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Pressure relief devices: Does anyone has experience in the case of gas blow by (also gas blow through or blow through) from a production separator to tank. Any guidelines about this subject?

Germán Snaider
Ingeniero de procesos en FLUOR DANIEL SOUTH AMERICA LTD

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Teunis

Teunis Wolzak
Senior Process Engineer 1 at Jacobs Engineering, Leiden

You must avoid this situation by good instrumentation on the low level in your production sparator. SIL classification!
teunis

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Rodrigo

Rodrigo Lopes de Siqueira
Process Engineer

I don't know if I understood exactly what you want to know, but let's try, you should put the devices on over flow of the vessel to relief gas phase, else, if you install the devices on bottom vessel, the design for the devices would be too big to can relief sufficient volumetric liquid to control the pressure. It would be that your question? Can I help something more?

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Harshit

Harshit Gupta
Process Optimisation Support Engineer at Petrloleum Development Oman

I am not sure if I have understood your question, but are you talking about installing level control device to control the liquid level in your separator.

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Ajith Kayapurathu

Ajith Kayapurathu Chandy
Senior Process Engineer at Haldor Topsoe India Pvt Ltd

Kindly elaborate the question. Is it with respect to sizing a PSV or an operational constraint? If its in case of sizing, the PSV should be sized for gas blowby from the production separator to the tank if the normal liquid volume within the separator less than the normal vapor space in the tank, i.e., liquid in the separator will deplete to the downstream tank.

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S M

S M Kumar
Process Design Consultant

I presume you are concerned about a pressure relief device on tank in case of gas blowby. This is an interesting topic with no single answer. There are several ways to skin the cat or the issue as I mention below. Cost the is final decider!!

AA: Usual Practice
(1) Usual practice is to consider full open Cv of the liquid control valve at Prodn Sep outlet and assume that only gas flows thru the failed open LCV after the liquids in the Prodn Sep gets pushed-out to tank. The tank's normal vent or PVV may not be handle this gas flow, pressurizing it with a potential top rupture. Worst case scenarios are (2) higher Cv of the bypass across the LCV and at times thru (3) both the LCV and its bypass, based on the assumption, that operator opened the bypass when the LCV malfunctioned and ended up opening it fully! So document your basis with your client. Since you work for an engg company Fluor, I assume it is engg work for a client. Whether (1) Cv of LCV alone or (2) Cv of bypass across LCV or (3) Both LCV and its bypass.

(2) Next get an agreement on upstream pressure - whether it is normal operating, PAH (Pressure high alarm) or PAHH (Pressure high high trip) or PSV set pressure. Each client has their own say. Good to assume PAH only as the PAHH would have closed the inlet and isolated the Prodn Separator.

(3) Similarly get the liquid level in tank fixed, Normal, LAL or LAHH level. Here again, good to take it at LAH

(4) Calculate the gas flowrate based on agreed Cv in step 1. [By the way: Remember to add the pressure drop in the fittings, bends in the pipeline between Prodn Sep and Tank. Usually process engineers forget this. This can add a lot of resistance and reduce flow.] Usually it will be a high number than the normal gas inflow into Prod Sep. The argument for the high flow is that when gas blowby occurs the instantaneous or momentary outflow can be high. Tanks usually come with or provided with a Blow-off hatch as a fire case PSV. Check if your atnk has one and if this rate can be handled by the Blowoff hatch (not by the PCV, if any on blanketing line or PVV). If yes. problem is closed, the easy way. If tank blow-off hatch can not handle the high flow, consider deleting the LCV bypass or add a RO on it to make its Cv the same as LCV. Similarly, consider 2 - 50% LCVs acted on by 2independent LTs if it helps. If yes. problem is closed. Still no way out??

