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Heat Exchanger Tube Rupture

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S M Kumar
Process Design Consultant

QUOTE:
We have a high pressure gas cooler. Gas side is designed at XX barg; Cooling Water side is designed to 10/13XX to its isolation valve and rest of the cooling water side is 0.40XX. In case of tube rupture we consider the following:

1. If HX cooling water outlet isolation valves are closed. LP side to isolation valves is designed for 10/13 rule. No PSV required for tube rupture case.
2. If HX cooling water outlet isolation valves are open. In this case vapor will relieve to cooling water side. Hence overpressure will not occur. No PSV required for tube rupture case.
3. If HX cooling water outlet isolation valves are open but LP side is isolated or blocked somewhere. This case could be considered as shutdown case and shutdown instances are supervisory controlled . Hence this case should be dealt by operating procedures. No PSV required for tube rupture case.

I have seen similar design in many such LP cooling water side without PSVs. Questions I have

1. Is my understanding correct , if not please add or correct me.
2. Have you seen other instance where tube rupture PSV are provided in cooling water side hence it can protect whole cooling water network
UNQUOTE

My Response:
Your cooling water side d/s of isolation is lower rated and is not protected. If you perform dynamic simulation analysis, you will notice that on tube rupture, it is difficult for the gas to push the massive cooling water in the LP side at the speed or velocity with which gas is issuing out of the rupture. Inertia. Like ant pushing an elephant. There will be pressure spikes and in plants I have designed, I have protected the LP side with 1+1 Rupture Disks. PSVs do not respond fast enough. Usually you need 2 RDs in series if the flare system pressure is likely to vary so that the first RD sees a constant backressure. Add a PI/PAH in between the RDs to protect against pinhole leaks in RD

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Amit Pawar likes this
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Mukund

Mukund Chiplunkar
Experienced Process & Technical Safety Consultant at SOHO

Kumar,
I have also seen RDs on low pressure rated cooling water systems. At that point in time, I thought it was an overkill. However, rapid overpressurisation does occur depending upon the size of leak, etc., and is verifiable by CFD analysis.
By the way, I have been told that API has removed that clause which lead to the interpretation of 10/13 rule that you have described. It is now stated that whether overpressure protection is required or not, is to be decided by the designer on case by case basis.

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ASHOK

ASHOK KUMAR
Process Engineer at MODEC International INC

Dear Sir

Though RD is installed , the low pressure side of the Heat Exchanger has to be designed for the higher transient overpressure Pressure,isn't it?

The Low pressure side (till the isolation valve as stated above) shall be designed for the Transient overpressure.

Please correct me.

Thanks and Regards

Ashokkumar

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S M

S M Kumar

Process Design Consultant

Good point Ashok. The CFD transient pressure is valid only if there is no relief device. As RDs are supposed to open in 2-5 milliseconds, so far I have designed the LP side to the set pressure of RD, that is LP System design pressure.

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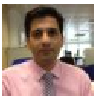

Saeid Rahimi Mofrad

Senior Specialty Process Engineer at Fluor

Before posting any answer, I would like to bring to your attention that there is a wide variety of tube rupture cases and the fact that each case may need different pressure relief device, design precautions and method for tube rupture flow and relief rate calculations.

First of all, the location of LP and HP fluids in heat exchanger is important because the capability of tube for handling extra fluid from shell side is totally different from shell side. For example, a HP liquid in tube which is released to LP gas in shell, most probably no pressure protection on the heat exchanger is needed as the liquid falls to the bottom of shell and is drained to the downstream vessel. Relief device is not even needed on the downstream vessel if the extra liquid can be handled by the interconnecting piping to downstream vessel and it does not cause any overpressure in that vessel. In the same exchanger, the situation would have been totally different, if the HP Liquid was in shell side.

Secondly, the number of phases in each side of heat exchanger is very important. See the following table which summarizes different combinations:

LP side---------------------------------- HP side

Gas-- Gas
Gas-- Liquid
Gas-- Two phase

Liquid----------------------------------- Gas
Liquid ----------------------------------- Liquid
Liquid ----------------------------------- Two phase

Two phase/flashing liquid ------------ Gas
Two phase/flashing liquid ------------ Liquid
Two phase/flashing liquid ------------ Two phase

This is important because the tube rupture flow rate calculation depends on the LP side fluid phase. The formula used for tube rupture calculation of gas or liquid or two phase flow or flashing liquid are different . Furthermore, depending on the type of fluid in LP side, the required relief device can be different. For example when HP gas is released into the LP liquid, a fast acting relief device is installed on the LP side.

