The Anatomy of Liquid Displacement and Vapor Breakthrough

Georges A. Melhem, PhD, FAIChE
ioMosaic Corporation
93 Stiles Road | Salem, NH 03079
melhem@ioMosaic.com

Matthew L. Brewer
BP
150 W. Warrenville Road
Naperville, IL 60563
brewerml@bp.com

Michael K. Porter
Chevron USA, Inc.
1400 Smith Street
Houston, Texas 77002
portemk@chevron.com

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brewerml@bp.com

Michael K. Porter
Chevron USA, Inc.
1400 Smith Street
Houston, Texas 77002
portemk@chevron.com

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Abstract

The loss of High Pressure (HP) / Low Pressure (LP) interface has to be evaluated in order to (a) ensure that the downstream equipment can handle the energy and/or mass accumulation and (b) to also ensure that the upstream equipment can handle the rapid depressurization. The loss of the HP/LP interface can occur as a result of automatic controls failure/malfunction, and/or inadvertent valve opening / human error as seen in Figure 1.

Potential outcomes of the loss of the HP/LP interface include but may not necessarily be limited to:

1. Overfilling of the downstream equipment, i.e. as the liquid is displaced from the upstream equipment into the downstream equipment.

2. High temperatures associated with the rapid compression of the vapor space of the downstream equipment associated with liquid displacement.

3. Multiphase flow associated with increasing liquid level, liquid entrainment, and/or high superficial vapor velocities caused by pressure relief device actuation in the downstream equipment.

4. Low temperatures caused by rapid depressurization in the upstream equipment and ultimately in the downstream equipment as pressure is further reduced when the gas breaks through to the downstream equipment.

5. Possible hydrate formation in the upstream and/or downstream equipment.

1 Sequence of Events Following Loss of HP/LP Interface

Upon loss of the HP/LP interface, for example a control valve fails wide open, the high pressure gas will initially push the liquid out of the HP vessel through the control valve and the associated piping to the downstream vessel. For most high pressure systems, the flow will be limited by and/or choked at the flow limiting element, typically the control valve or a flow restriction device that is placed downstream of the control valve.

For systems where the high pressure vessel contents are superheated relative to the downstream conditions, flashing will occur across the flow limiting element. If flashing flow occurs across the flow limiting element, the associated flow rates can be substantially lower than those for all liquid flow, and the choke points can range from as low as 30 % to as high as 90 % of the upstream pressure.

As the liquid is pushed downstream, the liquid level will increase in the downstream equipment causing the compression of the vapor space. As the vapor space of the downstream equipment is compressed, the pressure in the vapor space will increase until the lowest relief device opening pressure is reached. Depending on the liquid fill level at which the relief device(s) open, the nature of fluid (foamy, dirty service, non-foamy, clean service), and the initial superficial vapor or liquid entrainment velocities associated with the initial gas relief, the flow will either be single phase or two-phase.

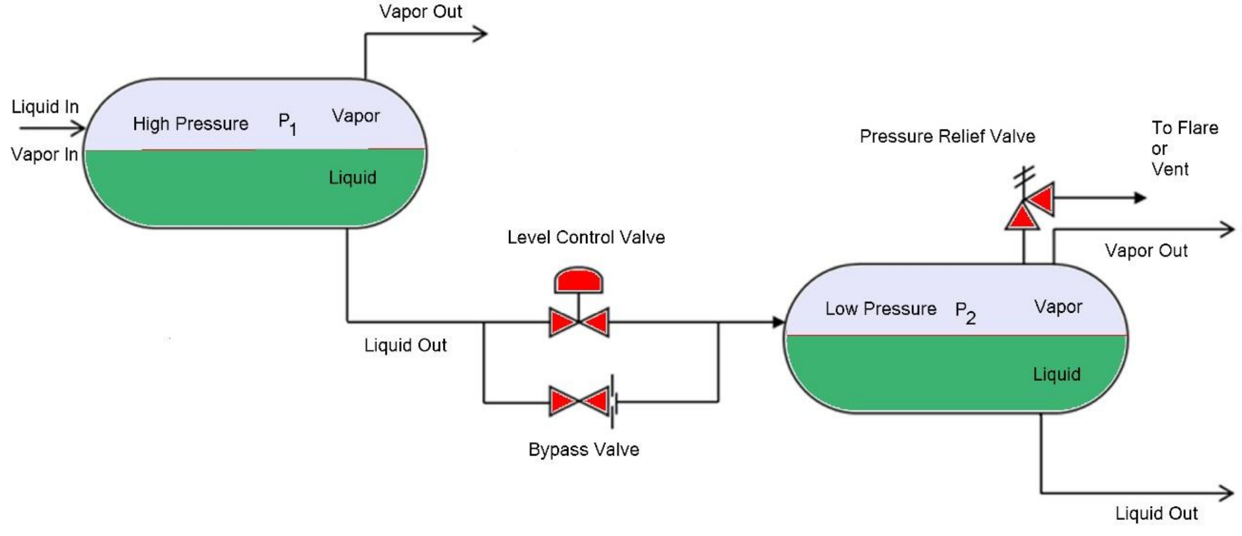


Figure 1. Typical High Pressure / Low Pressure Interface

If the flow is single phase due to the non-flashing conditions of the incoming liquid flow, the Pressure Relief Valves (PRVs) are expected to cycle rapidly (since most installations will have sized the PRVs for all gas flow and will have taken credit for operator intervention), until the downstream vessel becomes near liquid full. At near liquid full conditions, two-phase flow will occur. This is followed rapidly by single phase liquid flow. If the PRVs are not large enough to expel enough liquid to accommodate the volume expansion required by the expanding high pressure gas into the downstream volume (piping and equipment), the pressure will ultimately reach the upstream source pressure. This is a high consequence scenario, and is often referred to as liquid displacement.