BB: The Difficult but makes you feel happy and proud option.
If you have the aptitude and skill + access to Hysys Dynamics, set up a system from Prodn Sep upstream to Storage Tank vent. Based on the inputs (reduced bypass CV or 2-50% LCVs), while the instantaneous rate is high, the vapor volume on top of tank LAH may be able to handle the small gas quantity that gets transferred in the initial few milliseconds with hardly any pressure rise, mosquito bite on an elephant! Check. Then, ha ha, because of more gas outflow than inflow, the source pressure (Prodn sep pressure) will drop, that is pressure is not a constant number, reducing further gas flow. AND gas alone will not go continuously as the traditional method assumes; once the gas inventory in Prodn Sep is out, liquid will also go to tank via the fail open LCV, effectively reducing Cv available for gas blowby and gas blowby rate. At this time, gas:liquid flows will be the same as normal inflow. I am not able to visualize the numbers. Tank pressure will be decided by its vent capacity. Prodn Sep pressure will float on Tank pressure. As the LCV has failed FULLY open, it is possible that both gas and oil can pass with an increase or decrease in normal Prodn Sep pressure. If it falls, Prodn Sep vapor outlet PCV, if one will pinch to maintain pressure or it may open more to allow gas flow. That is vapor flow will be split between vapor and liquid outlets, as decided by system resistance. Usually the tank vent or blow-off hatch should be able to handle the resultant realistic flow.

See part 2.

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S M

S M Kumar
Process Design Consultant

Part 2. Had to split into 2 to stay within character limits

Note: Since you are looking at sizing Relief Valve, you can't take credit for any instrument response, as they could have been bypassed or failed latently. So no credit for PALL/LALL on Prodn Sep or PAHH/LAHH on tank closing Prodn Sep outlet SDV. If you wish to avoid a PSV, then HIPPS/IPF is the route that could complicate and become expensive needing independent transmitters, wiring etc. Do provide LALL/PALL on Prodn Sep to close its outlet SDV and PAHH/LAHH on Tank to close the same SDV. Since the Prodn Sep is directly sending liquid to tank, I guess the Prodn Sep operates at low a pressure, say below 250 kPa. In that case, the liquid outlet LCV/SDV may be large, considering static head liquid in tank. If the SDV is 16", it may take 16-20 sec to close. So make sure that LALL in Prodn Sep is set high enough above expected sand/ mud layer of 6"-8", so that liquid inventory between mud and LALL is adequate to allow SDV to close without gas blowby initiating. I hear in Hazops, "LALL provided" in chorus. But when I ask, is it high enough to stop gas blowby, silence is the usual answer!

Trust this helps

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Tochi

Tochi Offor
Manager, Process/SMR at TEP Nigeria

Some information are already provided to get you thinking and heading the right direction. In addition, I recommend to install the PSV upstream the Tank (on interconnecting pipe to storage tank) which is safer and more cost effective than losing part or all Tank. It's OK to install directly on "Tank" if this refers to an ATM separator.

For the system engineering, max flow thru LCV on the Prod Sep should be estimated. It is good approximation to assume only gas. Therefore, limiting flow can be derived from Eqns in Fig 4-30 & Eqns 4-9/4-10 in GPSA noting that X is obtained at inception of critical flow: Pc/P1= (2/k+1)exp (k/k-1) {Pc & P1 are absolute crititcal and inlet pressures respectively}. Note: X=(P1-Pc)/P1. For Eqn 4-9, in absence of vendor data, Y may be estimated using Xc values from Fig 4-32. The estimated max flow via the vena contracta should be applied to size the PSV. It is OK to add a margin to the flow rate for estimation error (don't forget to document that as well). Hope this helps.
Furthermore, to limit inventory to the PSV:
1) consider specifying the LCV bypass valve as LC/CSC if manual
2) optimize the LCV Cv to only what is essential (oversized LCV increases flow to the PSV); this could mean changing an existing LCV if cost and replacement impact is justified compared to the final PSV size.

Cheers.

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VIMALESH

VIMALESH AGNIHOTRI
Senior Process Engineer at Engineers India Limited

thanks kumar sir for giving such elaborated description.

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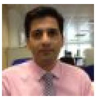

Saeid Rahimi Mofrad
Senior Specialty Process Engineer at Fluor

Since I like Kumar's comment, I am going to follow his style in writing:

AA) usual practice

1) Find the rated (100% open) CV of the LCV from instrument datasheet. Add Cv of fully open bypass valve to it if none of above recommendations about reducing system capacity is applied.

