Chemwork

PSV load calculation for HIPPS valve leakage rate

Mohammadreza Ebrahimi
Senior Process Eng. at Nargan Engineers & Constructors

Does everybody know how can calculate PSV load for valve leak rate scenario which is located downstream of HIPPS valve?

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andriyadi

andriyadi saputra
Process Engineer at PT. Tripatra Engineering

I have similar question on my mind.
Because my superior told me to assume that the 10% valve opening as leakage as worst case scenario that still come to smallest orifice "D" size as a result.
However, I still dont understand how to get that assumption.

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Mojtaba

Mojtaba Habibi
Process Engineer at Wood Group

I have seen some good engineering practice documents propose to consider HIPPS failure in case of multiple HIPPS (for example 1 HIPPS failure among 10 HIPPS packages) and HIPPS leake in case of single HIPPS package pending vendor information for exact value of leakage flow rate. But I have a doubt in my mind how likely is HIPPS failure scenario considering this fact that HIPPS is SIL rated package and how liekly is HIPPS leakage considering this fact that HIPPS valves are tight shut off valves. For current oil production project I asked HIPPS package vendor about HIPPS leakage scenario and the answer I received was this is not an issue to be concerned about. But on the other hand good engineering practice is developed based on real project experiences. So appreciate if other experts share their real project experiences on HIPPS failure and HIPPS leakage scenarions.

Best,
Mojtaba

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S M

S M Kumar
Process Design Consultant

All valves leak, including HIPPS valve. To qualify as a HIPPS valve, it is supposed to have ZERO leak at initial installation. It may leak in service. As and when a HIPPS valve is tested as required to meet SIL3, if it leaks either a waiver (or deviation request) is obtained or the valve is repaired or replaced. API 14H and 14C limits the leakage to 400 cc/min liquid or 15 SCFM for gas? Pls check the numbers.

The tricky part is whether to take X or Y times this rate to allow for margin before the leak is deducted and valve repaired/ replaced. Most likely the smallest D type valve is sufficient. I am sorry if the response is not categorical. RM (Recall Memory) button malfunction.

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Mohammadreza Ebrahimi
Senior Process Eng. at Nargan Engineers & Constructors

Mohammadreza

Dear Kumar
1. I checked the API 14C, you mentioned leak rate is correct.
2. can i ignore this case and consequently PSV (which is considered only for HIPPS valve leak scenario) ,If maximum operating pressure at upstream of HIPPS valve is less than PSV relief pressure ?

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andriyadi

andriyadi saputra
Process Engineer at PT. Tripatra Engineering

API 14C mention 15 SCFM for gas is for Check Valve and Underwater Safety Valve.
API14H mentions it for Surface Safety Valve and Underwater Safety Valve.
I am wondering that the HIPPS valve leakage test will be conduct in similar manner with Check Valve (FSV) / SSV / USV? which are used 15 SCFM as leakage criteria.

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S M

S M Kumar
Process Design Consultant

Mohammadreza: Dangerous to comment without knowing the full system configuration and knowing what section you are trying to protect. Say there are HIPPS valve with the PSV in between them; the primary source of leak is upstream (u/s) of your first HIPPS valve; and a low pressure (LP) section is d/s of the second HIPPS valve. The leak thru the first valve will eventually leak thru the second HIPPS valve and pressurize the LP section. If this is the configuration, how are you going to protect the LP section. That should decide the PSV set point.

When you say u/s pressure is < PSV set point, under what conditions - what is the source and shut in pressure.

Look at the full system and decide.

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Mohammadreza Ebrahimi
Senior Process Eng. at Nargan Engineers & Constructors

Mohammadreza

Dear Kumar
Let me describe the case in detail.(i send you by email support document)
The process unit operating and design pressure are respectively 70 and 80 barg and this design pressure is calculated based on upstream source pressure and process unit is constructed based on above condition.
But we have to design a new feed line for this unit from another source which it’s maximum operating pressure at plant B.L is 88 barg (more than downstream unit design pressure= 80 barg). Therefore a pressure control valve is considered at Plant B.L on inlet feed line in order to decrease pressure from 88 barg to 71 barg (1 barg pressure drop is considered from control valve up to unit B.L) and at downstream of control valve HIPPS valves (preliminary No. of HIPSS valves: 3) are considered (in order to ignore control valve failure scenario for PSV load , because flare network isn’t deigned for this case) and PSV at downstream of HIPPS valves is considered for HIPPS valve leakage scenario.
New feed line design pressure is 98 barg.
Thanks
Mohammad reza