In some circumstances, normal liquid flow will continue out of the downstream equipment throughout the loss of interface event. As the pressure increases beyond the set point of the relief device(s) and up to the allowable pressure accumulation, more flow can occur than the normal flow. If this flow is high enough compared to the incoming liquid flow and what is being vented by the relief devices, a liquid full condition may not be possible.

If the flow out of the downstream equipment relief device(s) is two-phase due to the flashing nature of the incoming fluid, the pressure relief device(s) would be expected to cycle with a lower frequency and to stay open (since they typically will have been sized for all vapor flow). Depending on the rate of flow of superheated liquid, the degree of superheat, the initial liquid fill level of the downstream equipment, and the set point of the relief devices, an all liquid full condition may never occur. Two-phase flashing flow is highly likely in this case. The vapor quality of what is vented and vapor/liquid disengagement characteristics can be estimated using the DIERS coupling equation (see [1]).

If the downstream equipment does not become liquid full and the liquid inventory in the upstream equipment is depleted, all vapor/gas vapor flow or break through will occur. In this case, the rate of depressurization and fluid characteristics will determine the low temperature levels reached in the upstream equipment. If the liquid level in the downstream equipment is high enough or the superficial vapor velocity is high enough two-phase flow will occur. If the liquid level is low enough, all vapor flow will occur.

One should note that the initial vapor phase composition in the downstream equipment is typically different from the vapor space composition of the upstream equipment. The difference in molecular weights should be considered in pressure relief design for the downstream equipment as well as the increase in temperature that is caused by the gas compression downstream due to the incoming upstream high pressure gas.

It can be shown (See [1]) that for a downstream vessel that is initially full of vapor, the gas temperature can reach $(C_p/C_v)*T_u$, where T_u is the upstream gas temperature. This can only occur if the pressure in the downstream vessel can reach the upstream pressure which typically does not happen if the pressure relief devices are properly sized. Still, the increase in downstream gas temperature up to allowable pressure accumulation limit should be considered in conjunction with the vapor space mixture molecular weights composition for developing the relief requirements.

2 Steady State Solutions

Despite the intricate nature of sequence of events following the loss of HP/LP interface, many companies still perform pressure relief systems design and evaluation using steady state methods. Although these methods are typically conservative, they can sometime be non-conservative - for example the difference in molecular weights between the upstream and downstream vapor spaces is not normally considered in steady state pressure relief design, two-phase flow is not considered, etc.

Two calculations are normally performed:

1. Gas blow through only – In this case the high pressure gas is expanded to 1.1 or 1.16 times the set point of the pressure relief devices downstream. The relief device(s) flow area(s) is selected such that no pressure accumulation is possible above the allowable limits (1.1 or 1.16 x MAWP).

2. Liquid displacement – In this case the rate of volume creation by the expanding gas must be accommodated by the rate of volume creation in the downstream equipment by the liquid being expelled/vented through the relief devices. This scenario will typically result in very large relief requirements primarily because of the difference of liquid and vapor densities (we need to conserve volume), and in many cases pressure relief is either not possible or not desirable. Large pressure relief valves will require substantial structural support, large downstream separation equipment, and will most likely cause chatter when they are actuated by scenarios other than the liquid displacement scenario.

   Prior to gas breakthrough the flow limiting element we can show, even under liquid full conditions, that the required downstream equipment relief flow area is a small multiple of the flow area of the flow limiting element. If we ignore frictional pressure loss, elevation changes, and assume incompressible flow, we can show the following:

$$\frac{A_{PRV}}{A_{CV}} = \sqrt{\frac{P_H - P_L}{P_L - P_B}}$$

   Where A is the ideal flow area, P_H is the upstream pressure, P_L is the maximum allowable downstream pressure (typically 1.1 or 1.16 x MAWP), and P_B is the ambient or flare header backpressure. We can also solve this equation for P_L for a given downstream relief flow area:

$$P_L = \frac{A_{CV}^2 P_H + A_{PRV}^2 P_B}{A_{CV}^2 + A_{PRV}^2}$$

   For typical installations where the backpressure is low compared to the upstream pressure,

$$P_L \approx \frac{A_{CV}^2}{A_{CV}^2 + A_{PRV}^2} P_H$$

Once the gas finally breaks through the flow limiting element, it expands and cools. For high pressure systems the gas flow will typically be choked at the flow limiting element. The gas will further expand from the choke as volume is created. This expansion or gas volume creation has to be accommodated by volume creation from all liquid flow through the pressure relief device(s). Large relief flow areas will typically be required in order to accommodate the expanding gas. If sufficient relief flow area is not available, the pressure in the downstream piping from the flow limiting element can reach the upstream source pressure.

3 Dynamic Solutions

These solutions are the most prudent choice for this class of scenarios, especially considering the implications of Section 2 above. Using dynamics one can address all the issues highlighted above dealing with temperature effects, multi-phase flow, and overfill, to name a few. More importantly, modeling the dynamics of the system will enable the quantification of the maximum pressure reached in the downstream piping and equipment if the relief capacity is undersized which will provide a more accurate representation of the overpressure risk while interim and permanent risk reductions measures are being explored.

Dynamic simulations are possible with computer codes like SuperChems™ Expert[1] and HYSYS® Dynamics. SuperChems™ Expert is more focused on relief systems estimates and has full dynamic implementation of the DIERS coupling equation. SuperChems™ Expert enables the user to connect and establish single or multiphase flow dynamics from one or more interconnected vessels with and without chemical reactions. This software is also used extensively to model simple and complex depressuring systems.