2) Assume P1 = PAH or PAHH or PSV set/relieving pressure

3) Assume P2 = tank operating pressure + LALL (not LAHH as recommended by Kumar) to get conservative flow rate)

where P1 and P2 are pressure at control valve inlet and outlet flanges.

4) Use control valve sizing equation to find gas flow rate (W) from separator to tank

5) Recalculate P1 and P2 considering the pressure drop of lines upstream and downstream of control valve flanges when W is passing through system (Note 1)

6) Go back to step 4 and recalculate W. Repeat till there is no major difference between W from subsequent calculations

Note 1:since usually large flow is established in gas blow-by condition, pressure drop of gas in control valve inlet and outlet lines can be so high that incompressible flow equation (dP= f L Ro v^2/2D) won’t be applicable(accurate). To minimize the error, inlet and outlet line can be cut in pieces and pressure drop of each piece can be calculated using mentioned equation. Density of the gas for different pieces will be different.

See Part 2

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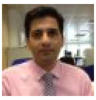

Saeid Rahimi Mofrad
Senior Specialty Process Engineer at Fluor

BB) best practice

1) Assume P1 = PAH or PAHH or PSV set/relieving pressure

2) Assume P2 = tank operating pressure + LALL (or even no level)

3) Calculate the total system resistance between separator and tank including nozzle entrances

and exits, elbows, tees, reducers, valves (even control valve) and piping resistance using K = sigma Ki + f Li/Di (i is just counter for fittings, valves and piping)

4) Determine maximum flow through piping systems (usually called chocked or critical flow but it can be non-choked or sub-critical depending on system hydraulics) using Isothermal or adiabatic flow equations. Adiabatic equation is more conservative. This method normally works based on Mach number. It is also iterative because f (friction factor) is function of W as the main variable in the calculations. (Note 2)

You can refer to:

1) “Easily design safety relief vent header systems” - CEP, Oct 1997- Martin Westman, to get step wise approach for adiabatic method
2) “Gas Calculations: Don´t Choke”- Chemical Engineering- Jan 2000- Trey Walters, to know more about compressible flow concepts

Note 2: AA method usually results in gas blow-by flow rate up to about 1.5 times of BB method.

Regards
Saeid

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Germán

Germán Snaider
Ingeniero de procesos en FLUOR DANIEL SOUTH AMERICA LTD

Thanks everyone for such good advices!!!

I have to say that we are calculating the gas blow by using the equation
Pc/P1= (2/k+1)exp (k/k-1)

and then if it is higher than atmospheric pressure, we obtain the density for this critical flow pressure and with that we calculate the gas mass flow. If it is lower than atmospheric pressure, we use the density at atmospheric pressure.

We also calculate a pressure drop due to gas expansion untill the liquid seal is broken by the gas flow. This task is performed with the gas equation and Z factor.

This way of solving the problem assumes 100 % flow, is conservative, and we know that the gas flow is instantaneous, like a ballon when it breaks. We had many discussions with clients about the assumptions.

The pressure drop of fittings and pipe in this case, doesn`t help to low down the gas flow. (in other cases we have equipment that acted as a restriction, and no gas blow by event would ocurr there).

The client don`t want restriction orifices, and because of prices, they prefer other options than a shutdown valve at the liquid outlet. But, in the project i am working on nowadays, the gas blow by required three hatches (PSE`s) at the tank, and that issue started the new discussion, because previous designed plants has smaller hatches for the same design basis flowrate of similar bateries. (i believe that you say hatches for the pressure relief devices used in tanks for emergency venting only)

Well, i will try the dinamic simulation once again, as S M Kumar says, to get an idea of the time that this high gas rate lasts. Nevertheless, the tanks are design for 50 mm of water column (API 650 tank) so they are very fragile.

Thank everyone for the help !!!

