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Chidi Benjamin Dibiat

Abstract

Thos masters thesis work tries to examine the current procedures used for testing pressure safety valves, and the bench mark used that are used in defining an inspection and testing interval for pressure safety valves. I have started by describing some basic elements of the design of safety valves, then go on to look at how these valves are tested. I have also examined the current criteria most owners of the pressure safety valves use in setting up their maintenance programs. The aim of this thesis in the end is to try and modify the current intervals being used in the oil and chemical industries today so as to cut down unwanted cost, guaranty the safety of personnel, and safe guard against any form or accidents in the plant. This thesis shows a conservative approach that is also in line with approved standards that can be used for setting an optimum inspection and testing interval.

Table of Contents

Chapter 1

Introduction

1.1 Background and Problem

There is a general problem in the oil and gas industry, as well as in the chemical industry, on how long a Pressure Safety Valve (PSV) should be allowed to be in operation before it is inspected and tested. This problem was not seen as a serious issue before now, but with the recent global financial crises and some very costly accidents, many firms dealing with PSVs have come to realise the need to cut down on unnecessary expenses by way of eradicating unnecessary testing and inspections, while still maintaining the integrity of the PSVs.

1.2 Scope and Objectives

IKM Laboratorium AS is a subsidiary of IKM Gruppen, Norway that specialize in testing and calibration of instruments. The range of instruments that IKM Laboratorium handle are classified as either pressure, electrical, temperature or mechanical instruments. In the pressure department, IKM Laboratorium has contracts with several clients to carry out pressure tests on their Pressure Safety Valves (PSVs). At the moment, these PSVs are just being tested at random times depending on the time frame their clients feel is safe and economical to them.

This thesis looks at a holistic view of pressure vessels and then tries to narrow the main problems to that of a PSV. I have attempted to understand how a PSV looks like and how it operates under normal operating conditions. I have gone through all the test results and test procedures at IKM Laboratorium AS, as well as apply the basic standards relating to pressure vessels and valves.

The objective of this work is to determine an optimum inspection and testing interval for PSVs that will also preserve the functional integrity of the PSVs. This would help the clients of IKM Laboratorium AS to plan their PSV maintenance programmes much better, as well as reduce operating cost of the PSVs. The long term advantages are: reduced expenses for the owners of the PSVs, better industrial services between the owners and calibration companies, and safer plants with highly reduce Fatal Accident Rates (FAR).

1.3 Methodology

This thesis is based on literatures and document reviews from IKM Laboratorium AS, and also various international standards such as the API, ASME and ISO standards. Also, various journals and articles, and information form the World Wide Web have being used for this thesis. Interviews with the test operators at IKM Laboratorium and the owners, were also organised and held in other to get more information about Pressure Safety Valves and other related areas for this project work.

1.4 Limitations

In this thesis, no visit to a site where PSVs were being utilized was considered due to logistics and other related problems. Most calibration companies only perform pressure tests on PSVs without doing a leak test. A combination of pressure test results and leak test results would have added more credence to this work. Also, the best pressure test for a PSV is that which subjects the PSV to the full operating conditions that it is supposed to endure. However, this would be very impractical and impossible to reproduce; hence I have confirmed myself to the results received from a bench-test done in a controlled laboratory at IKM Laboratorium AS. Finally, to be able to set a good inspection interval and test interval, a risk ranking method would ideally be the best measure for achieving this, and this would have meant conducting a proper consequence analysis on a sampled set of PSVs. This has not being done in this work due to lack of genuine data for a consequence evaluation.

Chapter 2

Pressure Safety Valves

2.1 Description of Pressure Safety Valves

Pressure Safety Valves (sometimes called Pressure relief valves or PSV, pressure relief devices, PRDs or simply safety valves) in the oil and chemical industry have existed for over four centuries now. The PSV is primarily used in protecting life and properties. It is a mechanical valve that is designed to open when a certain pressure value is exceeded in a process pressure system. This action helps protect life and all investments that have being put into such process plants.

The PSV is able to perform this function by acting as a path of least resistance in the event that the system pressure exceeds the set pressure of the PSV. This would allow a portion of the fluid to be diverted through an auxiliary route (usually a piping system called *flare header* or *relief header*) connected to a flaring system. As the fluid is being diverted, the pressure within the pressure system drops. When the pressure drops below the valves reseating pressure, the valves closes.

