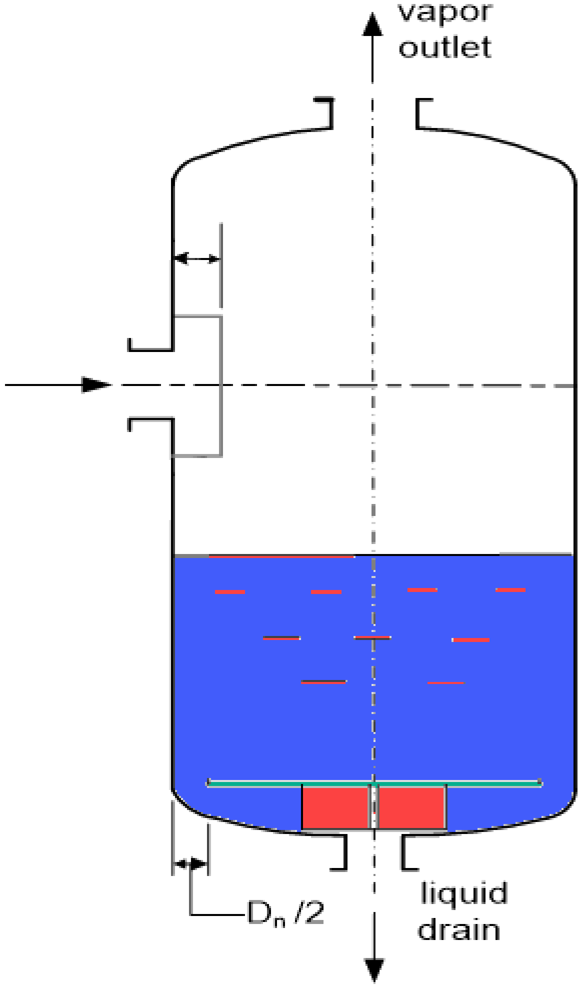
# Vent Containment Design for Emergency Relief Systems

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# Vent Containment Design for Emergency Relief Systems

*Pressure Relief and Effluent Handling Practices*

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# 1. Why Consider Effluent Handling and Vent Containment?

Emergency relief discharge from a chemical reactor or a process storage vessel will often need treatment before it can be vented to atmosphere. The discharge of flammable and/or toxic materials can create fire and explosion hazards as well toxicity hazards both on site and off site. Potential environmental impact can also lead to large cleanup and restoration costs.

Two-phase discharges are more challenging than all vapor discharges. Two-phase discharges can lead to the formation of aerosols and heavier than air clouds. More mass is airborne and as a result the resulting dispersion footprints are larger. Rainout of toxic and/or flammable materials (small liquid droplets) can occur near the release point and/or downstream of the release point.

Two-phase relief from chemical reactors and process vessels must be discharged to a safe location [2]. This can be particularly challenging for plants that are congested and/or that are near population centers. Although it is possible in some cases to eliminate two-phase relief by design or to discharge a two-phase effluent directly to the atmosphere, the use of vent containment can provide a more versatile approach to reducing two-phase relief risks.

# 2. What Technologies are Available?

A variety of vent containment and effluent handling technologies are available including disposal, collection, and treatment. Discharge of gases or vapors to atmosphere, burning of gases or vapors via flaring, and discharge of liquids to sewers represent common disposal strategies. Condensation of vapors in a quench vessel [1], containment of liquid for further processing, and vapor recovery systems represent some of the collection strategies used. Reacting of a hazardous effluent chemical component in a scrubber is an example of a treatment strategy.

Common effluent handling and vent containment equipment include vapor-liquid separators, scrubbers, quench vessels or pools, vent stacks, and flares. Vapor-liquid separation can be driven by gravity, vane impingement, or centrifugal force (cyclone).

# 3. General Considerations

In order to select and design an effective effluent handling and vent containment system we need to consider the quantity discharged and the rate at which the discharge occurs, fluid properties, and site specific conditions. Relief systems flow rates should be developed using what the actual [2] relief device will flow without any design derating factors or code flow safety factors used to select the relief device.

In addition to fluid quantity and rate of discharge, we need to consider if the effluent is toxic, corrosive, or presents other health hazards and/or fire and explosion hazards. Nuisance and potential environmental impact can also influence the design.

[1] Also commonly referred to as quench pool

[2] This is opposite of relief device sizing philosophy.

Fluid properties such as foaming, freezing, viscosity (polymers) can significantly influence the design and effectiveness of the vent containment system. Site specific conditions to consider include weather conditions, plant layout and constraints, overall geography and topography to name a few.

Foaminess (foam breakdown time) and liquid viscosity are two physical properties that strongly influence effluent handling systems design. For slightly viscous liquids ( $< 100$ cp) most effluent handling systems will work. For moderately viscous liquids (100 to 1000 cp), the trapped bubbles expand the liquid volume. As a result a two to three times accumulation volume is required for separators. High viscosity liquids ($> 1000$ cp) are best handled using quench vessels. In some cases small scale performance tests may be required for high viscosity liquids.

Relatively stable foams are best handled using quench vessels, although quench vessels are not recommended for systems with non-condensible vapors. Gravity phase separators are not suitable for relatively stable foams. Low stability foams can be effectively handled with quench vessels and gravity phase separators. A gravity phase separator with two to three times the clear liquid volume can be used for low stability foams. A cyclone separator is generally better than gravity separators for foamy fluids.

Liquid Separation efficiency depends on the size of liquid droplets in the vapor/liquid stream entering the effluent handling system. In general, 150 $\mu$m (microns) droplets can be removed by most effluent handling systems as shown in Table 1.

Table 1: Typical droplet size removal capabilities

| Droplet Size, $\mu$m | 1 | 5 | 10 | 50 | 150 |
|---|---|---|---|---|---|
| Gravity - Vertical | No | No | No | No | Yes |
| Gravity Horizontal | No | No | No | No | Yes |
| Vane Impingement | No | No | Yes | Yes | Yes |
| Cyclone | No | No | possibly | Yes | Yes |
| Flare | Yes | Yes | Yes | Yes | Possibly |
| Quench Vessel | No[1] | Possibly[1] | Possibly[1] | Possibly[1] | Yes |
| Scrubber | Possibly[2] | Possibly[2] | Possibly[2] | Possibly[2] | Yes |

[1]For vented quench vessel (no entrainment for non-vented quench vessels).
[2]Packed columns are very effective at removing small droplets. Columns with trays are less effective.

Table 2 illustrates typical ranges of liquid droplet sizes depending on the process condition.

Table 3 illustrates the expected liquid removal efficiency of effluent handling equipments for a 200 micron liquid droplet size.

# 4. Effluent Handling and Vent Containment Strategies

One of four effluent handling and vent containment strategies can be used depending on the relief systems discharge type, phase, and potential hazards. Emergency relief systems discharges can

Table 2: Typical droplet size vs. process condition

| Process Condition | Droplet Size, $\mu$m |
|---|---|
| Condensation with fogging | 0.1 - 30 |
| Gas atomization spray nozzle | 1 - 100 |
| Gas bubbling through boiling liquid | 20 1000 |
| Annular two-phase flow through pipe | 10 - 2000 |
| Liquid spray nozzle | 1000 - 5000 |

Table 3: Typical effluent handling equipment removal efficiency

| Type of Device | Typical Removal Efficiency (200 $\mu$m) |
|---|---|
| Gravity - Vertical | $> 90\%$ |
| Gravity - Horizontal | $> 90\%$ |
| Vane Impingement | $> 99\%$ |
| Cyclone | $> 98\%$ |
| Absorber | $> 99\%$ |
| Quench Vessel | $> 98\%$ |

consist of (a) all vapor flow, (b) two-phase vapor-liquid flow and/or (c) three-phase vapor-liquid-solid flow.

Various combinations of separators, flares, stacks, scrubbers are often considered (see Figure 1). It is important to note that most of the separation equipment will work well for low viscosity and non-foamy materials while quench vessels can work well for high viscosity and foamy materials.

Figure 2 illustrates a selection strategy for effluent handling and vent containment equipment that is recommended by CCPS [1]. It can be seen from Figure 2 that CCPS suggests directing emergency relief effluent to some form of containment/mitigation on a widespread basis.

### 4.01. Direct Discharge to Atmosphere

This is typically adequate for an all vapor discharge via a tall stack. The stack height is often determined using the peak flow rate to prevent ground level toxicity/flammability impact. The presence of nearby buildings and other structures/platforms where people may be working or living should be taken into consideration in any dispersion analysis that is performed to determine the required stack height [2]. Because of the short duration of emergency relief discharges, the use of time dependent flow rates, time dependent dispersion analysis, and AEGL [3, 4] thresholds can lead to more realistic and practical stack heights. Ignition of flammable vapors exiting the stack due to lightning and/or static should be considered in determining the final safe discharge stack height and location.

Figure 1: Typical effluent handling and vent containment systems configurations

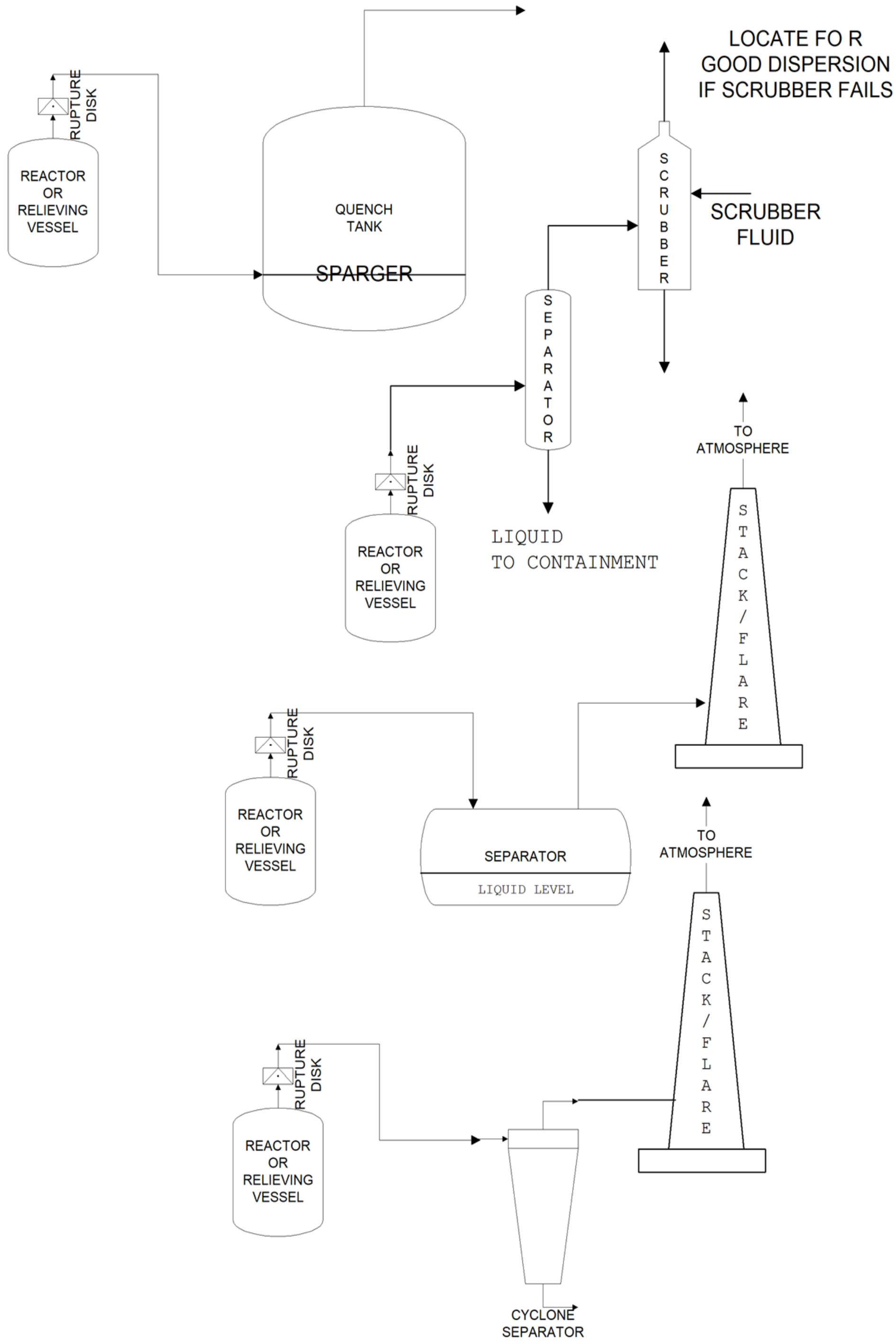
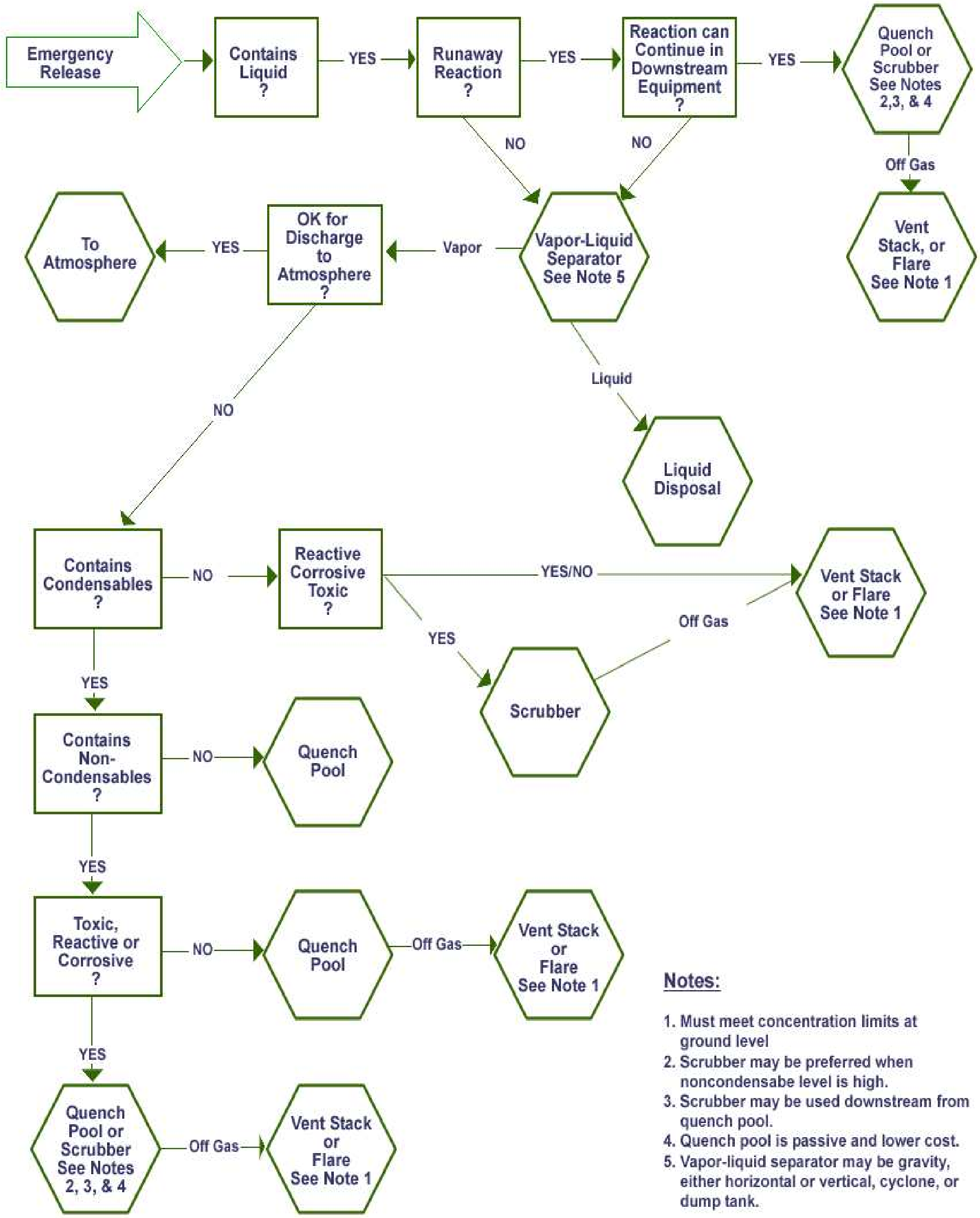