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Manish

Manish Juneja
Front End Engineering Department- Petronas

Mr. S.M. Kumar & Mr. Saeid Rahimi Mofrad,
I want to add details about tank (Downstream vessel) pressure.
We are calculating maximum gas flow through LCV using selected CV.
Flow through LCV dependent on differential pressure across LCV.
You are correct that we can use PAH pressure of production separator, but for me it will be more conservative if I use PAHH. During abnormality, pressure can be in-between PAH and PAHH.
Normally Production separator liquid goes to some other pressure vessel, where we have to design PSV considering gas-blow by. PSV will operate at it's set pressure (design pressure of tank / downstream vessel). At the time of PSV pop-up, pressure difference across the LCV will be less and flow will be lower. We should consider this differential pressure to calculate the flow

through the PSV.
P1 = PAHH of production seperator
P2 = PSV set pressure (design pressure) of downstream tank / Vessel
LCV CV = selected CV of control valve at 100% open (not the calculated / required CV)
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S M

S M Kumar
Process Design Consultant

German: Looks like you are well prepared than I had assumed.

I agree that it is not worth adding a SDV in liquid outlet.

I have no idea about the size and run length of line to tank + whether it is top or bottom entry into tank. If the entry is on top, after the liquid in the outlet is blown-out, gas may have free passage to flow. But depending upon liquid inventory, CFD studies would tell that it is a tough job to push and "accelerate" the liquid ahead of gas. Ant pushing an elephant. This is what you see when you analyze any heat exchanger tube rupture to a cooling water network. This will slowdown gas rate any momentary pressure peak will be in the 150# line close to tank.

If the entry to tank is thru bottom, then liquid static head in tank would "continue" to resist flow and reduce rate.

I had suggested taking tank liquid at LAH, as that would give the lowest gas volume in tank and highest pressure peak. For calculating highest flow, you may consider LALL or no liquid in tank as suggested by Saeid, but this gives large gas volume and may not show any pressure spike.

Difficult to say without actual numbers. Alice in Wonderland. Would be interesting to know what the numbers finally told you.

Manish: As I wrote, whether P1 is PAH/PAHH, it is your choice. P2 is not PSV set pressure, but PSV relieving pressure (110%set) + Pressure drop in the connecting piping. Here tank's design pressure would be 2 kPa. It hardly matters. That is why I wrote "Tank pressure will be decided by its vent capacity."
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Guntur

Guntur Wicaksono
Sr. Process Engineer at ConocoPhillips Indonesia

Hi German,

I think it would be better that you have to provide a emergency device in the tank (e.g. gauge hatch) and it shall be sized precisely with considering gas blowby case, otherwise installing SDV on the vessel outlet (tank inlet) if you provide and SDV it has to be provided based on SIL study (I believe it can be SIL 2 or 3), it more cosely than providing a gauge hatch.

to be honest with you can not rely on the pressure instrumentation on the vessel bottom, it can broken anytime.
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Germán

Germán Snaider
Ingeniero de procesos en FLUOR DANIEL SOUTH AMERICA LTD

Kumar,

The pipe size is 6" sch 40, the pipe run to the tank is about 60 m, with a top entry on the tank (very near the tank roof, about 7m height).

The LCV is 6", cv=412.

The production separator is horizontal L(T/T)= 8m and D=2m. Operating pressure is 5,5 kg/cm2g and 50 ºC.

The gas is 84% methane, oil is light, about 35ª API, 145 cP at 50ªC, and 50% water cut.

Tank is 50 mm of water column design pressure, and the hatches are set to 44 mm of water column.

so i understand that you agree with the gas expansion pressure drop correction i`d performed?
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S M

S M Kumar
Process Design Consultant

German:
As per the traditional - high flow method, given P1=5.5 kgg, CV=412, 6" x 100m eqL to tank, after the liquid is blown to tank, gas rate is ≈ 30,000 kg/h, P2 in tank at hatch set press, Pout of LCV ≈ 4 kgg, making flow thru LCV sub-critical or subsonic. This is an estimate. Any line sizing program can give you the right number; Hysys has a routine. Flarenet may also be used, keeping tank max press as tip pressure. Then, you can check against blow-off hatch vendor catalogue for the flow/ hatch and its set press of 44mm.

Comments (1) If line size is only 6", an outlet SDV is worth adding (2) Tank design press looks low. I vaguely remember 150-200mm as standard numbers (3) Pressure drop in outlet seems to be restricting flow; otherwise sonic flow thru LCV is ≈ 38,000 kg/h (4) I am not agreeing/disagreeing with expansion pressure drop correction. A small issue, part of standard line sizing calc. And once you do Dynamic Simulation/ CFD analysis of the system that considers P1 falling with outflow>inflow, you may get realistic number. Good Luck!