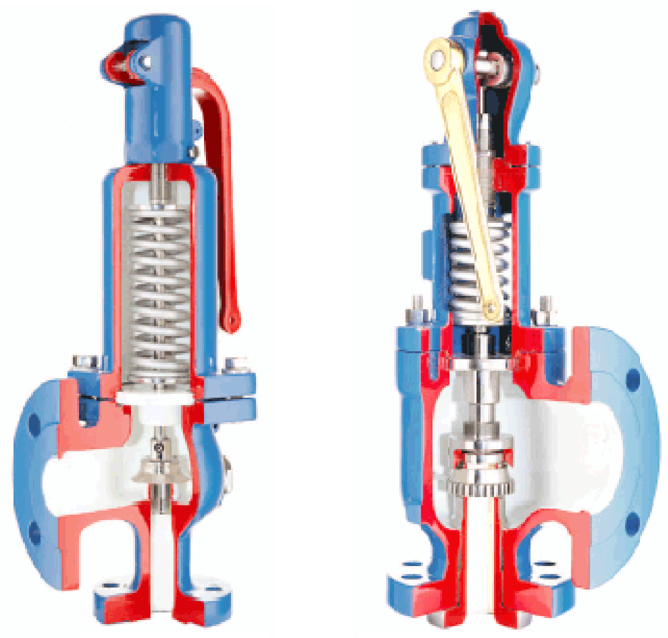

Fig. 1 Typical Safety Valves

2.2 Brief History of PSVs

Many people believe Papin was the inventor of safety valves, when he applied it in 1682 on his digester experiment in France. Papin kept his safety valve in place by means of a lever and a movable weight, sliding along the lever, thus allowing him to regulate steam pressure. It later turned out that Papin only managed to make improvements on an already existing system that was in use 50 years earlier by the German Glauber. Glauber attempted to prevent retorts and stills from bursting from an excessive pressure by using a conical valve which was fitted and loaded with a cap of lead. Many other scientist worked on the Glauber principle and many different designs were formed later on [3].

At the beginning of the 20th century, there were numerous boiler explosions in the United States. This prompted the government to look for a solution to these accidents, and they asked the American Society of Mechanical Engineers to formulate a design code. The boiler & pressure vessel committee was formed as a result of this, and the A.S.M.E. Section I for fired vessels was formulated and was made mandatory for all states in the United States. As the process industry grew larger, there was need for protection of life and property, and the need for unfired pressure vessels was identified. This gave rise to the A.S.M.E. section VIII. Other standards like the API standards and ISO standards were developed for safety valves. Also, in other to allow for free circulation of goods in Europe, member states had to conform to the pressure equipment directive (PED), which was published in 1997.

2.3 Safety Valve Design

We shall only consider the basic spring loaded pressure safety valves, also called conventional or standard valves shown in fig. 2