Figure 2: Recommended strategy for the selection of effluent handling and vent containment systems [1]



### 4.02. Flaring and Incineration

This is also recommended for all vapor flow because the presence of liquid droplets or mist can decrease the burning efficiency. The main premise of incineration/flaring is to turn chemicals into less hazardous combustion products. For example, the combustion of $CH$ compounds will lead to the formation of $CO$, $CO_2$ and $H_2O$. Thermal radiation calculations are used to set the flare height using criteria published for personnel and equipment exposure by API [5] and CCPS [1]. Additional dispersion and unconfined vapor cloud explosion calculations are also required in case the flare flames out.

### 4.03. Separation and Partial Containment

The main principle of this type of effluent handling and vent containment strategy is to first separate and contain the liquid and then send the vapor for further mitigation and risk reduction. Typical separation equipment that are used include:

1. Gravity horizontal separators
2. Gravity vertical separators
3. Cyclone separators
4. Open and partially vented quench vessels
5. Impingement separators

### 4.04. Total Containment

Under this type of effluent handling and vent containment strategy, another vessel with a large volume is used to catch and contain the emergency relief systems discharge. The large vessel volume is necessary to contain the effluent and to reduce backpressure on the relief system. Two types of total containment vessels can be used:

1. Vent vessel/tank. This is mostly used for systems where large quantities of highly toxic non-condensible gas are produced, usually from a runaway chemical reaction and/or decomposition.
2. Quench vessel/tank (closed). This is used for systems where the pressure is driven by the relief effluent liquid temperature (system vapor pressure) and the generation of condensible vapors. A quench liquid is used as a heat sink to cool, condense, dilute, and sometimes neutralize the relief effluent. This helps to reduce the overall system pressure and to slow down and/or stop chemical runaway reactions.

# 5. Phase Separation and Partial Containment

A phase separator traps the liquid portion of the relief system discharge. The potential hazard is mostly driven by the liquid portion of the discharge. Phase separators include gravity separators and cyclones. Separators can handle low to moderate flow rate and non-foamy liquids.

Sizing of separators depends on the vapor phase flow rate, the liquid phase flow rate, the liquid viscosity, and foaming tendency, and available space in the plant.

## 5.1 Cyclone Separators

A cyclone (see Figure 3) is a vertical cylindrical vessel with a tangential inlet and an internal concentric shroud (skirt). The combined effect of the tangential inlet and the internal concentric shroud creates a strong centrifugal force field for effective liquid separation. A cyclone is suitable for superficial vapor inlet velocities up to and including sonic velocity, and liquids ranging in viscosity from water-like liquids to those approaching molasses.

Phase separation occurs in cyclones because the tangential flow entry into the cyclone produces a spiraling flow pattern and a centrifugal force to drive droplets to the outer wall (see Figure 3). Upon entering the vessel, the bulk of the liquid collects on the outer wall, and both vapor and the liquid follow spiral paths downward. Part-way down the vessel, the vapor and the liquid mist carried by it, continue in an orbital path, but rise in an inner spiral toward the exit shroud. This tightening of the spiral intensifies the centrifugal force and excludes all but the very fine mist from leaving the cyclone separator.

A small fraction of the liquid spirals up the separator wall from the inlet, creeps across the underside of the top head, and drains down the outside of the internal shroud. It drips off the shroud without being re-entrained, provided that the superficial vapor velocity is maintained within the recommended design limits. The skirt around the outlet nozzle prevents re-entrainment of the liquid film climbing up the cyclone outer wall onto the cyclone top head. A sufficiently low superficial gas velocity inside the skirt allows the liquid film to drip off without re-entrainment. Either a sufficient liquid accumulation volume is added to the cyclone bottom (see Figure 4) or a vortex breaker is used to facilitate liquid removal to another vessel (Figure 5).

### 5.11. Cyclone Separator Design Procedures

These procedures deal mostly with the design of cyclones in the context of emergency relief systems design. The following information is required for the design:

- Transient volumetric flow rates of gas and liquid for the entire release duration
- Vapor and liquid densities
- The volume of liquid to be retained in the cyclone
- Degree of foaminess (foam breakdown time)

- Approximate liquid viscosity
- Operating pressure (usually slightly above ambient)

Cyclone inlet and outlet nozzles and piping should be determined based on an evaluation of the entire pressure drop throughout the entire relief system which includes the cyclone. The total cyclone pressure drop should include the cyclone body pressure drop and the cyclone outlet line pressure drop. Cyclone vapor outlet nozzle and piping can be selected such that the pressure in the relieving vessel does not exceed the maximum allowable pressure accumulation.

The cyclone inlet nozzle size is selected such that the flow inlet velocity is in the range of 30 to 45 m/s (100 to 150 ft/s). If the flow is choked in the relief device discharge line, the cyclone inlet nozzle should be the same size as the line size. The cyclone separator will also work at higher velocities although higher velocities are not recommended because of potential acoustic induced vibration risk to the piping. If the flow is not choked in the relief discharge line then the velocity can be reduced by making the inlet nozzle larger. Cyclone outlet nozzles are sized such that the vapor flow velocities are in the range of 15 to 30 m/s (50 to 100 ft/s).

Liquid can be removed from the cyclone during venting or after venting has stopped. This can be done by pumping or gravity. The liquid exit nozzle should be sized such that liquid velocities between 1 and 3 m/s are possible. When the liquid is pumped out, NPSH limitations should be considered.

The impact of continuing reaction and flashing due to pressure drop in the cyclone are also important and warrant consideration. More vapor may leave the cyclone than what enters due to flashing caused by pressure drop. The inlet velocity and cyclone pressure drop are limited by the allowable back pressure on the relief device. Increasing the allowable cyclone pressure will often result in smaller cyclones.

Cyclone separator design according to the procedures outlined in this document will typically remove liquid droplets that are larger than 100 microns when operating at peak inlet velocities. Liquid removal efficiencies in the range of 80 to 99 % have been observed experimentally.

A variety of cyclone and catch vessel arrangements are used depending on the application and availability of space.

## 5.12. Cyclone Separator with Integral Catch Vessel

This arrangement, shown in Figure 4, is similar to the one described in Section 5.13. below, except that the cyclone and catch vessel are combined in one vessel shell. This design is used when the vapor rate is quite high so that the cyclone diameter is large.

## 5.13. Cyclone Separator with Separate Catch Vessel

This arrangement, shown in Figure 5, is frequently used in chemical plants where plot space is limited. The cyclone performs the vapor-liquid separation, while the catch vessel accumulates the