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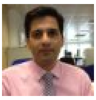

Saeid Rahimi Mofrad
Senior Specialty Process Engineer at Fluor

I did the same calculations based on following assumptions:

PAH: 1.1 x 5.5 barg = 6.0 barg
Gas MW: 20
Gas density: 5.32kg/m3 @ 6.0 barg & 50C
Gas viscosity: 0.01 Cp
Pipe Roughness: 0.0018"
Pipe upstream of LCV: 10m long, 6" dia
Pipe downstream of LCV: 100m long, 6" dia

AA method produces 29,600kg/hr based on Cv =412 (no bypass) , P1= 5.82 barg, P2= 4.53 barg.

I personally suggest adding SDV on separator outlet line closing at low-low liquid level (as first layer of over-pressure protection). But don't take it as an alternative to relief device unless it is upgraded to HIPPS requirements (and approved by Client). If you decided to install low-low level trip (SIL 2,3 or whatever rated), relief device as ultimate layer of protection would be still required.

Regards
Saeid

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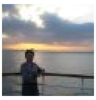

Susan

Susan Jiang
Senior Process Engineer at Technip

In addition, pressure relief systems may also include vent scrubbers, quench vessels, dump tanks, flame arrestors, vent silencers, high pressure relief systems, cold fluid relief systems or corrosive fluid relief systems. Note that vents used for the purpose of equipment pressure relief are considered to be part of a pressure relief system。

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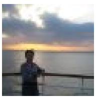

Susan

Susan Jiang
Senior Process Engineer at Technip

Codes, Standards of Practice and References：
API RP-520 (Part I and II) and API RP-521 recommended practices,
In particular, process engineers should be familiar with the relevant sections of the ASME Boiler and Pressure Vessel Code.

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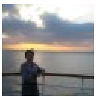

Susan

Susan Jiang
Senior Process Engineer at Technip

In very rare circumstances it is not practical or desirable to install a pressure relief device on a vessel per ASME Boiler and Pressure Vessel Code. This situation is possible, for example, when considering highly toxic chemicals, which may present a hazard in a pressure relief system. In these cases there is a code case exemption (Case 2211), which requires that overpressure protection be provided by system design. Application for this code case exemption should be approved by the Process Lead, Project Manager and Client and the requirements of the code case must be complied with. A copy of the case can be found in the AIChE “Guidelines for Pressure Relief and Effluent Handling Systems” or request the code case from the Mechanical Department.

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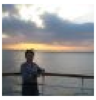

Susan

Susan Jiang

Senior Process Engineer at Technip

What you have to do is to analyze the process situation firstly,and find the right code and standard and process case to calculate.
Hope helpful!

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Lionel

Lionel Sheikboudhou

Process Engineer

I believe Kumar, Saeid and Manish have already given the key elements to calculate the relieving flowrate of this gas blow-by scenario.
I was just wondering if the scenario under discussion is credible? German, have you already compared the design pressures upstream and downstream the level control valve? I mean that if the downstream (tank side) design pressure is equal or higher that that upstream the LCV (separator side), the gas blow-by scenario study is irrelevant.

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Raheleh

Raheleh Sabounchi

Process Engineer

the failure of level control valve to regulate the flow of liquid from a higher to lower pressure system can result in loss of liquide inventory in upstream vessel and subsequent vapour flow to the lower pressure system.the relief load for this case shall be assumed to be the maximum vapour fow rate through the control valve.

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Germán

Germán Snaider

Ingeniero de procesos en FLUOR DANIEL SOUTH AMERICA LTD

Thanks everyone of you for your opinions!!!

I have to say that this issue is very common in upstream projects i am working on. The last one, was adressed considering the maximum instantenous flow rate of gas (peak flow of gas blow by) to design the gauge hatches emergency protection system for the stock tanks.
We also had to change the diameter of the level control valves of one production separator (that was oversized) from 6" to 4" to limit the flow rate of gas that can ocurr in this contingency.

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