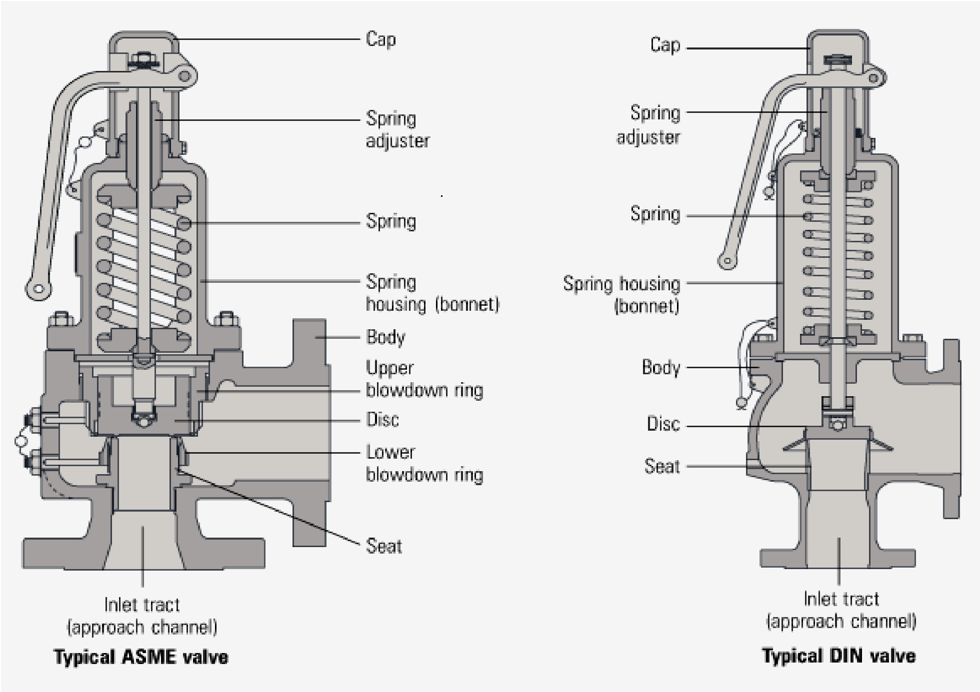

Fig. 2 Typical designs for Pressure Safety Valves

The basic elements of the designs shown above comprise a right angle pattern valve body with a valve inlet connection which is mounted on a pressure-containing system. The outlet connection is flanged for connection to a piped discharge system, or vented directly to the atmosphere for cases involving compressed air.

The valve inlet design can either be a full nozzle or a semi-nozzle type. Full nozzles are used in safety valves designed for process and high pressure applications, and for corrosive fluids. Semi-nozzle designs would normally have a seating ring fitted into the body that gives the seat of the valve, thus allowing for easy replacement of the seat without replacing the entire inlet.

The disc is held against the nozzle seat by a spring that is housed by a body called the bonnet. The discs used in rapid opening safety valves are usually surrounded by a disc holder or huddling chamber that helps to produce a very sharp response. The closing force on the disc is provided by the spring. This spring is made from carbon steel. The compression force on the spring can be adjusted by using the spring adjuster shown in the diagram above. This will help change the pressure at which the disc is lifted off its seat.

Normally the design principles of the conventional safety valves are similar, but the design details could vary considerably. In general, the DIN style valves which are common in Europe tend to use a simpler design with a fixed skirt arrangement while the ASME style valves have a more complex design that comes with one or two adjustable blow down rings.

2.4 Codes and Standards

Standards relevant to safety valves vary from country to country, and many are sections within codes relevant to Boilers or pressure containing vessels. The American Petroleum Institute (API) has developed the most commonly applied standards for the petroleum and chemical industries. API 521 provides excellent guidance for evaluating causes of over pressure and pressure relief systems. API 520 is the design manual that is used for the design, sizing and selection of component. API 526 provides a standard for manufacturers of flanged PSVs, and contains a set of installation dimensions, pressure and temperature ratings, set pressure limits, capacities and materials. API 527 provides a basis for the testing and acceptance for set pressure and seats tightness of PSVs. API 510 and API RBI 581 provide a guide for establishing inspection and testing intervals of safety valves or relief devices.

The NORSOK standard I-001 (Field Instrumentation) states the functional requirements and installation processes needed for various field instruments which includes the pressure safety valves. The NORSOK standard P-001 (Process Design) establishes the requirements for testing of pressure safety valves utilised in the Norwegian based industries.

Also the ISO 4126 standards (sections 1-7) are referred to by many manufacturers when they need to determine valve discharge coefficients.

These set of standards ensures that valves from various manufacturers are interchangeable both functionally and dimensionally. The API standards have being used extensively in this work because it is widely used by most companies that deal with pressure safety valves.

2.5 Inspection, Testing and Maintenance of PSVs

The PSV has no instrumentation or indicators that can give an indication of its status at any given time. This makes it very hard to carry out any form of condition monitoring processes on them. Hence there is need to be able to draw up a suitable inspection and maintenance scheme that would ensure the PSV operates properly when they are called upon at times of emergency. There are several guidelines that are utilized in recommending an effective inspection and maintenance program for pressure safety valves. Like I have stated earlier on, the API 510 and API 581 are the Pressure Vessel Inspection Code which I have utilised in this project extensively. This is because these are the only standards that have established any methodology for calculating test and inspection intervals from a test result that have being obtained from any field operation.

In other to establish a reliable maintenance program for PSVs, there is need to test the PSVs as often as the reliability of the PSV can be guarantied. The most desirable and common tests performed on PSVs is that which subjects the safety valve to full operating conditions which such valves are expected to endure like the set pressure, lift and blow-down acceptance. I shall discuss the testing procedures and protocols later on in this project work, and try to establish an optimum testing interval for PSVs which is the ultimate objective of this project.