4 Steps Involved in Steady State Estimates

The following steps are normally used when steady state estimates are used to establish (a) if liquid displacement will occur, and/or (b) to determine the required relief area once the gas breaks through.

1. Calculate the flow from the control valve or flow limiting element (liquid flow and gas flow). Note that typical Instrument Society of America (ISA) control flow equations are not suitable or recommended for multiphase flow or where gas compressibility is substantial or near the phase boundaries. (See [2]). If flashing flow occurs, estimate the vapor flash fraction at the downstream vessel maximum allowable pressure accumulation, typically 1.1 or 1.16 x MAWP. Also account for the piping resistance and pressure drop, especially for flashing multiphase flow and/or when both the control valve and/or bypass valve are assumed to be wide open, i.e., when there is significant pressure drop in the piping.

[1] SuperChems™ is a component of Process Safety Office™

In many practical cases and even after the liquid level is depleted in the upstream equipment, liquid and gas may continue to flow into the upstream equipment. The fluid that breaks through to the downstream equipment will not be 100 % gas but a mixture of liquid and gas that has the same vapor / liquid ratio as the incoming flow. Even if the flow is frozen (i.e., no flashing occurs through the flow limiting element), the flow rate through the flow limiting element can be substantially smaller than that for all gas flow depending on the vapor quality of the mixture.

2. Calculate the liquid inventory in the upstream vessel. This will be scenario dependent and may be different during startup, upset conditions, or normal operations. Usually, normal operating liquid level or high liquid levels are considered.

3. Calculate the liquid inventory in the piping (upstream and downstream of the control valve as appropriate). This is typically small compared to the inventory in the upstream vessel unless long runs and/or large diameter piping is used.

4. Calculate the liquid inventory in the downstream vessel and its available vapor space volume. Typically the nozzle height above the liquid level should be considered in establishing the vapor space volume. This will also be scenario dependent (startup, upset conditions, or normal operations). Normal operating liquid level and/or high liquid levels are usually considered.

5. Establish any normal inflows and outflows that need to be considered for both the upstream and downstream vessels.

6. Check if there is enough liquid inventory to fill the downstream vessel. If there is, check the required flow area for all liquid flow before gas breakthrough occurs at 1.1 or 1.16 x MAWP of the downstream vessel.

7. If liquid displacement is not possible, use the DIERS coupling equation to check the downstream relief requirement for two-phase flow. Most existing installations do not account for two-phase flow caused by liquid carryover and/or high superficial vapor velocities leading to two-phase flow. If two-phase flow is not predicted using the DIERS coupling equation, confirm that the existing relief capacity is adequate for all gas flow. Note the initial difference in molecular weights and temperatures for the incoming upstream high pressure gas and the vapor space contents of the downstream equipment.

8. If liquid displacement is possible, calculate the volumetric expansion of gas through the control valve down to 1.1 or 1.16 x MAWP. Note that typically the flow will be limited by and/or choked through the flow limiting element and that flow calculated earlier (see Step 1) is based on an isentropic flow path. Further expansion downstream from the choke point conserves stagnation enthalpy. Temperature changes due to expansion will have a significant impact on the relief volumetric flow rate of liquid needed to accommodate the expanding gas.

9. Calculate the required relief area at 1.1 or 1.6 x MAWP that can provide the same liquid volumetric flow rate as the expanding gas rate. If the required relief area is larger than

what is installed, consider modeling the dynamics of the two-phase vessel system, especially for flashing flow to establish if overfilling will occur since a sufficient amount of liquid can be vented during two-phase flow.

5 HPV / LPV Liquid Displacement Case Study

We consider the potential loss of HP/LP interface in a hydrocarbon processing facility from a high pressure vessel (HPV) into a low pressure vessel (LPV). We will consider several permutations of scenarios that can potentially lead to liquid displacement in LPV from HPV. These scenarios include different conditions for normal operation and for startup. Both steady state and dynamic solutions are provided to complete the pressure relief design dossier for LPV. Information pertaining to HPV and LPV is shown in Table 1.

Table 1. Liquid Displacement Scenarios Data Summaries

HPV (Upstream Vessel)	LPV (Downstream Vessel)	Flow Limiting Element	Relief System
Horizontal Cylindrical Vessel with Hemispherical Heads	Horizontal Cylindrical Vessel with 2:1 Elliptical Heads	Control Valve with a Port size = 2.875 in	Piping liquid inventory = 0.9 m^3
Liquid volume = 47.82 m^3 or 47.6 %	Liquid volume = 39.36 m^3 or 64.8 %	C_v = 23	Two 6Q8 PRVs in service
Total volume = 100.46 m^3	Total volume = 60.709 m^3	Effective control valve ideal flow area = 0.605 in^2	Set points of 150 and 157 psig
P = 1900 psig	P = 75 psig		API Flow Area = 11 in^2
T = 125 F	T = 133 F		Liquid C_d = 0.71
	MAWP = 150 psig		Gas / Two-phase C_d = 0.975
Initial contents in vapor space, hydrogen and some light hydrocarbons (M_w ranging from 2 to 3.9)	Initial contents in vapor space, hydrogen and some light hydrocarbons (M_w ranging from 2 to 3.9)		Total existing relief flow area = 22 in^2
Initial contents in liquid space, sour water	Initial contents in liquid space, sour water		
	Normal outflow rate for sour water = 600 GPM		

The governing scenario is a single control valve failure with the bypass valve closed. We need to calculate if liquid displacement followed by gas breakthrough is possible and what the required relief area will be using both steady state and dynamics simulation estimates. Four permutations are considered for this scenario as shown in Table 2.