Figure 3: Cyclone separator

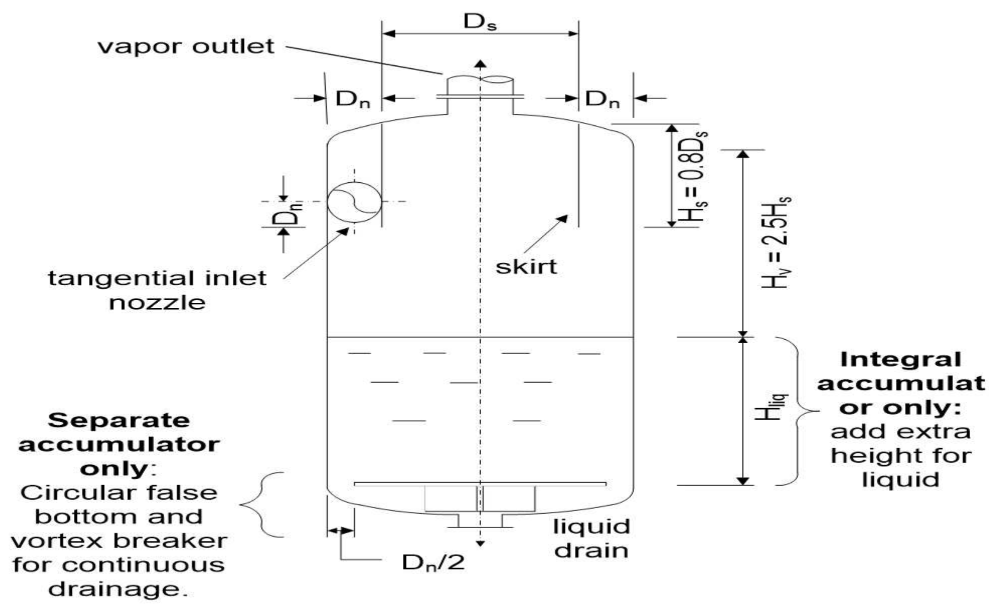



Figure 4: Combined cyclone separator and catch vessel

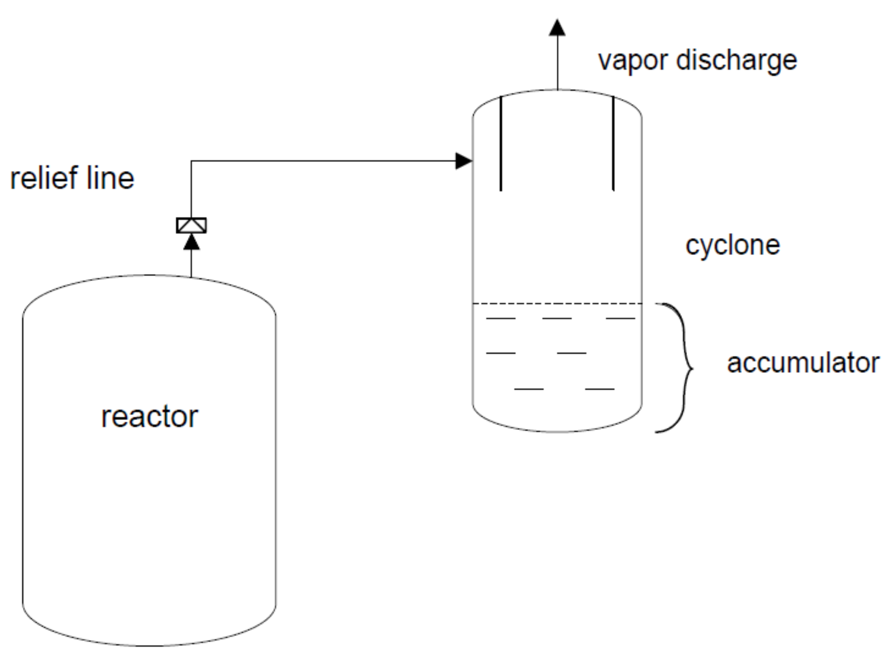
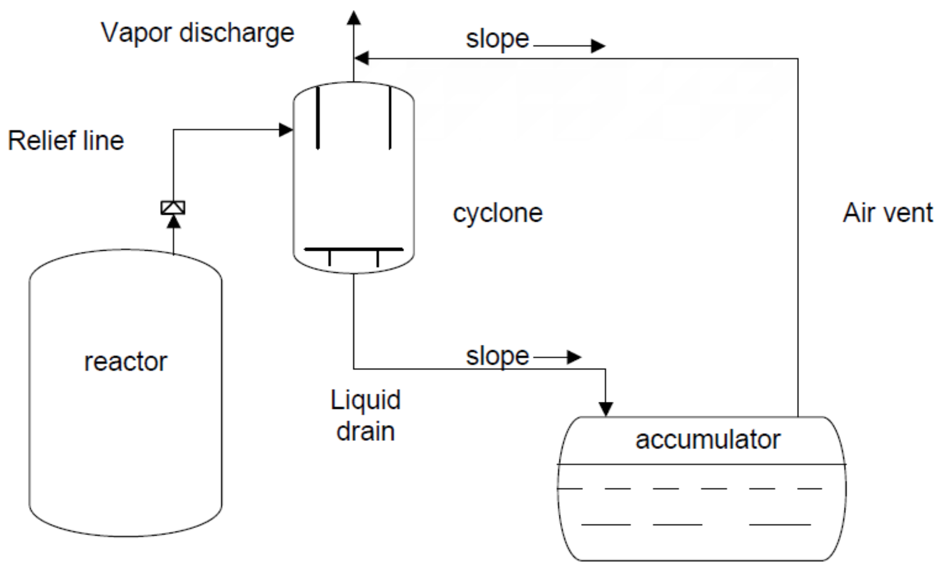


Figure 5: Cyclone separator with separate catch vessel




liquid from the cyclone. This arrangement allows location of the cyclone separator close to the reactor so that the length of the relief device discharge line can be minimized.

### 5.14. CCPS Cyclone Separator Design Method

The following step by step design procedure is based on several publications by Grossel [6] and is known as the CCPS cyclone design method:

1. Select a superficial F factor to be used for calculating the skirt diameter (see 6). Reported values for F in SI are 10 for water like fluids and 6 for molasses-like fluids (viscosity around 1500 cp).

2. Calculate the skirt mass velocity, area and diameter using the superficial F factor:

$$G_s = F\sqrt{\rho_v} \tag{1}$$

$$A_s = \frac{W_v}{G_s} \tag{2}$$

$$D_s = \sqrt{\frac{4A_s}{\pi}} \tag{3}$$

where $G_s$ is the skirt mass velocity in $(\mathrm{kg/m^2/s})$, $\rho_v$ is the vapor/gas density at cyclone outlet in $(\mathrm{kg/m^3})$, $A_s$ is the skirt cross sectional area in $(\mathrm{m^2})$, $D_s$ is the skirt diameter in (m) and $W_v$ is the maximum vapor/gas flow rate at the cyclone outlet in (kg/s) including vapor generated by liquid feed flashing.

3. Calculate other cyclone dimensions (as shown in Figure 6).

$$H_s = 0.8D_s \tag{4}$$
$$D_v = D_s + 2D_p \tag{5}$$
$$H_v = 2.5H_s \tag{6}$$

where $D_p$ is the inlet nozzle diameter (m), $H_s$ is skirt height, $D_v$ is the vessel diameter (m), and $H_v$ is the vessel straight side height. If the liquid is continuously drained to a separate accumulator, a vortex breaker and false bottom should be used. If the accumulator is integral with the cyclone, then $H_v$ must be increased to provide the required volume for liquid collection.

$$H_t = H_v + H_l = H_v + \frac{4V_l}{\pi D_v^2} \tag{7}$$

where $H_t$ is the total vessel straight side height in (m), $H_l$ is the liquid height to be added to $H_v$, and $V_l$ is the liquid volume collected in $(\mathrm{m}^3)$.

4. Calculate the cyclone pressure drop. This is approximately five velocity heads (includes inlet and outlet nozzle):

$$\Delta P = \frac{2.5G_v^2}{\rho_v} \tag{8}$$

where $\Delta P$ is the total unrecoverable pressure loss, including inlet/outlet losses, in (Pa), $G_v$ is the inlet mass flux of vapor in $(\mathrm{kg/m^2/s})$ and $\rho_v$ is the inlet vapor density in $(\mathrm{kg/m^3})$. If the pressure drop is too high, it may be necessary to increase the inlet line size for a distance of three to five diameters upstream of the cyclone. This pressure drop equation is only valid for when the pressure loss is less than 20 % of the inlet pressure. For higher pressure drops, the use of a flow resistance coefficient of five velocity heads is recommended so that compressibility of the vapor is taken into account.

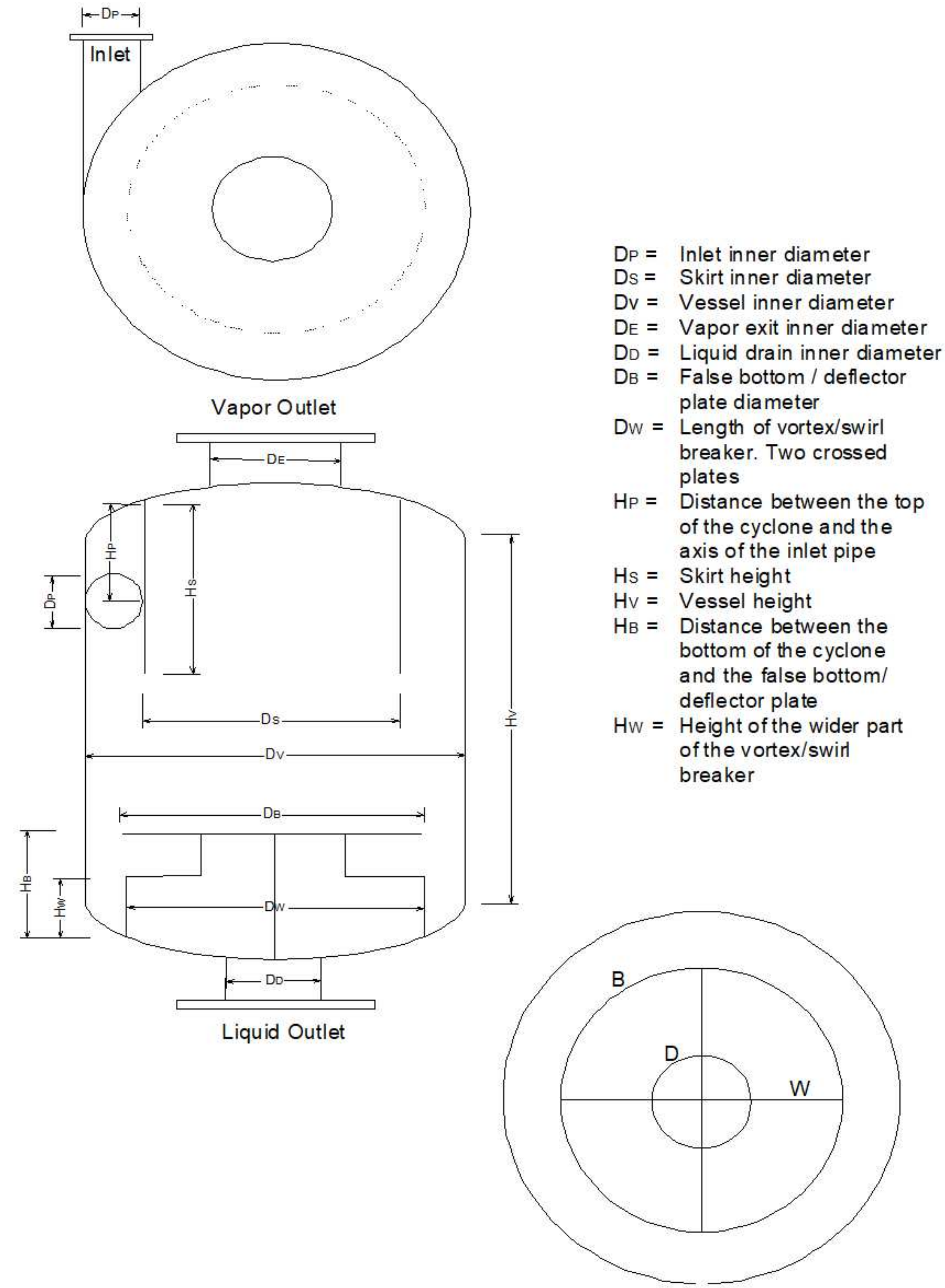
5. Size the liquid drain nozzle. For continuous draining of liquid, the nozzle should be sized for a maximum velocity, $u_d$ of about 1 to 3 m/s.

$$D_d = \sqrt{\frac{4W_l}{\pi \rho_l u_d}} \tag{9}$$

Vapor discharge during draining should be prevented through the use of a liquid seal leg or trap. For cyclones where the liquid is collected in the bottom, the size may be set by available time to drain or pump to liquid after the ovepressure event is over. If a pump is used, NPSH requirements should be considered is sizing the piping and in specifying the elevation of cyclone.

Figure 6: Cyclone separator with dimensions



### 5.15. BASF Cyclone Separator Design Method

Another method for the design of cyclone separators for emergency relief was published by Schmidt [7]. It is known as the BASF method.

$$
\begin{aligned}
D_v &= 0.42\frac{\sqrt{W_v}}{\rho_v^{1/4}} && (10)\\
H_v &= 2.00\ D_v && (11)\\
H_p &= 0.50\ D_v && (12)\\
H_s &= 0.75\ D_v && (13)\\
D_s &= 0.75\ D_v && (14)\\
D_e &= 0.50\ D_v && (15)\\
D_p &= 0.25\ D_v && (16)\\
D_B &= 0.75\ D_v && (17)\\
H_B &= 0.50\ D_v && (18)\\
D_w &= 1.00\ D_v && (19)\\
H_w &= 0.25\ D_v && (20)\\
& && (21)
\end{aligned}
$$

For typical values of vapor density and mass flow rates, the CCPS method yields a cyclone vessel diameter that is approximately 1.5 to 2 times larger than that calculated by the BASF method:

$$\frac{D_{v,CCPS}}{D_{v,BASF}} \simeq 1.0 + \frac{2D_p\rho_v^{1/4}}{0.42\sqrt{W_v}} \quad (22)$$

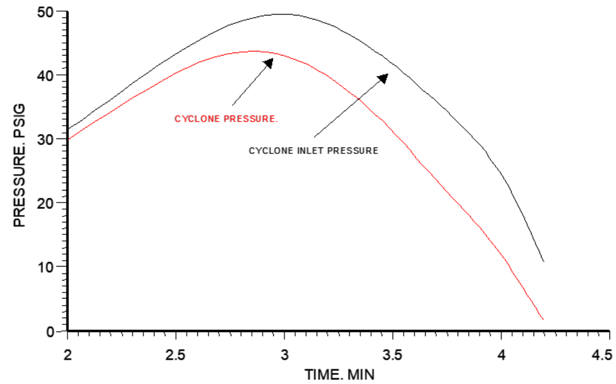
The BASF method recommends that the cyclone separator should be adequately anchored to support the high flow reaction force and bending (tilting) moment between the base and the cyclone inlet. They set the upper limit of the reaction force to twice the flow momentum in the inlet pipe:

$$
\begin{aligned}
\text{Force} &\leq 2\frac{(W_v+W_l)^2}{\rho_m A_p} \text{ where } \frac{1}{\rho_m} = \frac{x}{\rho_v} + \frac{1-x}{\rho_l} \text{ and } x = \frac{W_v}{W_v+W_l} && (23)\\
\text{Moment} &\leq \text{Force}\,(H_v - H_p) && (24)
\end{aligned}
$$

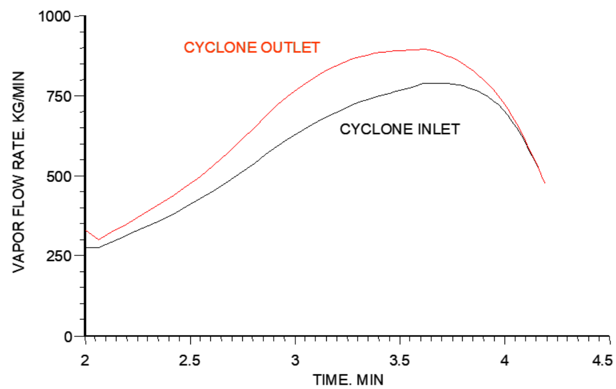
### 5.16. Cyclone Separator Design Example

This cyclone separator design example was performed using SuperChems Expert™ to demonstrator that the optimal cyclone dimensions depend on the transient nature of flow from a halogenation runaway reaction. During a runaway reaction, liquid and vapor flows can peak at different times during relief. Typically the initial two-phase flow is liquid rich at the beginning of relief and vapor rich towards the end of relief. Both reactants and products are toxic and flammable in this example. The reactor is protected by a rupture disk where the relief discharge line is tied into a common header which then connects to the cyclone separator. The liquid separated by the cyclone is routed to a separate collection vessel, while the cyclone vapor outlet is routed to a nearby common vent stack for safe dispersion.