Finally, the required relief rate can be different form the tube rupture flow rate (this is a mistake that many do – taking the results of tube rupture flow calculation directly to relief valve sizing). For HP liquid into LP gas case, thought what is passed through the ruptured tube is liquid, what is required to be released by relief device is the corresponding volumetric flow of gas. Furthermore, sometimes what is required to be released by relief device is not the fluid at HP side.
For example, in a gas-gas exchanger the HP gas is mixed with LP gas after tube rupture and therefore the composition and physical properties of stream which reaches relief valve can be totally different than HP gas. Therefore, the mixing and thermal effect of fluid at LP side (cooling due to pressure reduction/ heating as result of contacting the hot fluid at LP side / flashing because of both reasons) on the HP fluid should be considered and the resultant fluid (which can have different phase, composition or temperature) shall be used for relief device sizing.

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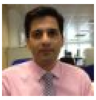

Saeid Rahimi Mofrad
Senior Specialty Process Engineer at Fluor

API 521, 5th Edition states that “The tube rupture scenario can be mitigated by increasing the design pressure of the low-pressure exchanger side (including upstream and downstream systems), AND/OR assuring that an open flow path can pass the tube rupture flow without exceeding the stipulated pressure, AND/OR providing pressure relief.”

It adds up that “It may be necessary to locate the relieving device either directly on the exchanger or immediately adjacent on the connected piping. This is especially important if the low-pressure side of the exchanger is liquid full. In this case, the time interval in which the shock wave is transmitted to the relieving device from the point of the tube failure increases if the device is located remotely. In addition, there is a time delay for the gas to overcome the momentum of the liquid-filled low-pressure side prior to establishing a full flow through the relief path. This can result in higher transient overpressure on the exchangers before operation of the rupture disk or relief valve.”

Increasing the design pressure of LP side to 10/13 of HP side design pressure is effective protection when it is applied on the entire LP side not the heat exchanger only and it ensures that LP side can confine the HP side pressure. But in some specific applications where HP gas/two phase flow/flashing liquid ruptures into LP liquid further precautions are needed further to increasing the design pressure of LP side. That is why API stat□s that “It can be impractical to protect some heat exchangers (and associated piping) by relief devices alone e.g. if there is a high pressure difference between the shell and tube sides. In these cases, different layers of protection, such as improved metallurgy, more frequent inspection and increasing the design pressure of the low pressure side (including upstream and downstream piping until the pressure is dissipated), can be necessary.”

In fact, there are more to do than what API standard mentions for this case, ranged from suitable heat exchanger selection and design considerations (to prevent the rupture or minimize the possibility of it) to process and relief system proper piping design (to safely handle the consequences of tube rupture).

For shell an tube heat exchanger in this specific application (HP gas/two phase flow/flashing liquid into LP liquid), two (2x100%) fast acting relief devices (rupture disc or buckling pin device) has been recommended in the recent studies. If LP side is tube, one RD on the inlet line and the other one on the outlet. If LP side is shell, one RD on the one side of the shell and the other one on the other side (especially for long shells).

2. Even if you prove that there is an open path for the gas to be relieved through low pressure side (in this case water basin or sea), you have to assess the possibility of water contamination with hydrocarbon, loss of products, hydrocarbon emission, environmental impact of releasing the gas in the cooling tower basin or see water return point, and the design of pipe support and anchors on the line handling slug of liquid which will be accelerated in front of the gas as it expands.

3. I don’t agree with this statement.

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Azhar Ali likes this

Ashraf

Ashraf Abufaris
Process Engineer at Petrofac

Saied,
I got a question on your post. If we asset the possibility of hydrocarbon contamination of cooling water network, what would be the solution ? We are recently designing cooling water network with the same configuration you described with hydrocarbon analyzer just upstream the sea water outfall structure but the path is open to sea.

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