Table 2. Dynamic Simulation Cases Initial Conditions

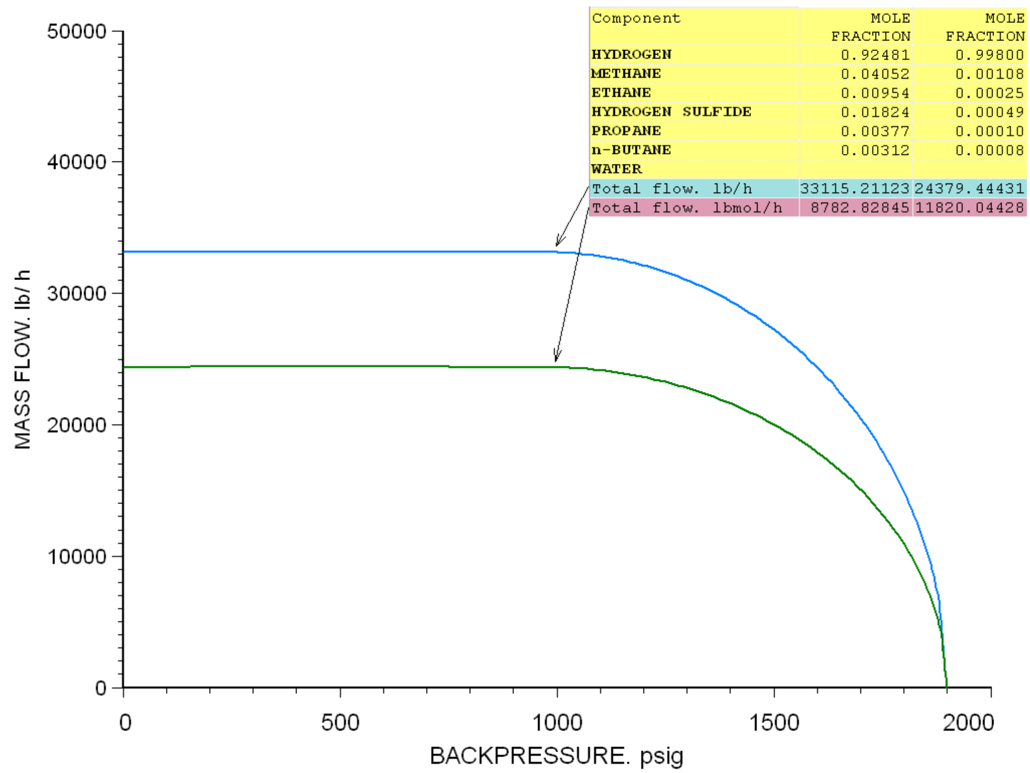
Case A	Case B	Case C	Case D
Startup	Startup	Startup	Normal Operation
Gas M_w = 2	Gas M_w = 3.7	Gas M_w = 2 and 3.7	Gas M_w = 3.7
Control valve wide open	Control valve wide open	Control valve wide open	Control valve wide open
Only gas flow into HPV	Only gas flow into HPV	Only gas flow into HPV	Gas and liquid flow into HPV
600 GPM liquid outflow from LPV	600 GPM liquid outflow from LPV	Outflow from LPV is blocked	Normal inflow and outflows from HPV and LPV
HPV P = 1900 psig	HPV P = 1900 psig	HPV P = 1900 psig	HPV P = 1835 psig

Steady State Solution – Step 1

Step one of the steady state solution (and also an important step in the dynamic estimate) is to calculate the flow rate that is limited by the control valve. A single value is required for steady state estimates at a backpressure of 174 psig (1.16 x 150) since there are two active relief devices on LPV. For dynamic source estimates, especially where an infinite pressure source is desired, a flow curve is typically established that provides flow mass and energy as a function of backpressure. This is shown in Figure 2.

More flow is calculated for the end of run conditions with a molecular weight of 3.77 than the start of run conditions with a molecular weight of 2.063. Note that the flow chokes at 977.5 psig and 34.7°F for the end of run conditions and 974 psig and 21.7°F for the start of run conditions. The flow curves illustrated above were calculated using VdP integration in SuperChems™ Expert. These methods are preferred over the ISA method for control valve, especially for two-phase flow. Subsequent constant stagnation enthalpy expansion to 174 psig yields temperatures of 18.3°F and 20.6°F respectively and a volumetric gas flow of 6,760 m^3/hr and 9,177 m^3/hr.

Because the starting sour water temperature is 125°F, the flow will be all liquid with a rate of 473,116 lbs/hr or 218.15 m^3/hr (960.5 GPM).



Component	MOLE FRACTION	MOLE FRACTION
HYDROGEN	0.92481	0.99800
METHANE	0.04052	0.00108
ETHANE	0.00954	0.00025
HYDROGEN SULFIDE	0.01824	0.00049
PROPANE	0.00377	0.00010
n-BUTANE	0.00312	0.00008
WATER		
Total flow. lb/h	33115.21123	24379.44431
Total flow. lbmol/h	8782.82845	11820.04428

Figure 2. Calculated Gas Flow Rates from a Control Valve with a C_v = 23 Wide Open as a Function of Backpressure

Steady State Solution – Step 2

The liquid inventory in the upstream vessel is specified by the scenario identification team to be at 47.82 m^3.

Steady State Solution – Step 3

The total liquid volume of the piping connecting HPV to LPV is calculated at 0.83 m^3 based on a detailed isometric.

Steady State Solution – Step 4

The liquid inventory in the LPV is established by the scenario identification team to be 39.36 m^3. The vapor space volume is then 60.709 – 39.36 = 21.35 m^3.

