non-fire conditions, and only to prevent vessel "overpressure", not to relieve operating pressures. A fire exposure may weaken a process vessel steel strength below the strength needed to contain its normal operating pressure. In this case the vessel may rupture before or during activation of the PSV, when its trying to relieve pressures above operating pressures.

Two major hazards may occur from hydrocarbon pressure vessel failures. The vessel rupture itself and the possible formation of a vapor cloud as a result of the rupture. If the vessel ruptures it will produce flying projectiles and usually release large quantities of flammable vapors. The flying projectiles may contain sufficient momentum to affect other areas. The projectiles could harm individuals or damage the hydrocarbon facility, possibly increasing the incident proportions. Secondly, the released gas from a pressurized vessel may cause a flammable vapor cloud to occur. If some amount of congestion is present or any amount of turbulence of the cloud occurs, an explosive blast may result if the cloud contains enough material and finds an ignition source.
Industry literature typically cites concern with open air explosions when 4,536 kgs (10,000 lbs.) or more of flammable gas is released, however, open air explosions at lower amounts of materials are not unheard of. When the release quantity is less than 4,536 kgs (10,000 lbs.), a flash fire is usually the result. The resulting fire or explosion damage can cripple a hydrocarbon processing facility. Extreme care must be taken to prevent the release of hydrocarbon from vessels resulting in vapor releases and resultant blast overpressure. Measures such as hydrotesting, weld inspections, pressure control valves, adequate pressure safety valves, etc., should all be prudently applied.

To overcome the possibility of a vessel rupture from a hydrocarbon fire exposure several methods are available. Depressuring, insulation, water cooling or draining are usually employed in some fashion to prevent of the possibility of a vessel rupture from it's own operating pressures. A generalized method to qualitatively determine the effect of a hydrocarbon fire on the strength of vessels constructed of steel is available. With this method one can estimate the time for a vessel to rupture and therefore the need to provide protective measures.

API conducted open pool hydrocarbon fire exposure tests (mostly naphtha and gasoline fires), on process vessels during the 1940's and 50's.

This data was plotted using the parameters of

1. Fire exposure temperature.
2. Rupture stress of the vessel.
3. Time for rupture.

These were plotted and are compiled in API RP 520, Chart D-2 (page 55). The data plotted is for vessels constructed of ASTM A-515, Grade 70 steel, a steel typically employed for process vessels. If other materials are used an allowance for their stress characteristics under heat application needs to be made.

Therefore a general determination of the need for protective measures, such as depressurization, can be made for a particular vessel by comparison to the D-2 chart and selected fire exposure temperatures. It should be noted that this is the best available fire test exposure data in the public domain. Improved methods and test data may be available in the future to refine the calculation methods.

Underwriters Laboratories (UL) high rise (hydrocarbon) fire test UL 1709, has an average fire temperature of 1093 °C (2,000 °F) after 5 minutes. Therefore unless the an actual fire exposure heat radiation input calculation has been made, either a worst case fire exposure temperature could be assumed or a standard temperature to the limits of UL 1709 could be applied.

Fireproofing material is considered to fail if any individual thermocouple reaches 649 °C (1200 °F). The API recommended practice does not define the surface temperature from a fire exposure to be applied for the purposes of calculating rupture periods, but provides data from 482 °C (900 °F) to 760 °C (1400 °F) in 38 °C (100 °F) increments to determine rupture times. It should be remembered that free burning fires as a rule do not achieve theoretical combustion temperatures for the fuels involved. Petroleum fires can reach as high as 1300 °C (2400 °F) but average 1000 °C (1850 °F) because of the various factors involved, i.e., cooling of the fire ball, winds, and geometry. Thus some engineering judgment of the arrangement of the vessel involved should be applied in selecting the appropriate fire exposure steel temperature. Typically 649 °C (1200 °F) is chosen as a starting point as this correlates well with fireproofing test requirements. A particular point noticed when using the API chart is that a 100 degree C (212 °F) difference in the fire exposure temperature can have a dramatic difference in the time till a vessel rupture. Therefore the chosen exposure temperature has to be chosen carefully and adequately justified.
The ASME pressure vessel rupture stress formula is applied to calculate a vessel stress is:

\[ S = P(R+0.6t)Et \]

Where:

- \( S \) = Rupture Stress
- \( P \) = Operating Pressure in Psig
- \( R \) = Shell Inside Radius, Inch
- \( t \) = Shell Wall Thickness, Inch
- \( E \) = Weld Joint Efficiency (generally assume 100%)

The shortest time known for a vessel to rupture from recorded incidents is thought to be 10 minutes. Rupture periods calculated for less than ten minutes should therefore not be assumed, as the historical evidence and the typical growth of a hydrocarbon fire would indicate that the immediate rupture of a vessel does not occur. Further investigations may be carried out verity if fire exposure conditions could produce such results (e.g., flange leak, gas fire exposures, etc.).