Chapter 3

PSV Test Procedure and Equipment

3.1 Scope and Testing Procedure

The procedure outlined here in this work is basically the same for both field test and laboratory test work.

The calibration covers both calibration of pressure safety valves irrespective of the operating medium (liquid, vapour or gas). The procedure covers the calibration of pressure safety valves where the operating medium is liquid, vapour, gas or air, or combinations of any of these. The procedure requires that the test object be disconnected from its original system and mounted on a test bench or similar arrangement that could provide the basic safety and operational requirement for operating personnel.

3.2 Test Apparatus and Instrument Set-up

The test apparatus comprises a piston, pressure indicator, a barometer, and a gas pressure reference. The set up is as shown below.

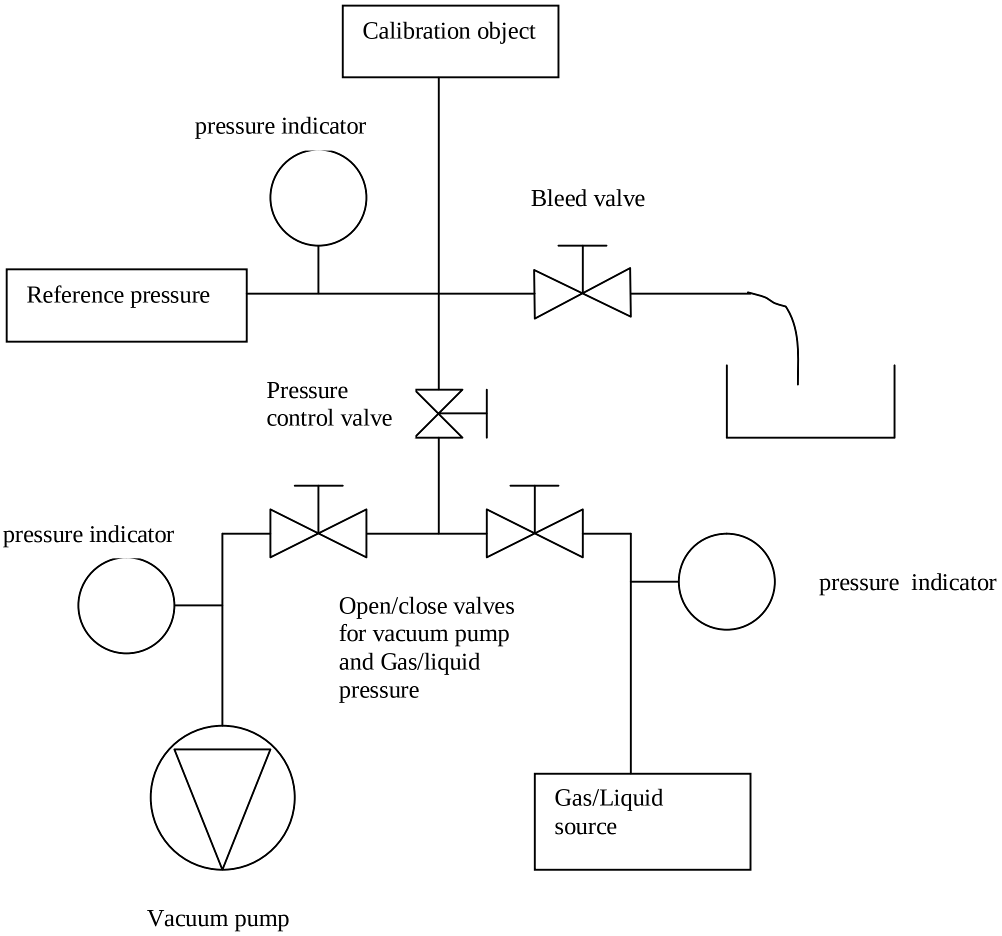

Fig. 3 set up diagram

3.3 Placement and Set Point of PSV

The first step in the placement stage is to identify the inlet and outlet sides by using the markings on the valve, or the manufacturer's assembly instructions. Apart from those cases where it is stated, the valves are always placed in a vertical position with the inlet down. Valve outlet is directed so that excess pressure is released in a responsible and safe manner. The test should take place in an environment without striking external influence factors such as vibrations, pollution, very large temperature fluctuations and so on.

The Cold Set Point (CSP) is the pressure needed to trigger 'ON' the safety valve when it is tested in the bench test rig. The CSP is calculated by correcting for the original system baked jerk and temperature.