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mehrnaz hamedi likes this

S M

S M Kumar
Process Design Consultant

Dear Md Reza,

As I understand your case, a HP system at a design pressure of 98 barg and operating at 88 barg is connected to a LP system with a design pressure of 80 barg (check PSV and sweetening are 82 barg) and operating at 70 barg. A single PCV is letting down the HP gas to LP. Its fail open case can pressurize the LP system to its last shutdown or block valve wherever it is – not necessarily to the ESDV you are providing d/s of the HIPPS valve. You are providing 3 HIPPS valves and a PSV downstream of the HIPPS valves to take care of leakage thru the HIPPS valves.

Query: What is the size of new feed and the existing line to Gas Sweetening? Are they both 600# flanges.

Suggestion: Read fine print disclaimer on all such online advice

(1) From the design pressure of 82 or 98 barg it looks like both the existing and new system are 600# system capable of 98 barg. Is there any way your piping engineer can look into the existing system to the last valve without PSV protection in the Gas Sweetening system and check if it can be upgraded to 98 barg. [Note: Flange ratings cover 2” to bigger sizes. Smaller pipes can take more pressure. Good piping engineers can analyse piping and flanges if they can withstand higher pressure, do a suitable hydrotest and certify it to 98 barg. Issue closed

(2) Can you introduce 2 PCVs in series – one letting down from 88 to 80 barg; and the second from 80 to 71 barg. The first one can be SAPCV (self-actuating PCV) and the second the normal type. This will eliminate simultaneous failure of both the PCVs from a common cause and eliminate a relief scenario on account of failure one PCV. Would your client or owner buy this

(3) Don’t have HIPPS at all. Any pressure > 88 upstream of the PCV to close the u/s ESDV; any pressure > 71 barg d/s of PCV to close both u/s and d/s ESDV. Ask your Safety Engineer to perform a probability analysis. In worst case, the LP pipe will reach 88-98 barg or 1.07 to 1. 2 of its design pressure or less than hydrotest pressure. Code allows pressure excursion for infrequent events. Read piping code. 1.07 is less than 1.1 with a PSV when it relieves. If required provide a single HIPPS.

Present 3 HIPPS valves looks like an overkill considering the likely pressure excursion scenario. I will go for 1 or 2 as above.

Coming back to your original query: Can you eliminate the PSV as u/s op pressure < d/s PSV relief pressure, that is 88 < 82*1.1= 90.2. Usually upstream design pressure or PAHH set point is taken. If you consider alternative suggestions as above, you may not need the PSV.

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mehrnaz hamedi likes this

Mohammadreza Ebrahimi
Senior Process Eng. at Nargan Engineers & Constructors

Dear Kumar

Mohammadreza

PLease consiser 80 barg as the PSV set pressure and sweetening design pressure (a minor mistake in previous e-mail).

New feed line rating is 900# till the d/s of HIPPS valve and rating for the rest is 600# (same as existing pipe and gas sweetening unit).

1. For first item i am waiting to hear from piping department side to check this matter. If they confirm that pipes can suffer these high pressures PSV will be deleted otherwise this device shall be kept. There is a main point which HIPPS valve consideration is mandatory because the PSV in the feed surge drum in sweetening unit is sized only for fire case and cannot handel the load of the new scenario.

2. There are some problems for considering two series PCV as the below:

a) In the case of first PCV malfunction (Full open), the second PCV meets more pressure drop (88-71= 17 bar) than its design (80-71= 9 bar). It seems more flow will be passed through the second PCV. So downstream pipe and sweeting unit will be pressurized. As the above mentioned in part one, the PSV which is located in the sweetening unit feed surge drum is small for this case.
b) In the case of second PCV malfunction (full open), the sweeting unit will be operated in design condition and the same story for feed surge drum PSV will be occured. .

3. In my opinion the capabilities of normal operation for sweeting units in the range of 71-80 barg will be decreased base on your third configuration . In this senario if the pressure is increased more than 71 barg, the down stream ESV will be closed and feed will be intruppted. While sweeting unit has been designed to be in operateion of more than 71 barg.

I would like to draw your attention that at this stage Three valves for HIPPS are recommended from the instrument department as the perliminary design. But in future after HIPPS SIL verification study, the No. of HIPPS valves will be finalized.