If vessel is insulated, some credit can be taken on the reduced heat input rate provided by the insulation, but this depends upon the quality and thickness of the insulation, plus the time for the insulation to raise the ambient exposure temperature. Typically in sizing relief valves, it is normally assumed that lightweight concrete insulation (fireproofing) reduces the heat input to approximately one third of its original value. Therefore depending on the rating of the fireproofing, the time till a vessel rupture from operating pressures can be increased. The time delay of the fireproofing material can be added to the time it takes to cause the steel to weaken and rupture. Commercially available hydrocarbon fire rated fireproofing materials are available in several hours of fire resistance periods. If connecting pipelines are not isolated with an ESD valve or insulated from fire sources, they could also be a source of hydrocarbon release that have to be taken into account when making these assumptions.

Similarly if a vessel is provided with a reliable and dependable water cooling (i.e., firewater deluge water spray), according to recognized standards, (e.g., NFPA 15, Water Spray Systems for Fire Protection), that would not be affected by an explosion blast pressures or the fire exposure, it may theoretically demonstrate that a vessel does not need a depressuring system for the prevention of rupture from fire exposures. Similarly API RP 2000 does not allow credit for water cooling sizing pressure relief valves unless they are demonstrated to have extremely high integrity during accidental events.

If the area under the vessel is provided with adequate drainage capability credit may also be taken for a reduced heat input due to the runoff of any flammable liquids producing the fire exposure. Usually drainage requirements to NFPA 30 (Flammable and Combustible Liquids Code), would have to be met, namely 1 percent to a 15.2 meter (50 ft.) radius. Published literature suggests that an uninsulated vessel rupture time could be increased 100% for a highly effective drainage system.

Two examples on the technique to calculate vessel rupture periods have been prepared and are shown in Figures 4 and 5.
SEPARATOR (Horizontal)

ASSUMPTIONS:

Size: 10'-0" I.D. x 50'-0" s/s
Shell Wall Thickness: 1/2"
Liquid Sp. Gr.: 1.0
Material of Construction: A515 Gr. 70
Operating Pressure: 50 Psig
Design Pressure: 90 Psig
Normal Liquid Level: 5'-0" from bottom

\[ S = \frac{P (R + 0.6t)}{Et} \]

Where: 
- \( S \) = Rupture Stress
- \( P \) = Operating Pressure in Psig
- \( R \) = Shell inside radius, Inch
- \( t \) = Shell Wall Thickness, Inch
- \( E \) = Joint efficiency (assumed 100%)

From Figure D-2 (API 520, page 55)

Time before rupture at 6,030 psi and 1,300 deg. F is approximately 5 Hrs.

CONCLUSION: Depressurization system is not required.

HORIZONTAL SEPARATOR

Figure 4
Depressurization, Blowdown and Venting

**CRUDE STABILIZER**

**ASSUMPTIONS:**
- Size: 5'-0" I.D. x 40'-0" s/s
- Shell Wall Thickness: 7/16"n
- Liquid Sp. Gr.: 0.85
- Material of Construction: A515 Gr. 70
- Operating Pressure: 150 Psig
- Design Pressure: 175 Psig
- Normal Liquid Level: 5'-0" from bottom seam
- Vessel is insulated but no credit given for insulation

\[ S = P \left( R + 0.6t \right) E \]  
(Ref. ASME, DIV. VIII FOR CIRCUMFERENTIAL STRESS)

\[ S = 150 \left( 30 + 0.6 \times 0.4375 \right) / 1.0 \times 0.4375 \]

Where:
- \( S \) = Rupture Stress
- \( P \) = Operating Pressure in Psig
- \( R \) = Shell inside radius, inch
- \( t \) = Shell Wall Thickness, inch
- \( E \) = Joint efficiency (assumed 100%)

From Figure D-2 (API 520, page 55)

Time before rupture at 10,374 psi and 1,300 deg. F is approximately 0.3 hours

**CONCLUSION:** Depressurization system is required.

**CRUDE STABILIZER**

**Figure 5**
Once a time period has been established when a vessel might be expected to rupture, this needs to be compared against the WCCE for the facility. A very short duration fire exposure would imply that a vessel depressurization may not be necessary. Typically most hydrocarbon facilities are provided with an ESD system, which at the very minimum should isolate the incoming and outgoing pipelines. In this fashion the remaining major fuel inventory at the facility is what remains in vessels, tanks and the piping infrastructure. It should also be considered that after 2 to 4 hours of a hydrocarbon high temperature fire, usually equipment or facility salvage value is minimal. So beyond these periods, little value is gained in additional protection measures. Typically if the rupture period is of a magnitude of several hours, the need for depressuring (or blowdown) is not highly demonstrated or recommended.