Steady State Solution – Step 5

The normal inflows and outflows are shown in Table 1. For Cases A, B, and C only gas flows into HPV and for the startup cases, it is assumed that there is sufficient supply to keep the gas inflow for a long period of time, as seen in Figure 3. For Cases A, B, and D the sour water outflow out of LPV is specified at 600 GPM or 136.26 m^3/hr. The normal outflow of sour water is assumed to be blocked for Case C. Normal inflows and outflows are usually established from operating data and/or heat and material balances.

Steady State Solution – Step 6

The upstream liquid inventory is 47.82 + 0.83 = 48.65 m^3, while the available vapor space in the downstream equipment is 21.35 m^3. The net flow rate of liquid into the downstream equipment if outflow is not disabled for LPV is 81.89 m^3/hr. The total flow time of liquid from HPV is 48.65 / 218.15 = 0.223 /hr. The total amount of liquid that would be transferred to LPV with outflow enabled in LPV is 0.223 x 81.89 = 18.26 m^3. This would take the liquid level in LPV to 95 %. This existing PRVs on LPV have not been sized for two-phase flow and they should be checked.

If normal outflow is disabled on LPV, it would take 21.35/218.15 = 0.0978 /hr to fill LPV. If LPV becomes liquid full, the pressure exhibited by the downstream equipment and piping downstream of the control valve would reach 1900 psig in 0.0978 hours.

Steady State Solution – Step 7

Since liquid displacement is not possible for Cases A, B, and D, the existing relief devices would need to be checked for two-phase flow. This estimate is duplicated in the dynamics solution provides later in this paper.

Steady State Solution – Step 8

Since liquid displacement is possible for Case C, we need to calculate the required relief area needed to relieve 6,760 m^3/hr and 9,177 m^3/hr of liquid for the end of run and start of run conditions.

Steady State Solution – Step 9

The required relief ideal flow area is calculated to be 61.58 in^2 and 83.6 in^2 respectively, well in excess of the existing relief capacity. These would be the ideal flow areas needed to flow 6,760 m^3/hr and 9,177 m^3/hr of sour water to allow a sufficient volume for gas expansion.

6 Gas Dynamics Solution

The gas dynamics solution was generated using SuperChems™ Expert. For this solution the user first creates a variety of objects to represent components of the systems such as vessels, relief devices, piping segment, mixtures, etc. A built-in thermodynamics package is selected. Scenarios are then created to collect the objects. Once the scenarios are created, flow dynamics models area associated with the individual scenarios and the dynamic solutions are produced. Figure 3 shows the pressure history in LPV for Cases A and B with outflow continuing in LPV.

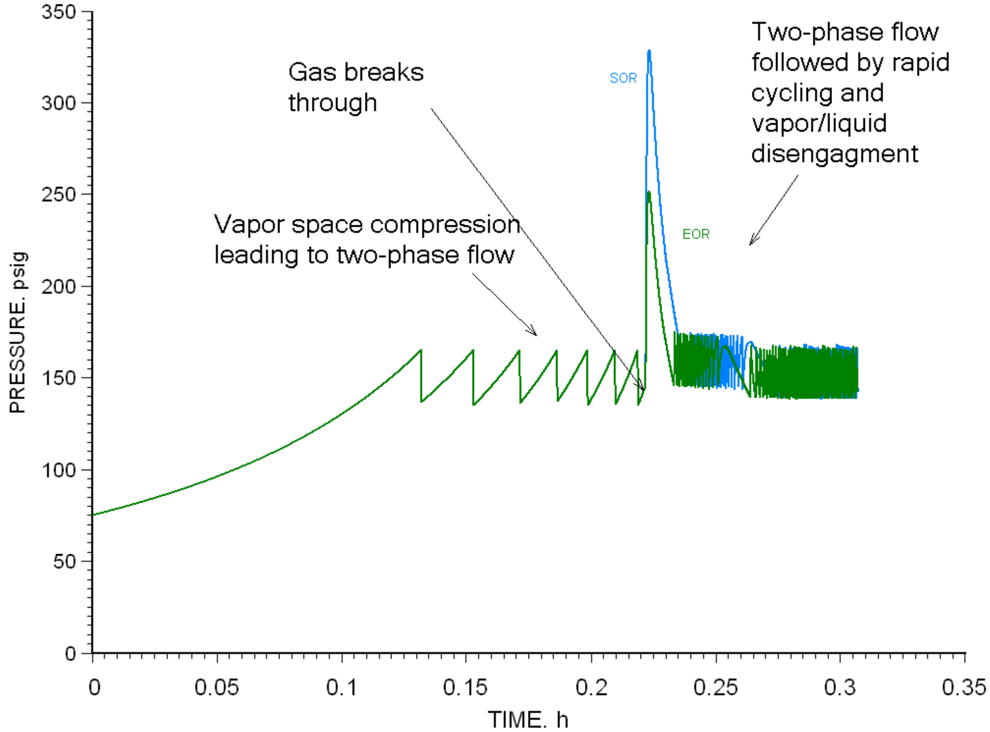


Figure 3. Pressure Histories for Cases A and B in LPV

We note that the start of run condition with a molecular weight of 2 causes a higher pressure level in LPV than the end of run condition with a molecular weight of 3.77. We also note that the existing 6Q8 relief devices are oversized for single phase gas flow.

Figures 4 and 5 illustrate the flow rates from both relief devices (Primary and secondary) piping as well as the constant liquid outflow (Bottom) for Case A with the start of run gas molecular weight of 2. We note that a single PRV is open during the period where the vapor space gas is compressed and that the flow is predominantly liquid. We also note that when the gas finally breaks through both PRVs open for short duration followed by one PRV flowing only. The flow out of the single PRV becomes all vapor when the liquid level drops to a level where the DIERS coupling equation predicts vapor/liquid disengagement.