Conclusion: If PSV relief pressure is less than u/s maximum operating pressure, this PSV is not necessary.

Best Regards
Mohammad reza

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S M

S M Kumar
Process Design Consultant

Dear Md Reza,

(1) Noted

(2) (a) It does not matter if the second PCV gets more pressure drop. Its function is to maintain d/s pressure. It will close or reduce its port size to suit flow and pressure drop.
(b) in the case of second PCV failure, you can reset the set point of the first PCV to 71 barg after getting the alarm. Do what you are comfortable and able to sell to your team.

(3) Sorry, I did not write clearly. I should have said existing PAHH set points instead saying >88 and >71

Conclusion: Many may not buy your conclusion of using u/s operating pressure. Usually upstream design pressure or PAHH set point is taken

Regards
Kumar

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Mohammadreza

Mohammadreza Ebrahimi
Senior Process Eng. at Nargan Engineers & Constructors

Dear Kumar

You are right. i have to consider U/S design pressure or PSHH.

API Per 14C the leakage is 15 SCFM for gas which it seems is constant, but i think leakage rate is depend on below items.

1- Valve size
2- Leakage class
3- Differential pressure between U/s (deign pressure) and PSV relief pressure.

B.R
Mohammad reza

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S M

S M Kumar
Process Design Consultant

That settles it Md Reza.

You may like to look at another issue - minimum length of a fortified section d/s of 900# HIPPS valve to be technically correct. The 900#/600 spec break location??

Let us say, due to hydrate blockage or stuck pig or a closed valve, the LP section reaches PAHH and triggers HIPPS valve to close. There is a time lapse by the time PAHH is sensed, relayed and the HIPPS valves fully close. In that time, the pressure in the u/s of blockage may build up and exceed the MAWP of the LP section. The hydrate blockage or stuck pig can occur at the 900#/600# spec break point.

Based on rate of maximum flow * time elapse, you may need a HP section of sufficient length (=volume) d/s of HIPPS valve (fortified section), instead of taking the spec break immediately at the outlet of 900# HIPPS valves.

Google search <pipeline fortified section> to read more.

In your case, you may not have hydrate or stuck pig issue and the first manual valve in the 600# line may be away giving the required fortified length.

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Sampath

Sampath Kumar R
Upstream Process Engineer at Technip

Dear Friends
Good Day:) Interesting Topic.
Extract from Shell DEP regarding the Pressure Relief Device d/s of HIPPS valve:

Quote:

A Pressure Relief Valve shall be installed downstream of the HIPPS Valves to accommodate possible HIPPS Valve leakage, unless it is demonstrated that operational response is fast enough and dependable to prevent overpressuring that could result from this leakage. The minimum effective diameter shall be 25 mm.
Unquote:

From the above, minimum effective diameter shall be 25 mm which is equivalent to the area of 0.76 inch2. This will call for "H" orifice size.

Kind Regards
Sampath Kumar R

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S M

S M Kumar
Process Design Consultant

Thanks Sampath

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Padmanabha

Padmanabha NC
Lead / Senior Process & Flow Assurance Engineer at PT. Meindo Elang Indah

Hi Sampat,

i prefer to take the valve leakage rate from HIPPS vendor ...the API 14 C values are for FSV......and i have seen some PSV sizing for HIPPS valve leakage on de-rating of piping from 600 # to 300# the pipe header size 28 " design Gas flow is 1004 MMSCFD operating pressure is 41 barg ....the installed PSV orifice size is 1E2 ...i have P&IDs for this ..but no sizing details of PSV .....if required i will share....

thanks
Padmanabha

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Mohammadreza

Mohammadreza Ebrahimi
Senior Process Eng. at Nargan Engineers & Constructors

Dear Padmanbha

Would you please send me P& ID? (mohammadreza.ebrahimi@yahoo.com)

Thanks
Mohammad reza

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Sampath

Sampath Kumar R
Upstream Process Engineer at Technip

It's my pleasure Kumar Sir:)
Dear Padmanabha.
Regarding the relief device size d/s of the HIPPS valve, the quote I have mentioned should apply only if you are following Shell DEP. Otherwise it shall be based on the general engineering practice or based on the leakage rate from HIPSS valve vendor.

Btw I missed to provide the Shell DEP number. Here it is:

32.80.10.10-Gen. Clause: A1.7 (Page No:81)

Kind Regards
Sampath Kumar R

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