Normally, emergency vessel depressuring is to be automatically activated through a facility ESD level 1 (worst case) interface and completed in less than 15 minutes. A vessel should be depressurized to a minimum of 50% of its design operating pressure or preferably completely depressurized, remaining interconnecting vessels are also depressurized. If vessels are not completely depressurized, there is still a risk of vapor release from the remaining pressure (i.e., fuel inventory) in the vessel or piping. An engineering evaluation of depressurization arrangements and calculation of depressuring periods should be performed.

Certain conditions and arrangements (process restarts for example) may preclude the provision of an automatic and immediate depressuring system for all vessels. Some volumes of gaseous products may be necessary for an adequate process restart. If the facility where to accidentally depressurize, the operation may suffer an economic business interruption loss if gas supplies have to be obtained from outside the facility. In these cases alternative protection methods may be employed. Where vessels are not located in dense processing areas, subject to adjacent impact time delay and local fusible plug activated depressuring outlet valves, insulation (fireproofing), dedicated firewater deluge, adequate and immediate drainage, etc., should all be considered. Remote placement of the subject vessel is another yet costly alternative. An engineering analysis should be performed when a fully automatic (ESD) depressuring system is not provided.

Published literature also suggests that explosions and major damage are very unlikely when less than one ton of material is released. API RP 520 additionally suggest that vessels operated at or below 690 kPa (100 psi) typically are not provided with depressurization capability.
Depressurization, Blowdown and Venting

The following is a general guideline that may be used to generally classify which process vessels may need depressurization capabilities:

**Vessels Requiring Depressurization Capability:**

- A vessel operated above 690 kPa (100 psi).
- The vessel contains volatile liquids (e.g., butanes, propanes, ethanes, etc.) with vapor pressures above atmospheric.
- Operational requirements exist (e.g., compressor blowdowns).
- A fire condition may occur that weakens a vessel to below safe strength levels (as defined by API RP 520, Part 1, figure D-2), within several hours, which may cause significant exposure losses.

**Vessels Which May Not Require Depressuring Capability:**

- A vessel operated at or less than 690 kPa (100 psi).
- A vessel containing less than 907 kgs (2,000 lbs.) of vapors.
- A vessel whose time to rupture from a fire exposure is more than several hours.
- A vessel provided with fireproofing insulation rated to withstand the expected fire exposures until other protection measures are employed (e.g., effective manual fire fighting is available).
- A vessel provided with a firewater deluge system to protect against hydrocarbon fire exposures for the duration the worst case plausible incident.
- A vessel whose time to rupture, insulation, fixed fire water protection or drainage arrangements would not cause the vessel to rupture during the process incident.
- A vessel, which if a rupture occurred due to a fire exposure would not endanger personnel, damage important or critical facilities, cause significant financial impacts, create an environmental hazard or create an undesirable reaction from the general public.

The objectives of depressuring are (1) to prevent a vessel from rupturing during a major fire exposure (from the weakened condition of the vessel steel), (2) to prevent further fire escalation and (3) to minimize the impacts to the vessel itself. It is therefore incumbent to depressure a vessel so that its stress is less than the stress to cause a rupture from fire conditions. These stresses and rupture periods can be estimated to determine the need for depressuring systems for hydrocarbon vessels. These estimates can provide a rough estimate for the need of a depressurization system for a particular hydrocarbon process.

Vapors from depressuring valves are typically routed to a blowdown header and then to a flare to safely remove the vapors from the area and dispose of it without impact to the environment. A special concern when high levels of pressurized gases are released into a piping system is the possibility of auto-refrigeration of the piping material that may cause a brittle fracture. A process engineer should verify which pipe materials and flow rates, specified for the depressurization system, are suitable for the pressures, flows and gases contemplated.

Once calculations are completed on a depressurization system it will become readily apparent high volumes of gases will be flowing through the header to a flare. In some cases the practicality of simultaneously depressurizing all of the process equipment and vessels will be difficult to accomplish. In these cases a sequential blowdown of the vessels should be considered. Providing for the "worst" vessels first or controlling the system to blowdown the area most affected first are desirable options.

High noise levels may also be generated when high flows are encountered. In these circumstances, special noise reducing fittings are available to limit noise impacts from the system to the surrounding area.
Figure 6

PROCESS VESSEL DEPRESSURIZATION
FLOW CHART

Figure 6
Blowdown

Blowdown is the removal of the liquid content of vessels and equipment to prevent their contribution to a fire or explosion incident. Blowdown is similar to depressurization but entails liquids instead of gases. A liquid blowdown should never be sent to the facility flare that is designed to only handle gaseous materials. A liquid release out of the flare may result in a flare out, and if the flare is elevated a shower of liquids onto the process facilities. Ideally liquid blowdowns should be routed to facilities that are specifically designed to handle large quantities of liquid materials. The blowdown could be routed to a storage tank, open pit, another process area or the pressurized sewer. A blowdown to a tank is generally avoided since entrained gases may cause the tank to rupture. Similarly disposal to a pit is not desired as it poses the hazard of exposed combustible liquids that can ignite.