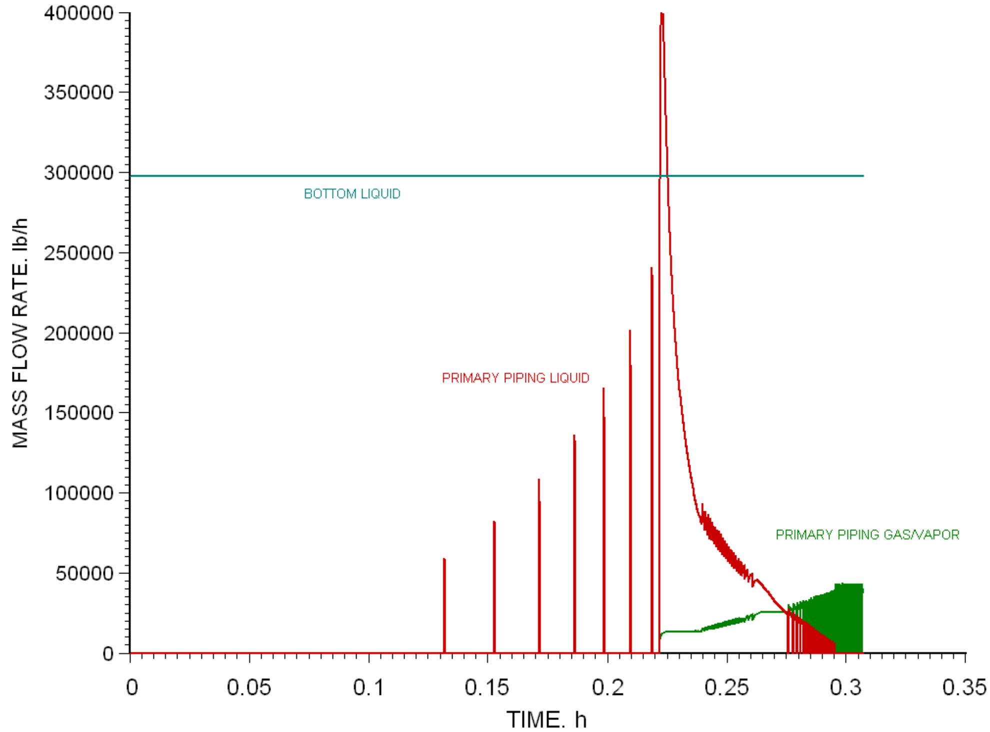


Figure 4. Flow History from First PRV Set at 150 psig

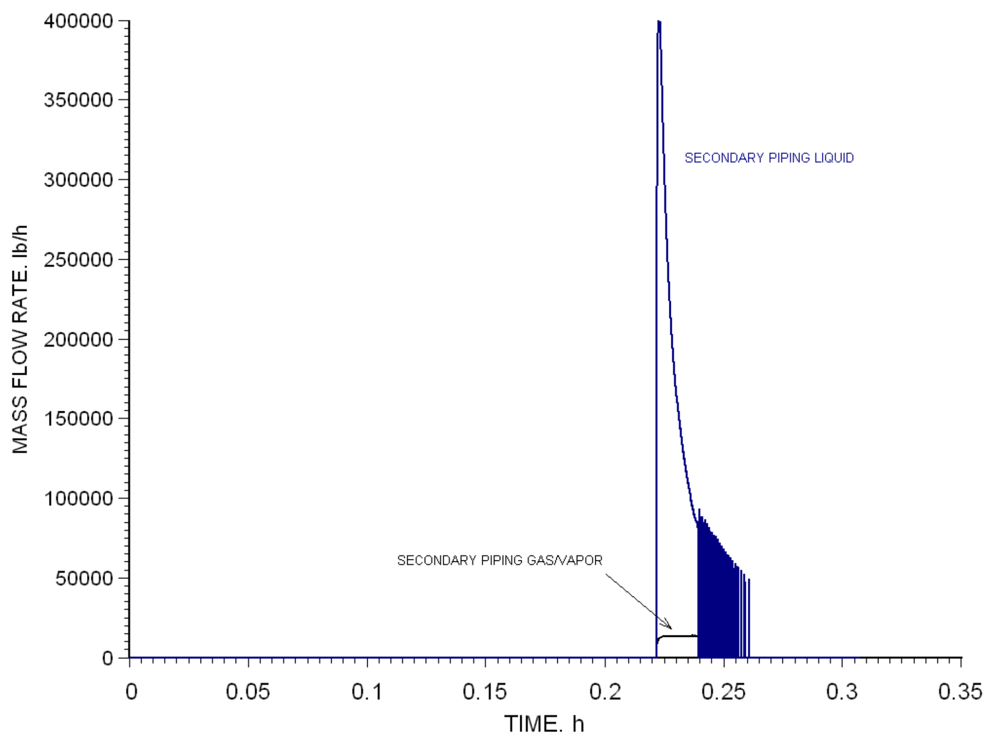


Figure 5. Flow History from Second PRV Set at 157 psig

Figure 6 illustrates the liquid level reached in LPV for Cases A, B, and C. In Case C, without normal outflow from LPV, liquid full conditions are reached and the pressure in LPV and any piping downstream of the control valve is expected to reach the upstream pressure value of 1900 psig.