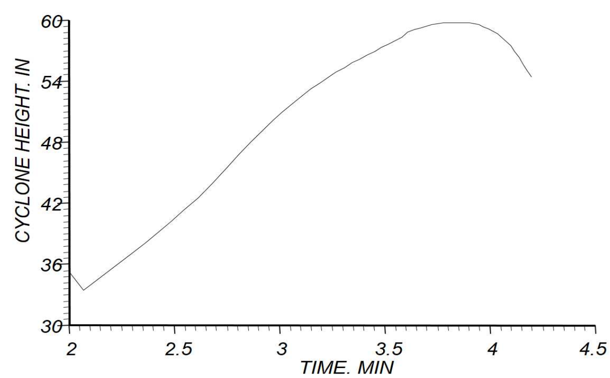
The dynamic cyclone pressure history calculated by SuperChems Expert shows that the cyclone pressure is lower than the inlet line pressure due to pressure drop in the cyclone




The simulation results also show that more flashing occurs in the cyclone and increases the vapor flow rate to the vent stack




Finally, the simulation results show that the optimal cyclone height is calculated at approximately four minutes during the relief transient

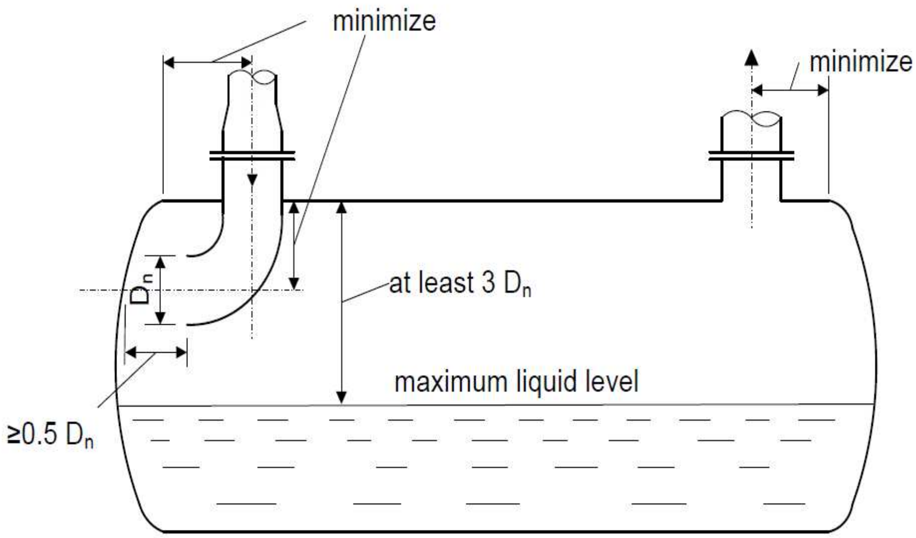



## 5.2 Gravity Separators

This type of effluent handling and vent containment system, shown in Figures 7 and 8 often combines both the vapor-liquid separation and containment (holdup) functions in one vessel. Horizontal separators are commonly used where space is plentiful, such as in petroleum refineries and petrochemical plants. The two-phase mixture usually enters at one end, and the vapor exits at the other end. For two-phase streams with very high vapor flow rates, inlets may be provided at each end, with the vapor outlet at the center of the drum, thus minimizing vapor velocities at the inlet and aiding vapor-liquid separation. API RP 521 contains a good discussion of this type of equipment.

Figure 7: Horizontal gravity separator and catch vessel



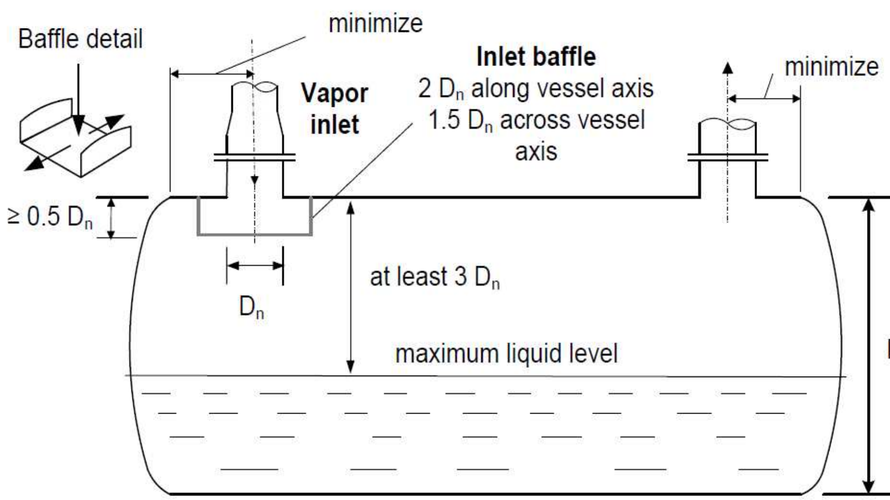
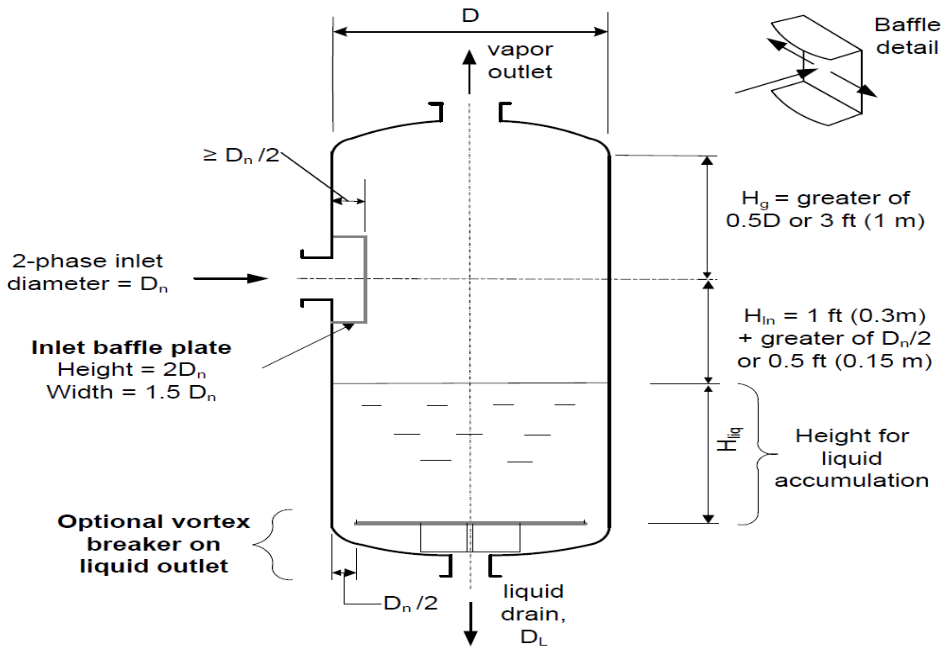



Figure 8: Vertical gravity separator and catch vessel



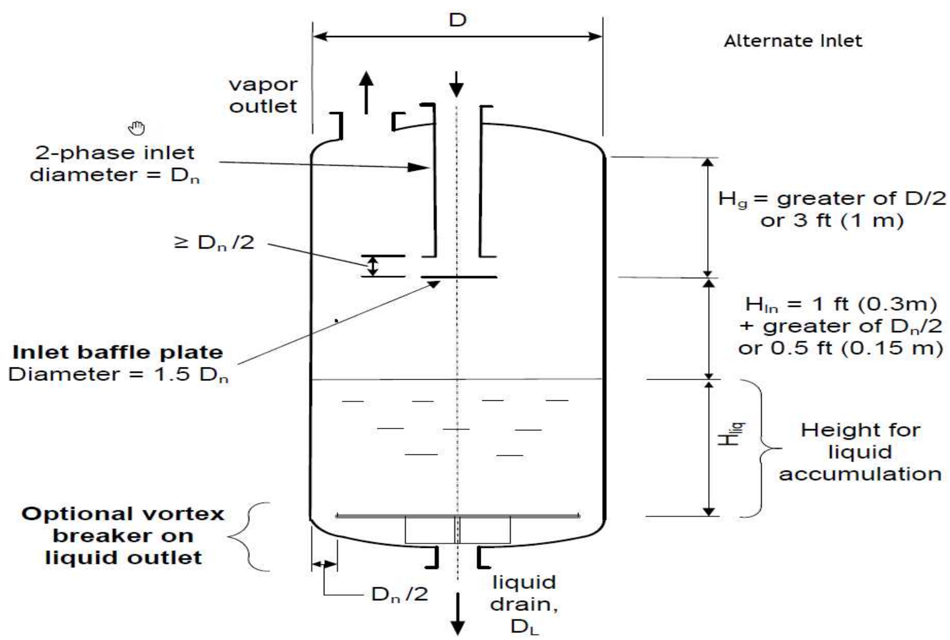
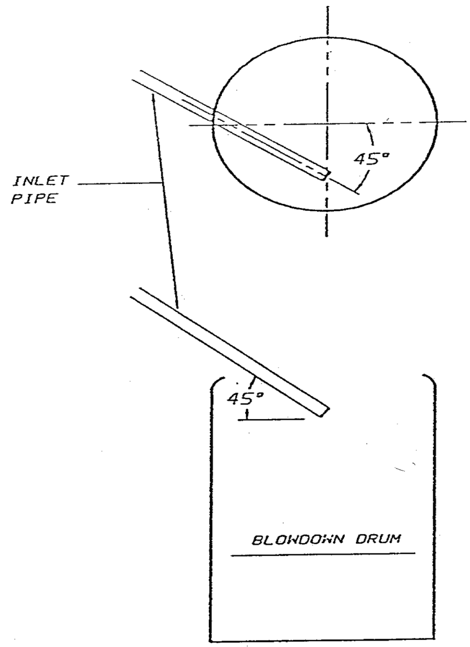



Figure 9: Open top vertical separator and catch vessel [1]




An integral open top separator and catch vessel variation is sometimes used when it is desirable to prevent the containment of highly explosive gases (e.g., hydrogen-air mixtures). Open pits have also been used as alternatives to open-top vessels (see Figure 9).

Two-phase separators can be oriented either vertically or horizontally. Selection of orientation will typically depend on cost, available space, and the vapor to liquid ratio of what needs to be separated [8].

Vertical separators are recommended for separation of mixtures where the vapor to liquid ratio is high. Horizontal separators are preferred for separation of mixtures where the vapor to liquid ratio is low.

Separation occurs in three stages:

1. Cancellation of momentum of entering two-phase mixture. This can be accomplished using a diverter where the liquid droplets will impinge upon such that small droplets will coalesce and form large droplets that will fall out by gravity.

2. Gravity separation. This accomplished by providing a large vapor disengagement space such that droplets will fall out due to their terminal settling velocity being smaller than the superficial velocity of the vapor leaving the separator.

3. Mist elimination. This is where very small droplets are collected and coalesced to form larger droplet that will drop due to gravity.

For item 2, the allowable superficial vapor velocity can be calculated such that it is smaller than the terminal settling velocity of the liquid droplets. The terminal settling velocity of a spherical liquid droplet can be estimated using a force balance between gravity and drag:

$$F_{drag} = \frac{\pi C_D D_d^2 U_v^2 \rho_v}{8} = F_{Gravity} = \frac{M_d(\rho_l - \rho_v)g}{\rho_v} \tag{25}$$

$C_D$ is given by the following expressions:

$$C_D = \frac{24}{N_{Re}} \quad , \mathrm{N_{Re}} < 0.1 \tag{26}$$

$$C_D = \frac{24}{N_{Re}}\left[1 + \frac{3}{16}N_{Re} + \frac{9}{160}N_{Re}^2 \ln(2N_{Re})\right] \quad , 0.1 < \mathrm{N_{Re}} \leq 2 \tag{27}$$

$$C_D = \frac{24}{N_{Re}}\left[1 + 0.15 N_{Re}^{0.687}\right] \quad , 2 \leq \mathrm{N_{Re}} < 500 \tag{28}$$

$$C_D = 0.44 \quad , 500 \leq \mathrm{N_{Re}} < 200,000 \tag{29}$$

$N_{Re}$ is the droplet Reynolds number and is based on the relative velocity, $u_t$, between the droplet and the surrounding medium:

$$N_{Re} = \frac{D_d |u_t| \rho_v}{\mu_v} \tag{30}$$

where $C_D$ is the drag coefficient, $D_d$ is the droplet diameter, $U_v$ is the vapor velocity, $M_d$ is the droplet mass, $g$ is the gravitation constant, and $\rho$ is mass density.