Venting

Direct venting of hydrocarbon or toxic gases to the atmosphere should be avoided for the following reasons.

1. It may create a combustible vapor cloud with fire or explosion possibilities.
2. It may be harmful to personnel.
3. It may be an environmental pollutant.
4. It is a waste of the fuel gas.
5. It represents a poor community or public image to release wastes to the atmosphere.
6. It may be a violation of the national or local environmental governmental regulations.
7. Remotely vented gases may not adequately disperse, then drift considerable distances and ignite.

Whenever possible waste vapors or gases should be disposed of through the facility flare system or reinjected into the production process for recovery. Non polluting materials such as steam can be freely vented to atmosphere if they do not pose burn hazards to personnel.

Flares

In most hydrocarbon operations excess gas and vapors have to disposed of safety, quickly without environmental impact. Where the gas or vapor cannot be converted into useful energy they are routed to a remote point for safe incineration, called flaring. Flares are the most economical and customary means of disposing of excess light hydrocarbon gases in the petroleum and chemical industries. The primary function of a flare is to convert flammable, toxic or corrosive vapors to environmentally acceptable gases for release into the atmosphere. Both elevated or ground flares can be used.

The type of flare used depends on several factors including:

1. Available space of onshore or offshore arrangements.
2. Characteristics of the flare gas - composition, quantity, pressure, etc.
3. Economics both initial capital costs and periodic maintenance.
4. Public impression (i.e., if flaring is smoky or noisy the general public will object to its operation).

The primary features of a flare are safety and reliability, while the primary objective of the flare is to prevent the release of any unburned gases. In reviewing existing facilities worldwide, from Russia to South America, onshore and offshore, most installations have admitted either officially or unofficially that on occasion, a liquid release has occurred from the tip of the flare stack. This has occurred even with the installation of a flare header liquid knock out drum. In most cases this has caused no apparent problems although in a few
cases it has been disastrous. It is suspected liquid releases occur much more frequently than is actually reported. Technically the problem may be because most flare systems are designed for unrestricted gas flow through the flare header and knock out drum, but which induce liquids to carryover. Therefore the possibility of liquid releases from vapor disposal flares cannot be entirely ruled out.

During a typical plant design, the flare location is usually examined. Some experts suggest that a flare should be located downwind and while others propose it should be upwind of the facility. This based on the assumption that a flare may overflow with liquids and therefore should be downwind so these don't fall on the plant, while vice versa, a plant gas release will travel downwind and be ignited. The solution of course, is to locate the flare perpendicular to the prevailing wind with adequate spacing from the facility. This avoids both the vapor dispersions from the flare and liquid releases onto the plant. Preferably the flare should also be at a lower elevation than the rest of the facility. This is in case it releases heavy vapors that have not been properly combusted in the flare exhaust. Because of the larger spacing distances and risk factors associated with flares they should be one of the first items sited for a new facility.

Flare safety precautions should include:

1. Use of an automatic flame monitoring device to warn of flameout conditions.
2. Provision of a liquid knock out (KO) drum, which is equipped with high level alarms to warn of an excessive accumulation of liquid.
3. Prevention of the introduction vapors into the system when it is not operational.

The important safety aspects of flares include the following:

a. The flare is a readily available ignition source to vapors that can reach it or the radiant heat it produces.

b. Flame-out (flame lift-off or blow-outs), sometimes occurs at a flare, at which time flammable vapors will be discharged. If heavier than air and wind conditions permit, they will travel along the ground to other areas until dissipated. Provision of a windshield around the flare tip will assist in prevent a flame-out from occurring.

c. Flare may emit liquids under certain conditions, which even if a flare is lit it can endanger processes that are placed to close to it. Provisions to entrap and contain liquids in the flare header, for worst case conditions should be provided at the flare tower.

Liquid separators or knockout drums are normally used to remove any liquid from gas streams flowing to flares designed to burn vapors. The drums should be designed not only to collect liquids running along the bottom of the pipe, but to disengage entrained liquid droplets. API RP 521, Section 5.4.2.1 recommends that particles 300 to 600 micrometers in diameter or larger should be removed before flaring the gas. Additionally the knockout drum should be sized to accommodate the maximum amount of liquid that might be required to be withdrawn during depressurization of the entire or any portion of the facility as the design of the system may dictate. If large quantities of propanes and butanes low temperatures may be reached in the flare header and drum due to auto refrigeration, which must be taken account of.
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**Table 15**

General Guidelines for Material Disposal Methods
Bibliography