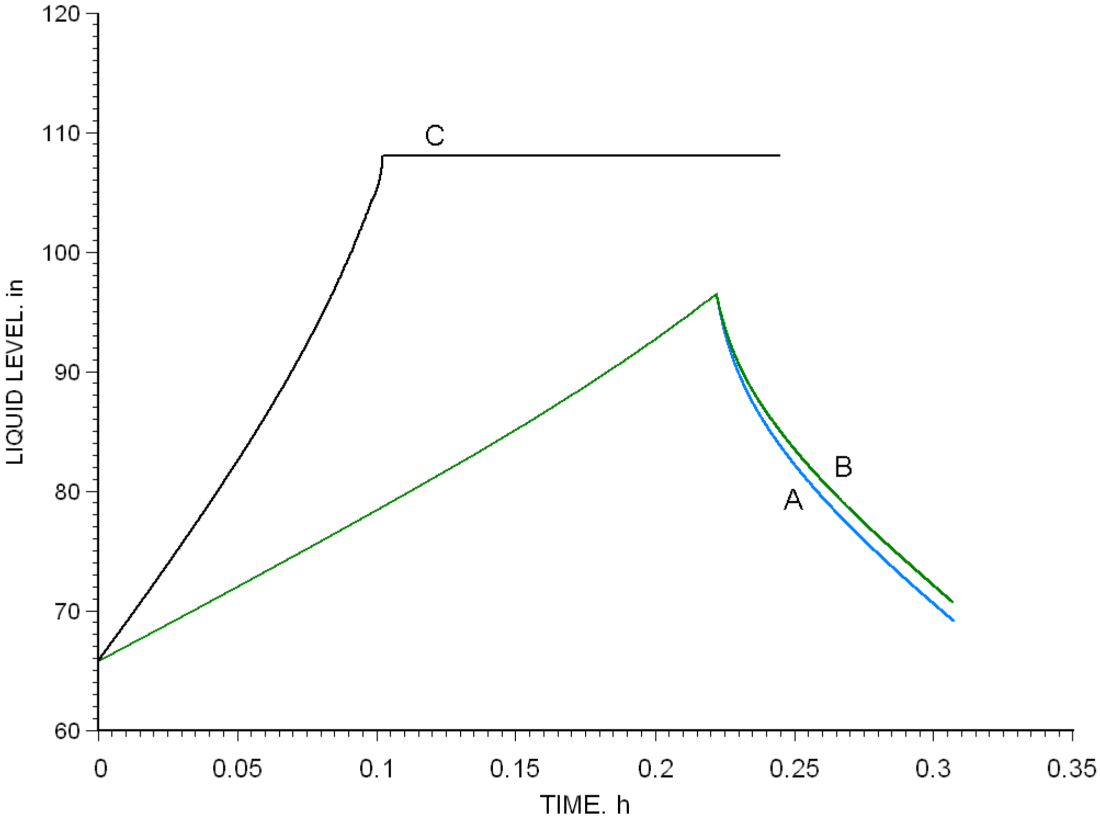


Figure 6. Liquid Level Histories for Cases A, B, and C

7 Practical Solutions for Existing Systems

The relief requirements for liquid displacement scenarios can be very large. It may not be practical or desirable to outfit the downstream low pressure equipment with large or multiple relief devices because of reaction forces, support requirements, increased loads on process separation equipment, and capacity of disposal systems, to name a few reasons.

If large pressure relief devices are to be installed, ensure that other process induced scenarios are vented using smaller devices at lower set points. One should also consider the tradeoffs of lifecycle cost between inherently safer, passive, active and procedural risk reduction measures. All four categories can produce tolerable risk levels but their associated lifecycle cost may be different. Inherently safer alternatives involve a higher upfront initial capital cost while procedural risk reduction alternatives will have continuing maintenance and operating costs.

There are many practical solutions that can be used to reduce and manage the risk of these scenarios including but not limited to:

1. Use of a relief device on the inlet line of the downstream LP equipment
2. Use of Safety Instrumented Systems (SIS) chopper valve(s)
3. Use of a restriction orifice on the primary piping
4. Use of a smaller or modified control valve
5. Provide alternate relief paths
6. Decrease operating liquid levels

7.1 Use of Inlet Line PRV

As indicated earlier, most existing oil and gas installations relief systems are typically designed for all vapor flow. A high pressure gas and/or two-phase breakthrough leading to two-phase flow out of the downstream low pressure equipment cannot be adequately handled by the exiting relief devices. This will often lead to vessel overpressure beyond what is allowed by Recognized and Generally Accepted Good Engineering Practices (RAGAGEP).