We can solve the above equation for the terminal settling velocity:

$$u_t = \sqrt{\frac{4gD_d(\rho_l - \rho_v)}{3C_D\rho_v}} = K\sqrt{\frac{\rho_l - \rho_v}{\rho_v}} \tag{31}$$

where $K$ is defined by:

$$K = \sqrt{\frac{4gD_d}{3C_D}} \text{ where K is in m/s} \tag{32}$$

One has to guess the value of $u_t$, estimate $N_{Re}$, solve for $C_d$, and then solve for $u_t$ until convergence is achieved.

A droplet of a diameter $D_d$ will drop out due to gravity if the vapor velocity $U_v$ is less than $U_t$, the droplet terminal settling velocity. This suggests that in order to design an effective separator we need to define the minimum droplet size/diameter that we need to separate. The estimation of stable droplet diameter is difficult and uncertain. Table 4 summarizes several recommended $K$ values to be used based on empirical data and best recommended practice: The value recommended by Grossel (K=0.27 ft/s or 0.0823 m/s) is good for separating droplets between 300 to 600 microns in diameter. Recently, Monnery and Svrcek [10] conducted experimental and analytical studies to determine the separation efficiency of flare knockout drums. They recommend that drums should be designed to remove an average droplet of 300 microns in order to avoid carryover. API-521, for example, does not include a check for possible entrainment from the liquid surface in the separator if the vapor velocity is too high.

Table 4: Separator K Values

**Mist Eliminator**

$$\begin{aligned} K &= 0.1821 + 0.0029P + 0.0460\ln(P) \;\; 1 \leq P \leq 15 \\ K &= 0.35 \;\; 15 \leq P \leq 40 \\ K &= 0.430 - 0.023\ln(P) \;\; 40 \leq P \leq 5,500 \end{aligned}$$

where $P$ is in psia and $K$ is in ft/s.

**GPSA** Gas Processor's Supplier Association engineering data book [9].

$$K = 0.35 - 0.01\frac{(P-100)}{100} \;\; 0 \leq P \leq 1,500$$

- Most vapors under vacuum, K = 0.20
- For glycol and amine solutions, multiply K by 0.6-0.8
- For vertical vessels without mist eliminators, divide K by 2
- For compressor suctions, scrubbers, mole sieve scrubbers and expander inlet separators multiply K by 0.7 - 0.8

where $P$ is in psia and $K$ is in ft/s.

**Grossel [6]** The recommended value is between 0.157 and 0.4 ft/s. A value of 0.27 is conservative for droplet sizes ranging from 300 to 600 microns.

Table 5: Important definitions and criteria used in separator design

**Holdup** Time it takes to reduce the liquid level from normal (NLL) to empty (LLL) while maintaining a normal outlet flow without feed makeup. This is based on the reserve required to maintain good control and safe operation of downstream facilities/systems.

**Surge Time** The time it takes for the liquid level to rise from normal (NLL) to maximum (HLL) while maintaining a normal feed without any outlet flow. This is based on the requirements for accumulation of liquid as a result of downstream/upstream variations and upsets.

Table 5 provides definitions of important terms to be used in the design procedures to follow. These terms are also shown in Figures 10 and 11.

### 5.21. Vertical Separator Design Procedure

The vapor disengagement area is the entire cross-sectional area of the vessel. The vessel diameter is calculated from a simple mass balance knowing the volumetric vapor flow rate (including flashing) and based on the allowable vapor velocity with respect to a selected value of droplet terminal settling velocity:

$$D_{vd} = \sqrt{\frac{4Q_v}{\pi u_v}} \tag{33}$$

where $Q_v$ is the volumetric flow rate in $(\mathrm{m}^3/\mathrm{s})$ and $u_v$ is the allowable vapor velocity, typicall set from 75 to 100 % of $u_t$.

The total vessel height can be calculated by adding the individual section heights as shown in Figure 10:

$$H_T = H_{LLL} + H_H + H_S + H_{LIN} + H_D + H_{ME} \tag{34}$$

The following is a step-by-step procedure for the design of a vertical separator:

1. Specify peak vapor flow rate, $Q_v$, in $\mathrm{m}^3/\mathrm{s}$
2. Specify peak liquid flow rate, $Q_l$, in $\mathrm{m}^3/\mathrm{s}$
3. Calculate the terminal settling velocity from Equation 31. See table 4 for typical values of $K$ or estimate $K$ using a specified droplet diameter / size.
4. Set $u_v$ to 0.75 $u_t$
5. Calculate vessel diameter using Equation 33. If there is a mist eliminator, add 3 to 6 inches to $D_{vd}$ for the support ring and round up to the next 6 inches.

6. Calculate the holdup volume:
$$V_H = t_H Q_l \tag{35}$$
where $t_H$ is obtained from Table 6

7. Calculate the surge volume:
$$V_S = t_S Q_l \tag{36}$$
where $t_S$ is obtained from Table 6.

8. Obtain low liquid level height, $H_{LLL}$, from Table 7.

9. Calculate the height from the low liquid level to the normal liquid level:
$$H_H = \frac{4V_H}{\pi D_{vd}^2} \tag{37}$$
If $H_H$ is less than 0.30 m, set $H_H$ to 0.30 m

10. Calculate the height from the low liquid level to high liquid level:
$$H_S = \frac{4V_S}{\pi D_{vd}^2} \tag{38}$$
If $H_S$ is less than 0.15 m, set $H_S$ to 0.15 m.

11. Calculate the height from high liquid level to the centerline of the inlet nozzle:
$$\begin{aligned}
H_{LIN} &= 0.30 + d_N \text{ with inlet diverter} \\
H_{LIN} &= 0.30 + \frac{d_N}{2} \text{ without inlet diverter} \\
d_N &\geq \frac{0.2337}{\sqrt{\pi}} \left(\sqrt{Q_v + Q_l}\right) (\rho_m)^{1/4} \\
\rho_m &= \rho_l \alpha + (1-\alpha)\rho_v \\
\alpha &= \frac{Q_l}{Q_l + Q_v}
\end{aligned}$$

12. Calculate the disengagement height, from the centerline of the inlet nozzle to:
    (a) the vessel top tangent line if there is no mist eliminator, or
$$H_D = 0.5 D_{vd} \tag{39}$$
    (b) the bottom of the demister pad, minimum of
$$H_D = 0.92 + \frac{d_N}{2} \text{ without mist eliminator or} \tag{40}$$
$$H_D = 0.60 + \frac{d_N}{2} \text{ with mist eliminator} \tag{41}$$

13. If there is a mist eliminator, take 0.15 m for the mist eliminator pad and take 0.30 m from the top of the mist eliminator to the top of the tangent line of the vessel.

$$H_{ME} = 0.15 + 0.30 = 0.45 \text{ m} \tag{42}$$

If there is not mist eliminator

$$H_{ME} = 0 \tag{43}$$

14. Calculate the total height of the vessel, $H_T$:

$$H_T = H_{LLL} + H_H + H_S + H_{LIN} + H_D + H_{ME} \tag{44}$$

## 5.22. Horizontal Separator Design Procedure

The procedure for sizing horizontal separators is similar to that used for vertical separators. The following should be noted, however:

- The cross-sectional area of a horizontal separator is occupied by both vapor and liquid
- The liquid droplets to be separated have a horizontal drag force that is not directly opposite to gravity as is in the vertical case. The allowable horizontal velocity can be higher than the terminal settling velocity. The time it takes to travel a horizontal length between the inlet and the outlet nozzles must be greater than the time it takes to settle vertical distance to the liquid surface:

$$\frac{L}{u_{AH}} \leq \frac{H_V}{u_t} \tag{45}$$

- The design requires iterative calculations.

The following volume balance equation will drive the sizing for a horizontal separator:

$$V_H + V_S = L(A_T - A_{VD} - A_{LLL}) \tag{46}$$

where $A_{VD}$ is the vapor/liquid disengagement area. $A_{VD}$ is typically specified as 0.3 or 0.7 m or 20 % of the vessel inside diameter, whichever is greater.

The following is a step-by-step design procedure:

1. Specify peak vapor flow rate, $Q_v$, in $\mathrm{m}^3/\mathrm{s}$
2. Specify peak liquid flow rate, $Q_l$, in $\mathrm{m}^3/\mathrm{s}$
3. Calculate the terminal settling velocity from Equation 31. See table 4 for typical values of $K$ or estimate $K$ using a specified droplet diameter / size.
4. Set $u_v$ to 0.75 $u_t$

Table 6: Liquid holdup and surge times

| Service | Holdup Time (s) | Surge Time (s) |
|---|---|---|
| Unit Feed Drum | 600 | 300 |
| Separators | | |
| Feed to column | 300 | 180 |
| Feed to other drum or tankage with pump or through exchanger | 300 | 120 |
| Feed to other drum or tankage without pump | 120 | 60 |
| Feed to fired heater | 600 | 180 |
| Reflux or product accumulator | | |
| Reflux only | 180 | 120 |
| Reflux and product | 180+ | 120+ |
| Based on reflux (180 s) + appropriate holdup time of ovhd products | | |
| Column bottoms | | |
| Feed to another column | 300 | 120 |
| Feed to other drum or tankage with pump or through exchanger | 300 | 120 |
| Feed to other drum or tankage without pump | 120 | 60 |
| Feed to fired boiler | 300-480 | 120-240 |
| Based on reboiler vapor expressed as liquid (180s) + appropriate holdup time for the bottom product | | |

- For compressor suction / interstage scrubber
  - 180 s between HLL (HLA) and HLSD
  - 600 s from bottom tangent line to HLA
- Fuel gas knockout drum
  - 20 ft slug in the incoming fuel line between NLL and HLSD
- Flare knockout drum
  - 1,200 s to 1,800 s to HLL
- Personnel factor. 1 for experienced, 1.2 for trained, and 1.5 for inexperienced
- Instrumentation factor. 1 for well instrumented, 1.2 for standard instrumentation, and 1.5 for poorly instrumented

Table 7: Low liquid level height

| Vessel Diameter (m) | $<20$ bars Vertical LLL (m) | $>20$ bars Vertical LLL (m) | Horizontal LLL (m) |
|---|---|---|---|
| $\leq$1.22 | 0.38 | 0.15 | 0.23 |
| 1.83 | 0.38 | 0.15 | 0.25 |
| 2.44 | 0.38 | 0.15 | 0.28 |
| 3.05 | 0.15 | 0.15 | 0.31 |
| 3.66 | 0.15 | 0.15 | 0.33 |
| 4.88 | 0.15 | 0.15 | 0.38 |

5. Calculate the holdup volume:
$$V_H = t_H Q_l \tag{47}$$
where $t_H$ is obtained from Table 6

6. Calculate the surge volume:
$$V_S = t_S Q_l \tag{48}$$
where $t_S$ is obtained from Table 6.

7. Obtain an estimate for L/D from Table 8.

8. Calculate initial vessel diameter from the following equation:
$$D = \left(\frac{4(V_H + V_S)}{\pi(0.6)(L/D)}\right)^{1/3} \tag{49}$$
Round to the nearest 0.15 m.

9. Calculate the total cross-sectional area:
$$A_T = \frac{\pi D^2}{4} \tag{50}$$

10. Calculate the low liquid level height $H_{LLL}$ using Table 7 or
$$H_{LLL} = \frac{D}{20} + 0.178 \tag{51}$$
If $D \leq 1.22$ m, set $H_{LLL}$ to 0.22 m.

11. Calculate $A_{LLL}/A_T$ from the following equations:

$$\begin{aligned}\frac{A_{LLL}}{A_T} = & \left(0.00004755930 + \frac{0.174875H_{LLL}}{D} + \frac{5.668973H_{LLL}^2}{D^2} - \frac{4.916411H_{LLL}^3}{D^3} - \frac{0.145348H_{LLL}^4}{D^4}\right) / \\ & \left(1 + \frac{3.924091H_{LLL}}{D} - \frac{6.358805H_{LLL}^2}{D^2} + \frac{4.018448H_{LLL}^3}{D^3} - \frac{1.801705H_{LLL}^4}{D^4}\right)\end{aligned}$$

12. Calculate $A_{LLL}$:
$$A_{LLL} = \left(\frac{A_{LLL}}{A_T}\right) A_T \tag{52}$$

13. Calculate the minimum height required for the vapor disengagement area.

$$\begin{aligned} H_V &= \max(0.2D, 0.30) \text{ without a mist eliminator} \\ H_V &= \max(0.2D, 0.70) \text{ with mist eliminator} \end{aligned}$$

14. Calculate $A_V/A_T$ using $H_V/D$ from the same equation used for $A_{LLL}/A_T$.

15. Calculate $A_V$
$$A_V = \left(\frac{A_V}{A_T}\right) A_T \tag{53}$$

16. Calculate minimum length to accommodate the liquid surge/holdup:
$$L = \frac{V_H + V_S}{A_T - A_V - A_{LLL}} \tag{54}$$

17. Calculate the liquid dropout time
$$\phi = \frac{H_V}{u_V} \tag{55}$$

18. Calculate the actual vapor velocity, $u_{VA}$:
$$u_{VA} = \frac{Q_V}{A_V} \tag{56}$$

19. Calculate the minimum length required for vapor/liquid disengagement:
$$L_{MIN} = u_{VA}\phi \tag{57}$$

20. Depending of the values of $L$ and $L_{MIN}$ the following steps are required:
    (a) $L < L_{MIN}$. Set $L$ equal to $L_{MIN}$. This will simply result in some extra holdup.
    (b) $L << L_{MIN}$. Increase $H_V$ and repeat from step 13.
    (c) $L > L_{MIN}$. The design is acceptable for vapor/liquid separation.
    (d) $L >> L_{MIN}$. Liquid holdup is controlling. To increase $L_{MIN}$ and decrease $L$ we need to decrease $H_V$. $H_V$ can be decreased if it is larger than the minimum specified in step 13. Repeat calculation from step 13.

21. Calculate $L/D$.
    (a) if $L/D > 6$, increase $D$ and repeat the calculation from step 8
    (b) if $L/D < 1.5$, decrease $D$ and repeat the calculation from step 8

22. Select the head type for the vessel:
    (a) If $D < 4.57m$ and $P > 100$ psig, use 2 to 1 elliptical heads
    (b) If $D > 4.57m$, use hemispherical heads regardless of P value
    (c) If $D < 4.57m$ and $P < 100$ psig, use dished heads with knuckle radius of 0.6 D.
23. Calculate the thickness and surface area of the shell and heads from the following equations:

    **Shell**

$$\begin{aligned} t_{shell} &= \frac{PD}{2SE - 1.2P} + t_c \\ A_{shell} &= \pi DL \end{aligned}$$

    **2 to 1 Elliptical Heads**

$$\begin{aligned} t_{head} &= \frac{PD}{2SE - 0.2P} + t_c \\ A_{head} &= 1.09D^2 \end{aligned}$$

    **Hemispherical Heads**

$$\begin{aligned} t_{head} &= \frac{PD}{4SE - 0.4P} + t_c \\ A_{head} &= 1.571D^2 \end{aligned}$$

    **Dished Heads**

$$\begin{aligned} t_{head} &= \frac{0.885PD}{SE - 0.1P} + t_c \\ A_{head} &= 0.842D^2 \end{aligned}$$

    where $t_c$ is the corrosion allowance (in), $E$ is the joint efficiency ranging from 0.6 to 1 (0.85 for spot examined joints, 1 for 100 % x-ray joints), $S$ is the allowable stress in $psi$, $D$ is the diameter in (in), $P$ is the design pressure in (psig) typically set at operating pressure plus the greater of 15-30 psi or 10 to 15 % P. The temperature at which the metal properties are computed should be at the operating temperature plus 25 to 50 F if the operating temperature is more than 200 F. If the operating temperature is less than 200 F, use 250 F. If the overpressure is caused by boiling, use the boiling temperature.
24. Calculate the vessel weight:

$$M_{metal} = 7,800\,(0.02548 \max(t_{shell}, t_{head}))\,(A_{Shell} + 2A_{head})\,(0.0254^2) \tag{58}$$

25. Increase and decrease the diameter $D$ by 0.15 m and repeat the calculations until $L/D$ ranges from 1.5 to 6 until minimum vessel weight is obtained.

26. Calculate the normal liquid level $H_{NLL}$:

$$
\begin{aligned}
A_{NLL} &= A_{LLL} + \frac{V_H}{L} \\
H_{NLL} &= D\left(0.00153756 + \frac{3.299201 A_{NLL}}{A_T} + \frac{24.353518 A_{NLL}^2}{{A_T}^2} - \frac{36.999376 A_{NLL}^3}{{A_T}^3} + \frac{9.892851 A_{NLL}^4}{{A_T}^4}\right) \\
&/ \left(1 + \frac{26.787101 A_{NLL}}{A_T} - \frac{22.923932 A_{NLL}^2}{{A_T}^2} - \frac{14.844824 A_{NLL}^3}{{A_T}^3} + \frac{10.529572 A_{NLL}^4}{{A_T}^4}\right)
\end{aligned}
$$

27. Calculate the high liquid level:

$$H_{HLL} = D - H_V \tag{59}$$

Table 8: L/D ratio guidelines for horizontal separators

| Vessel Operating Pressure (psig) | L/D |
|---|---|
| 0 < P <= 250 | 1.5 to 3.0 |
| 250 < P <= 500 | 3.0 to 4.0 |
| 500 < P | 4.0 to 6.0 |

## 5.23. Separator Safety Considerations

A pressure relief device may be required for the separator if one or more of the following conditions are satisfied:

1. External fire exposure, vapor outlet should be able to handle vapor boiloff rate
2. Blocked vessel outlet
3. Continuing chemical reaction

Freeze protection may also be required if high viscosity fluids are being separated.

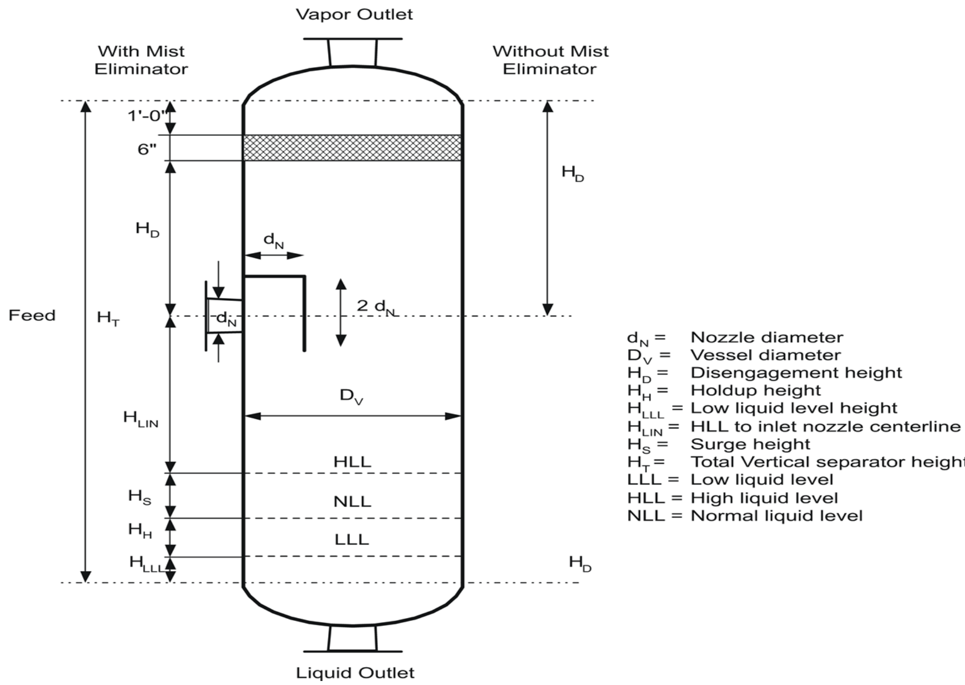
## 5.24. Separator Design and Instrumentation

For separators that operate at ambient pressure a design pressure of 50 to 75 psig is recommended. This pressure rating will adequate to protect against vessel rupture from an internal deflagration. For a higher operating pressure level, the design basis should be adjusted according to ASME code or NFPA 68.

The following instrumentation and equipment are recommended:

1. Low and high liquid level alarms
2. Temperature and pressure indication

Figure 10: Vertical separator design




3. Pressure relief device

4. Manhole for maintenance and cleaning

5. Pump for transferring accumulated liquid

6. Anti-vortex baffle above the liquid outlet line to the pump

7. Bracing or wall stiffeners to allow for jet reaction force loading on the vessel wall, and for intermittent vibration if two-phase slug flow occurs.

## 5.3 Integral Vortex Separator and Catch Vessel [1]

As can be seen from Figure 12, one obvious advantage of this type of separator is that it can be mounted directly above a reactor so that the vent line between the reactor and separator is vertical with no elbows. The guide blades (vanes) create a vortex motion so that the liquid is deposited on the walls of the conical section and flows out through the annular area between the vapor line and the conical section. Because the liquid is separated from the vapor-liquid mixture vented during a runaway, the relief device can be smaller in size than what would be required if it was mounted on the reactor, as it can now be sized for essentially all vapor flow.

Figure 11: Horizontal separator design

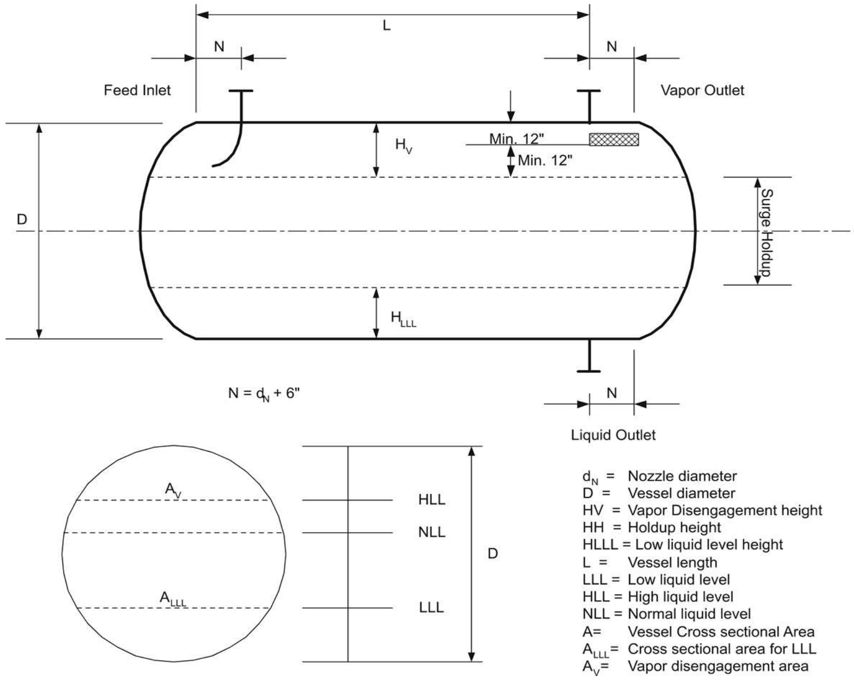
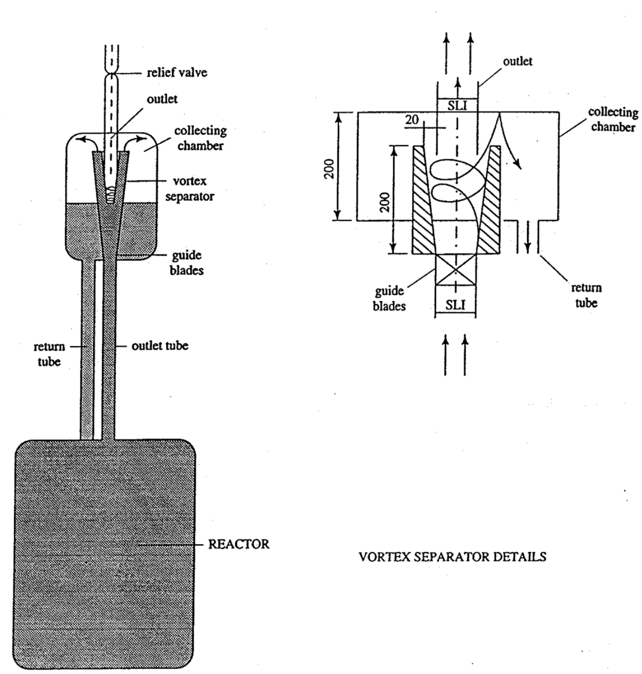



Figure 12: Vortex separator and catch vessel



# 6. Quenching

This type of system, as shown in Figures 13 and 14, is used when it is desired to remove condensible vapors from a flammable or toxic vent mixture by passing them through a pool of liquid in the quench vessel. This arrangement often eliminates the need for a subsequent scrubber and/or flare stack. The design of the sparger is critical to efficient condensation and avoidance of water hammer.

Figure 13 is the more conventional ”passive” type quench vessel used in the chemical and nuclear industry. The quench vessel shown in 14, with a superimposed baffle plate section, is used when complete condensation of vapors is not expected. This type is often used in petroleum refineries.