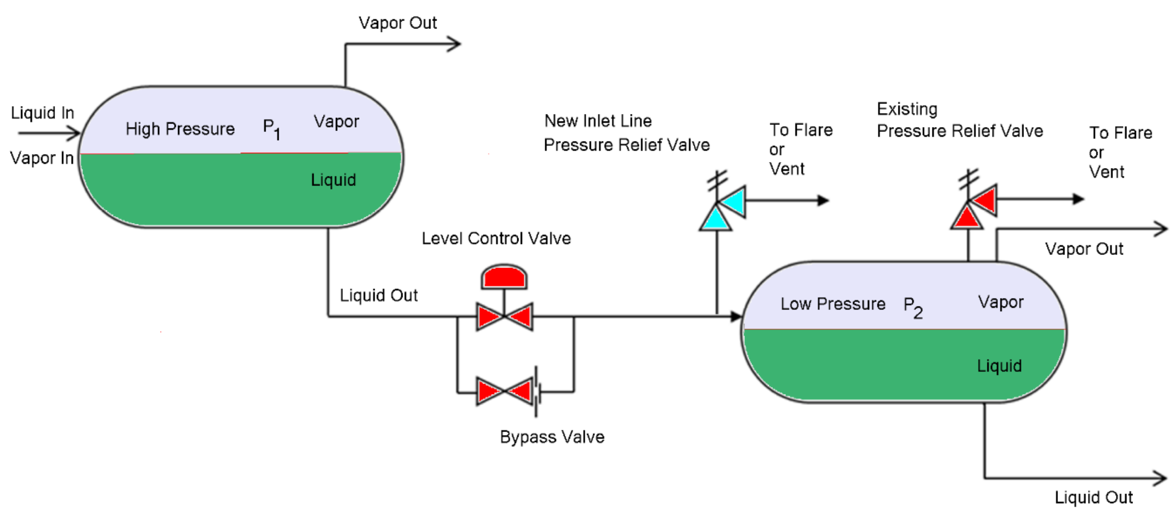


Figure 7. Use of Inlet Line PRV to Mitigate Liquid Displacement

A pressure relief valve can be installed on the inlet line connecting the HP interface to the LP interface, downstream of the flow limiting element, i.e. orifice, control and/or bypass valve. The need for locating pressure relief devices on the inlet line to the low pressure vessel is recognized in the relief system standards of some major hydrocarbon processing companies. A single pressure relief valve is preferred (with a spare). The set point is typically set higher than the set point for the existing pressure relief valves on the downstream equipment such that all other pressure relief scenarios can be handled by the existing valves and the gas / two-phase breakthrough can be handled by the inlet line valve and the existing valve(s). Note that if the LP vessel and the inlet piping downstream of the HP level control valve are liquid packed prior to a full gas breakthrough scenario, then even having the pressure relief device on the inlet line may still results in impractical pressure relief device sizes.

Careful consideration should be given to set point selection, structural support, vortex shedding, and valve trim selection. Liquid or universal trims are recommended where two-phase and/or liquid flow cannot be completely ruled out to minimize the potential for valve chatter. Note that valves with liquid trims will typically have higher blowdown values than valves with vapor trims. One may also want to consider installation of one or more check valves to isolate the new pressure relief valve from reverse flow from the downstream equipment. Many companies prohibit the use of check valves between the protected equipment and relief device.

Rupture disks or rupture pins can be used in combination with pressure relief valves to provide a more economical solution when compared with pressure relief valves alone for this scenario due to the large relief area requirements. Rupture discs/pins can be installed either on the vessel inlet or outlet to handle liquid displacement. The burst pressure of the rupture disk should be set higher than the primary pressure relief valves so that it only functions in the event of a liquid displacement event. Rupture disks or pins may not be acceptable due to prolonged product loss, but may be acceptable for lower pressure or utility systems.

In many cases, a sufficiently large single inlet line pressure relief valve is not feasible due to the inlet line size. Unless the pressure relief valve is large enough, overfilling may still occur. However, depending on the size of the inlet line pressure relief valve, more time may become available for operator response as a result.

7.2 SIS Chopper Valves

SIS Chopper valves can provide a practical solution when compared with the installation of multiple large pressure relief devices. SIS Chopper valves can be installed to stop flow either on low level in the upstream system or high level in the downstream system. For systems which have many multiple inlets, it may be less desirable to install multiple chopper valves due to the increased lifecycle cost and complexity of the system. Multiple chopper valves in series may be preferable to achieve an acceptable overall Safety Integrity Level (SIL) rating over providing a single chopper valve at a higher SIL rating.

7.3 Restriction Orifice on the Primary Piping

A restriction orifice can be installed to limit the inlet flow into the LP side system below the rate of outflow of the downstream system. If flow cannot be guaranteed during the relief event, then restricted flow can only prolong the event. The orifice should be designed to ensure that inflow does not exceed outflow of the system. This option can be designed to take into account the capacity of existing pressure relief systems. Consideration should be given to design of the orifice to limit flow to handle both the liquid displacement and full gas breakthrough scenarios.

7.4 Use of a Smaller or Modified Control Valve

Installing mechanical stops can provide a means to limit the travel of the control valve and thus limits flow. Mechanical stops may not be acceptable where the upstream maximum pressure is above the corrected hydrotest pressure of the downstream vessel. Mechanical stops may be able to provide an acceptable interim risk reduction method before permanent solutions can be implemented. Restricting the travel of the control valve by modifying the stem or providing a smaller trim valve can also reduce the C_v of the valve. The amount of travel that can be restricted can be limited by process considerations such as the amount of flow that is required.

7.5 Provide Alternate Relief Paths

Alternate relief paths can also provide an effective means of assuring that the LP side vessel does not become overfilled. Consideration can be made to ensure that liquid outflow continues during the event. If the system is pumped or pressured out, normal outflow may continue during the failure (e.g., failure of automatic controls) provided that there is not a cascading failure (power failure causing instrument air failure). Vapor outlet lines or overflow lines can also provide an effective means of ensuring that the liquid full condition does not occur. Credit for liquid flow through vapor lines is not typically taken as a risk reduction credit. This credit may be considered if the vapor line can remain open during the event and can be shown to have sufficient capacity to handle the incoming liquid. Startup conditions should also be considered as systems may be blocked in or offline during initial inventory of equipment.

7.6 Decrease Operating Liquid Levels

If operating levels in the upstream system are greater than the downstream vessel vapor space, then liquid displacement can occur. Operating liquid levels may be adjusted either in the HP or LP side vessels to ensure that the liquid that is transferred does not overfill the downstream vessel. Consideration can be made to lower the operating level enough to avoid two-phase flow through the relief devices in addition to liquid displacement. Some companies may use the maximum level tap or high level alarm as the high level point for comparison for liquid displacement volumes, so this option may not be viable in many circumstances.

8 Conclusions

The protection of downstream low pressure vessels from loss of high pressure (HP) / low pressure (LP) interface is very challenging. Steady state methods often result in impractical relief requirements. The use of dynamic simulation tools can provide a lot of insight into how much time is available for operator response as well as the sensitivity of relief systems design to key variables such as composition, vessel geometry, fill levels, and initial conditions. Two-phase flow from downstream vessels should be considered in any relief systems design and/or evaluation.

9 Acknowledgements

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10 References

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