A quench vessel traps virtually the entire discharge of a relief system. The relief system discharge, liquid and vapor, is sparged into a cold pool of liquid in the quench vessel. As a result the vapor phase is condensed and the reaction is quenched due to dilution and temperature reduction. A quench vessel will be typically two to four times the size of the reactor volume. The size of a quench vessel is influenced by the total heat of reaction, the volume of liquid carryover, the operating pressure, and the initial quench liquid temperature and heat capacity.

Quench vessels can offer a high degree of containment and can handle low to very high flow rates, foamy or moderately viscous liquids, and condensible vapors. Their efficacy degrades if the effluent contains significant amount of non-condensible vapors.

## 6.1 How Quench Vessels Work

The relief system effluent is sparged into the quench vessel liquid pool using a sparger as shown in Figures 13 and 14.

The sparger divides the relief stream into multiple high velocity jets. The resulting high velocity jets provide violent contact and mixing with the quench liquid. Nearly instant heat transfer condenses all relief vapors. The momentum of jets mixes the liquid pool and minimizes hot spots. The liquid pool thermal capacity (heat sink) absorbs the heat of reaction and keeps the liquid pool below the boiling point.

A quench vessel should be designed to withstand a deflagration in the vapor space unless the vapor space is inerted. A quench vessel can operate at atmospheric pressure as an open vessel with relatively small emissions due to vapor space displacement and produces low backpressure on the relief system. Quench vessels can also be operated as closed vessels. A closed quench vessel requires a higher design pressure rating and generates a higher backpressure on the relief systems. However, a closed quench vessel has zero emissions. A quench vessel can also be partially vented via a rupture disk or pressure relief valve. This enables the use of lower vessel design pressures where the quench vessel relief device(s) control the pressure. The vapor flow from the partially vented quench vessel is usually routed to a vapor stack, flare, or scrubber.

The quench vessel should be protected from overpressure and if it is partially vented, it should discharge to a safe location. Overpressure scenarios to consider for the quench vessel when developing its relief requirements include maximum displacement rate of air and vapor, fire exposure,

Figure 13: Quencher knock-out drum/catch vessel

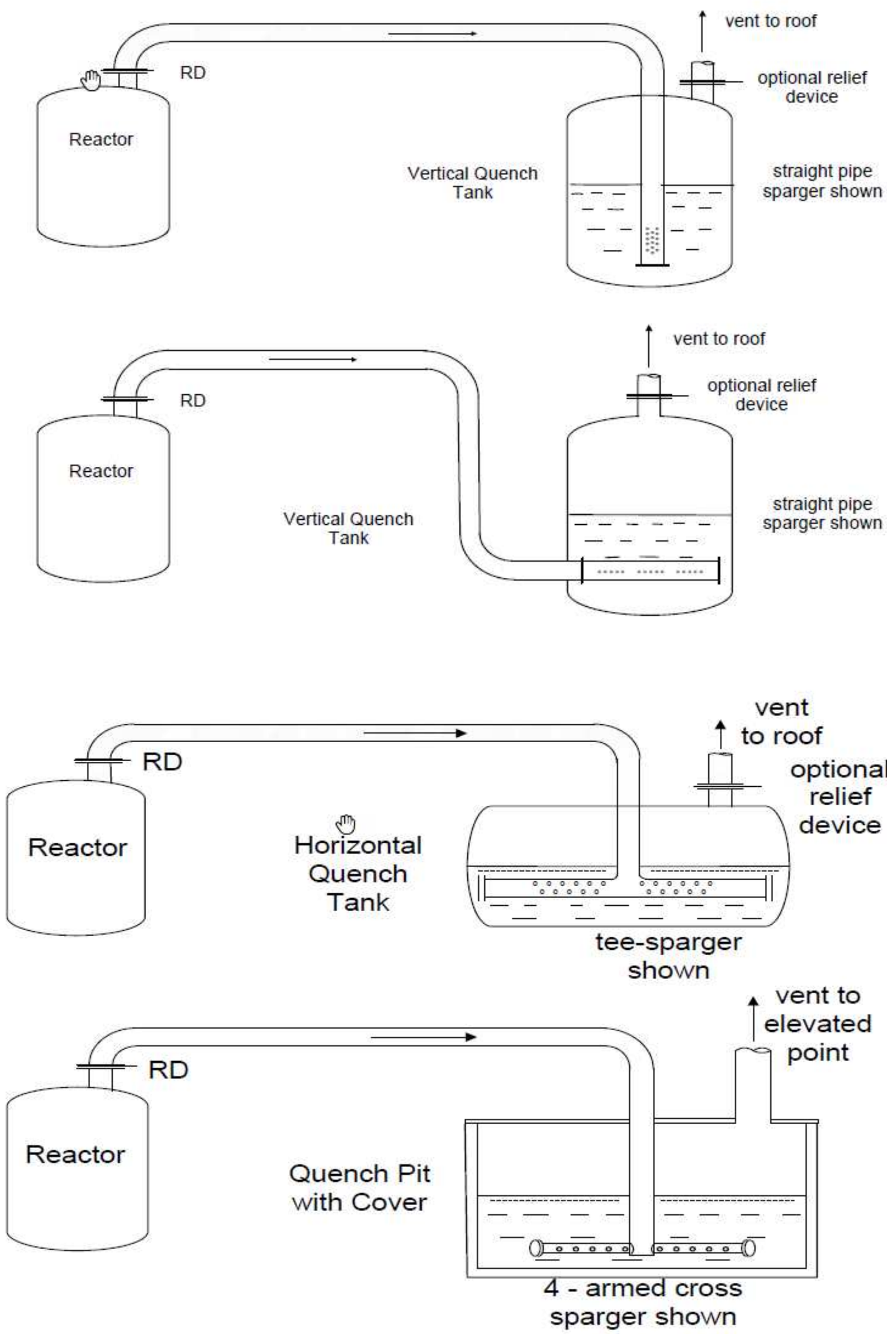
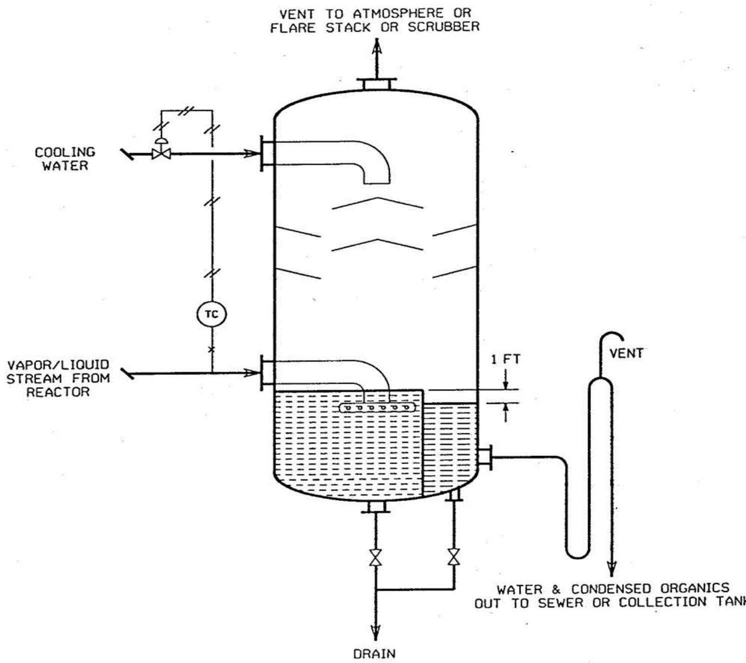


Figure 14: Quench vessel with direct contact baffle tray section




continuing reaction, omission of the quench liquid, and a vapor space deflagration.

## 6.2 Required Quantity of Quench Liquid

Selection of the quench liquid is important. The ideal quench liquid is one that is chemically compatible with the relief system discharged material, has a low vapor pressure, has a high specific heat capacity, has a low viscosity, has a low freezing point, has high thermal conductivity, and a high flash point. The quench liquid can either be miscible or immiscible with the relief discharged materials. The required quantity of a quench liquid is obtained via a simple overall heat balance.

$$m_Q \simeq \frac{m_R r_C \Delta H_R + m_R C_{p,R} (T_R - T_F)}{C_{p,Q} (T_F - T_Q)} \tag{60}$$

where $m_Q$ is the required mass of quench liquid, $m_R$ is the total mass in the reactor, $C_{p,R}$ is the specific heat capacity of the reactor contents, $C_{p,Q}$ is the specific heat capacity of the quench liquid, $T_R$ is the initial temperature of the reactor contents, $T_Q$ is the initial temperature of the quench liquid, $T_F$ is the final temperature of the mixed quench liquid and reactor contents, $\Delta H_R$ is the heat of reaction of the reactor contents, and $r_C$ is fractional degree of reaction.

When using this simple overall heat balance for an immiscible quench liquid with the reactor contents we assume that heat losses are negligible, the reactor empties all of its contents, and that all vapor condenses in the quench vessel. The initial quench liquid temperature $T_Q$ is selected as the highest 24 hour average ambient temperature and is locale dependent. The final temperature $T_F$ is selected to be 10 Kelvin below the boiling point. The fraction of reaction heat released $r_C$ is set to 1.0 or the actual conversion fraction required to reach the relief pressure in the quench vessel for a partially vented quench vessel.

For a miscible quench liquid, the final quench liquid temperature $T_F$ is selected 10 Kelvin below the lower of the boiling points of the relief discharge alone or the quench liquid alone.

The quench vessel volume must be large enough to accommodate the condensed vapor and liquid relief discharge as well as the quench liquid. The vapor space is typically set at 15 % of the total volume for a partially vented quench vessel, 30 to 50 % for an open quench vessel or pit, and as needed to limit quench vessel pressure for a closed quench vessel but no less than 10 %.

The dynamics of quench vessel with continuing chemical reaction(s) can be readily modeled using SuperChems Expert . The quench vessel is initially specified and partially filled with all the quench liquid. An initial estimate of the required quench liquid mass is provided from Equation 60. The transient time dependent relief of the reactor is then connected to the bottom of the quench vessel. The coupled reactor and quench vessels dynamics are simulated to yield time dependent flow, backpressure, temperature, pressure, composition, etc. for both vessels. SuperChems Expert also enables the quench vessel to be outfitted with a relief system and partially vented. More complex simulations where several reactors can be connected to the same quench vessel can also be performed. SuperChems Expert simulations assume that the incoming reactors flows are well mixed with the quench vessel liquid using an effective sparger design.

## 6.3 Sparger and Quench Vessel Design Considerations

An important aspect of quench vessels is the design and performance of the sparger. The primary function of the sparger is to divide the incoming relief stream into multiple high momentum small jets in good contact with the quench liquid and to have those jets well distributed in the quench liquid. The total flow area of all the sparger holes should be large enough such that the performance of the relief system is not negatively impacted due to backpressure. The holes should be sized and distributed such that the momentum forces are balanced, shocks caused by vapor bubble collapse are minimized and prevented, and momentum induced quench liquid recirculation is maximized.

Large momentum and shock forces can act on the sparger and quench vessel walls during relief. The sparger should be securely anchored to the vessel. The vessel should be securely anchored to ground or other heavy structures. The relief lines should be braced, especially elbows where the flow changes direction. The quench vessel walls must have sufficient stiffness to tolerate the reaction forces.

Smaller sparger holes, 6 to 12 mm (1/4 to 1/2 inch) are typically better. If fouling is possible, larger holes, 25 to 50 mm (1 to 2 inch) can be used. If the flow is choked at the end of the relief line, the total holes area should be equal to the relief line flow area divided by 0.6 or 1.67 times the line flow area. If the flow is subsonic (unchoked), allow for a pressure drop of 0.3 bar ($\geq 5$ psi and use a hole discharge coefficient of 0.6 with a frozen flow assumption, i.e. constant vapor fraction or quality.

The sparger arms should be arranged symmetrically in order to balance the momentum forces. The total sparger arms flow area should be equal to two times the total holes flow area. Many sparger types have been used including vertical dip-pipe, horizontal tee, horizontal radial arm, and horizontal ladder.

Even with high viscosity flow through the sparger, plugging may be minimized using large holes, 50 mm (2 inch). Note that polymer skins are usually soft at reaction temperature but may not be once cooled and plugging may be more likely. As a result, the sparger may be difficult to clean. When in doubt, small scale quench tests can provide more insight to the performance of the sparger.

A quench vessel that is properly vented or operating at atmospheric pressure can be used to handle multiple relief streams. On the other hand, non-vented quench vessels may not be able to because of possible pressure interaction between the incoming streams.

In some situations reverse flow can occur from the quench vessel back to the reactor. This can lead to reactor thermal shock, partial reactor vacuum, water hammer in reactor if it fills and a gas pad is not present, autoignition if air is drawn into the reactor, or vapor ignition of air is drawn into the quench vessel. These potential hazards can be mitigated by using vacuum breaker or designing the reactor for full vacuum, inerting the reactor and/or inerting the quench vessel. More guidance on the design of quench vessels can be found in reference [1].

### 6.4 Quench Vessel Design Example For $PCl3$-Water

This example deals with a 5,000 gal reactor in which phosphorus trichloride ($PCl_3$) is used. A scenario was identified where it is possible for a heel of phosphorus trichloride (2,700 kgs) to remain in the reactor undetected (below detection level) at 40 C, and for an operator to attempt to flush the vessel with water. This can lead to the generation of gaseous hydrogen chloride and excessive system pressure.

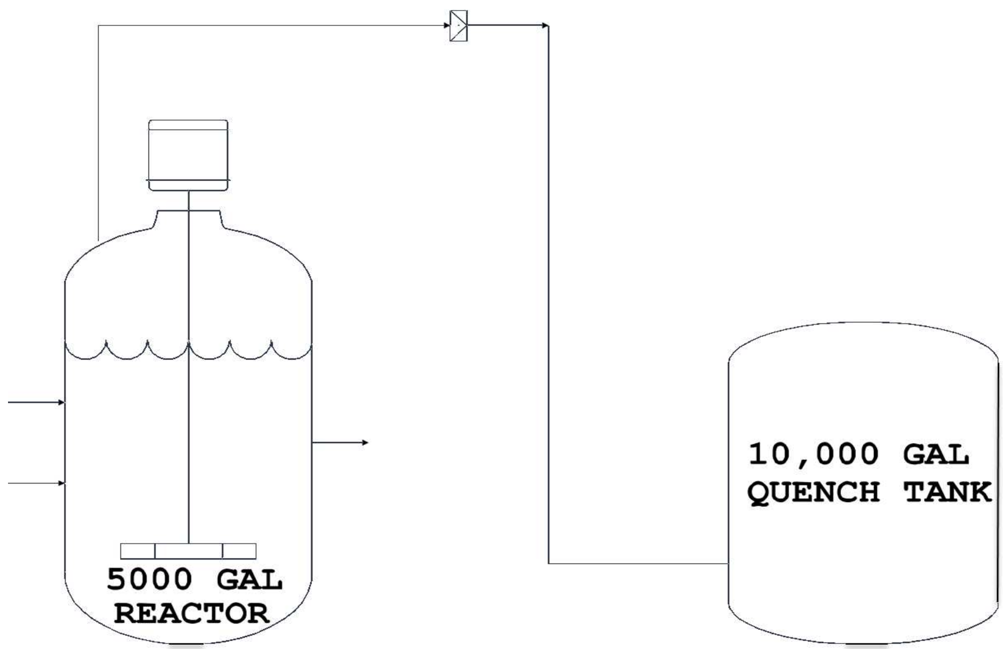
Water can be introduced into the reactor at the rate of 15 kg/s for 38 seconds. The reactor has a 12-inch rupture disk set at 20 psig. The effluent is discharged into a quench vessel. The quench vessel has a volume of 10,000 gal and initially contains 24,000 kgs of water at ambient conditions. The process equipment is illustrated in Figure 15. The reaction rate and characteristics of $PCl_3 - H_2O$ reaction are described by Melhem and Reid in reference [11].

Figure 16 illustrates the calculated time history of pressure, temperature, and individual component flow rates for the reactor and the quench vessel. Please note that both $HCl$ and $PCl_3$ are discharged from the reactor and that $PCl_3$ reacts with the quench vessel water to form phosphorous acid and $HCl$. One should also note that the temperature and pressure rise in the quench vessel are caused primarily by the hydrogen chloride heat of solution.

## 7. Conclusions

The effluent handling and vent containment equipment design procedures (models) are available in SuperChems Expert . These detailed models can be used as stand alone with external user specified

Figure 15: A Simple reactor/quench vessel arrangement



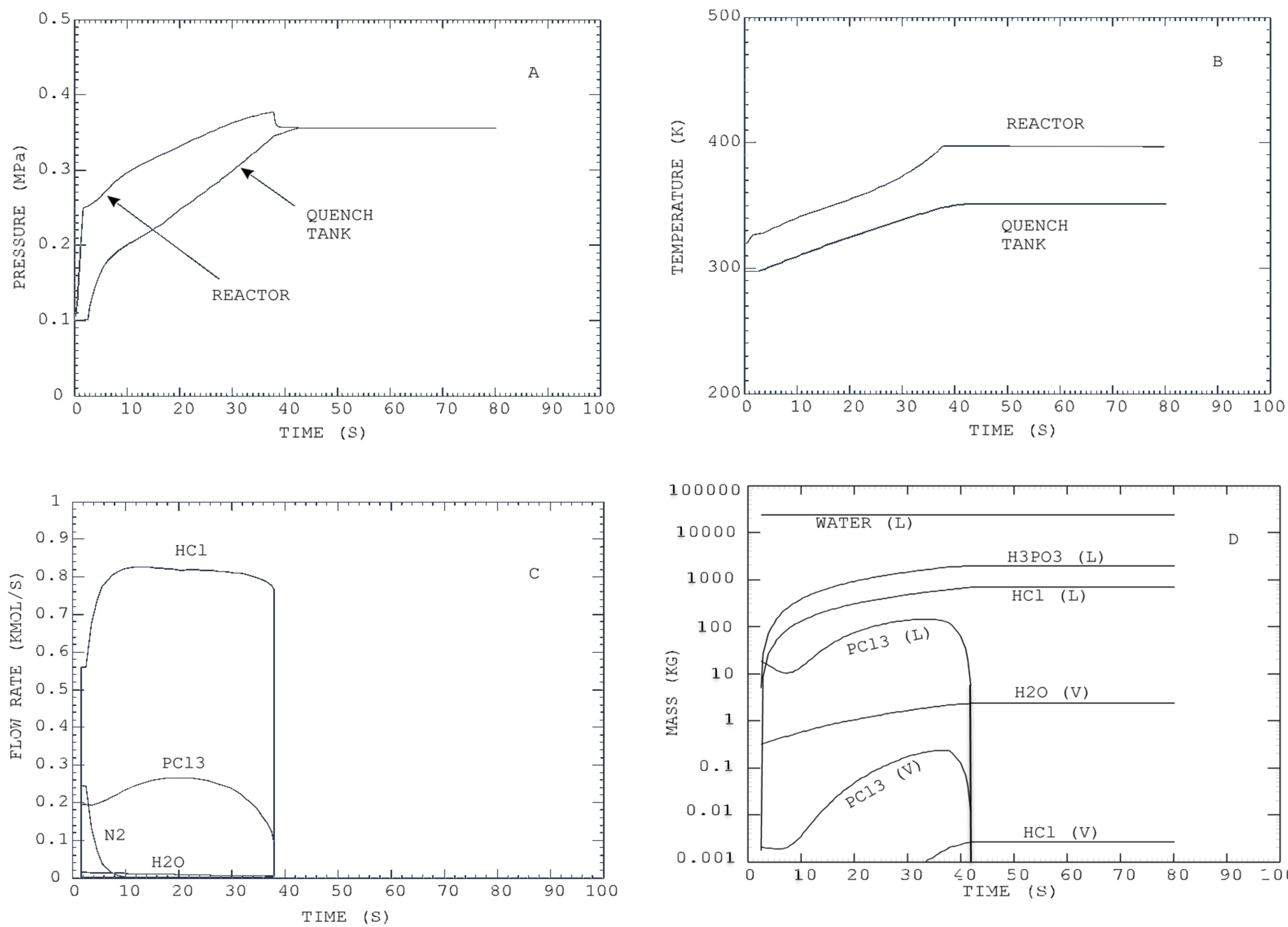

steady state flow streams. They can also be coupled with SuperChems Expert complex vessel multiphase relief dynamics (see examples) to provide optimal equipment design and risk reduction. SuperChems Expert also integrates detailed dispersion, explosion, and thermal radiation models with complex vessel relief dynamics to ensure optimal design, selection, and documentation of safe discharge location for single and mutiphase flow.

No matter what vent containment system is used one should always plan for the clean up and waste [3] disposal in order to minimize plant down time. There will almost always be a mess. In addition, vent containment system internals must be removed for cleaning.

[3]In some instances, the quench fluid maybe expensive and can be distilled and reused.

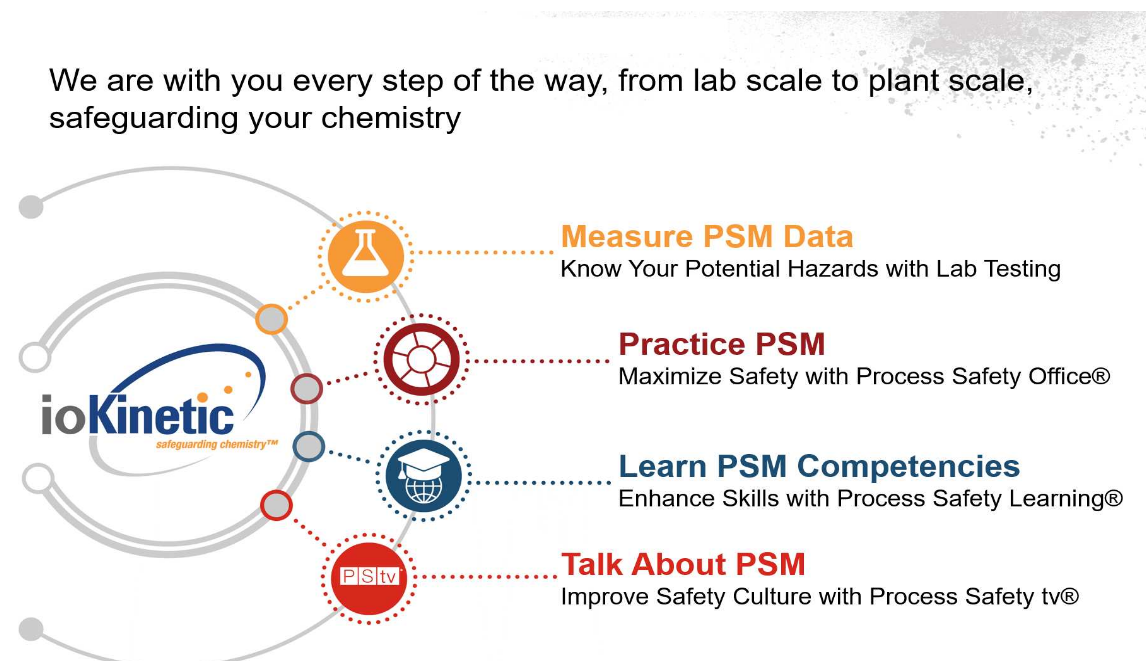
Figure 16: Estimated transient profiles for pressure, temperature, mass and flow rate using SuperChems Expert for the $PCl_3$-Water quench vessel design example



# 8. How can we help?

In addition to our deep experience in chemical reaction systems and the conduct of large-scale site wide relief systems evaluations by both static and dynamic methods, we understand the many non-technical and subtle aspects of compliance and legal requirements. When you work with ioMosaic you have a trusted partner that you can rely on for assistance and support with the lifecycle costs of relief systems to achieve optimal risk reduction and compliance that you can evergreen. We invite you to connect the dots with ioMosaic and ioKinetic.

Figure 17: Connect the dots with ioMosaic® and ioKinetic™

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## About the Author

Dr. Melhem is an internationally known pressure relief and flare systems, chemical reaction systems, process safety, and risk analysis expert. In this regard he has provided consulting, design services, expert testimony, incident investigation, and incident reconstruction for a large number of clients. Since 1988, he has conducted and participated in numerous studies focused on the risks associated with process industries fixed facilities, facility siting, business interruption, and transportation.

Prior to founding ioMosaic Corporation, Dr. Melhem was president of Pyxsys Corporation; a technology subsidiary of Arthur D. Little Inc. Prior to Pyxsys and during his twelve years tenure at Arthur D. Little, Dr. Melhem was a vice president of Arthur D. Little and managing director of its Global Safety and Risk Management Practice and Process Safety and Reaction Engineering Laboratories.

Dr. Melhem holds a Ph.D. and an M.S. in Chemical Engineering, as well as a B.S. in Chemical Engineering with a minor in Industrial Engineering, all from Northeastern University. In addition, he has completed executive training in the areas of Finance and Strategic Sales Management at the Harvard Business School. Dr. Melhem is a Fellow of the American Institute of Chemical Engineers (AIChE) and Vice Chair of the AIChE Design Institute for Emergency Relief Systems (DiERS).

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## About ioMosaic Corporation

Through innovation and dedication to continual improvement, ioMosaic has become a leading provider of integrated process safety and risk management solutions. ioMosaic has expertise in a wide variety of areas, including pressure relief systems design, process safety management, expert litigation support, laboratory services, training and software development.

As a certified ISO 9001:2015 Quality Management System (QMS) company, ioMosaic offers integrated process safety and risk management services to help you manage and reduce episodic risk. Because when safety, efficiency, and compliance are improved, you can sleep better at night. Our extensive expertise allows us the flexibility, resources, and capabilities to determine what you need to reduce and manage episodic risk, maintain compliance, and prevent injuries and catastrophic incidents.

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