3.4 Visual Inspection and Functional Tests

It is very important that the safety valve is checked and inspected visually for defects and deformities that would normally affect the valve functionality.

The test object is then connected to the pressure reference on the test bench, and all connections used must have appropriate transitions with current pressure ratings. A vacuum pump must be used if the object is to be tested and calibrated for pressure ranges lower than atmospheric pressure. The valve should be prepared in accordance with the valve Standards and user manuals.

The Functional Test involves generating a pressure at a controlled steady pace into the safety valve until it opens. Care must me taken with valves that come in a metal-metal sealing so that they do not "pop" from the application of excessive pressure otherwise there could be 'knocking' and this may damage the valves.

It is nice to check the valves and connections for any leaks during the first pressure rise as pressure is applied to the valves. If the safety valve does not open at the intended set point, it should never be exposed to pressures beyond the valve pressure range stated in the manufacturers' data sheets. When this data sheet is unavailable, then the valve is only subjected to a pressure limited to 110% of CSP.

Note: A safety valve that has not being triggered for a long time may have a slightly higher set point than the label on the valve might suggest when being pressurised initially.

Incorrect valve lift during testing according to the above mentioned conditions would result in the termination of the functionality test, and repair data could then be obtained according to the instruction manuals.

3.5 Adjustment of PSVs.

Normally a leak test should be performed after a visual inspection to see if the sealing lips of the valves are intact, however a leak test is beyond the scope of this project hence we would try to adjust the PSVs without considering the results of a leak test.

When there is a large deviation from the allowed pressure tolerance range, the valves should be adjusted in accordance with the product service handbook or manual.

Safety valves with spring set are adjusted normally ‘IN’, to facilitate or increase the load on the feathers of the valves. During this process, it is important that the sealing is held completely still to avoid damaging them. New set points are established and then checked against the referenced pressure values, and if deviation is still too high, the adjustment is continued until the deviation is eliminated or within acceptable limits, otherwise the valves should be subjected to a pressure test.

3.6 Measurement procedure and Repeatability

The tests under gauge pressure conditions are done with atmospheric pressure as the reference, and the tests under absolute pressure conditions are done with vacuum pressure as reference. Both cases require that the reference be started according to its user manual. The following procedures are usually performed.

1. The minimum pressure reading and initial calibration are checked.
2. Pressure is then generated at a controlled pace into the safety valve until it opens. This is the set point
3. The set point is recorded.
4. The safety valve will reset itself when the pressure is reduced. The pressure at which the valve stabilizes is then noted and recorded.
5. The safety valve is then depressurised.

After the approved test has being performed, the company’s brand stickers or labels are applied to the valves and sealed. The labels would normally contain the serial numbers, set pressure, date of testing or calibration, proof number.

3.7 Discussions and Interviews with the clients on current practice

I had a few meetings with representatives from some of the owners of the safety valves being tested at IKM Laboratorium AS. Most clients claimed that a Risk Management team usually performed a Failure Mode Effect and Criticality Analysis (FMECA) or just a criticality analysis on the PSVs for various locations, and the results are classed according to in-service time period, operating pressure to set Pressure ratio (OP/SP), Temperature of operating environment, the condition of different service locations or process units, and the normal pipe size of the PSV at the inlet (inlet size). From the results of the FMECA, these clients try to fix an appropriate testing interval for various locations according to the risk results or criticality of the location. One client attested to the fact that the best way to go about finding an appropriate inspection and testing interval would be to perform a risk assessment on the PSVs. However, he made it clear to me that many plant managers are not willing to abide to the results of such assessments as many of them do not see the need to shut down their plants so often as dictated by the tests results. For this reason many plant managers stick to the time frame suggested by the API 510 standard.

Most of the clients depend on the as test results in the workshop to evaluate the aging condition of the PSVs. This is done by taking the ratio of the Test Pressure (TP) to the set pressure (SP). They think an increased in the value of the ratio TP/SP is a very good health indicator and this is very good engineering practice for inspection and maintenance. Hence, the PSVs are removed from their static equipment and a pressure test is normally conducted. How the pressure test results are used in trending the aging of the PSVs would be looked at in the next chapter.

However, it is important to bear in mind that PSVs are standby emergency devices that must function when called upon in the event of a pressure build-up. Its is very important that the test intervals set aside are within acceptable limits that guaranties that the PSVs are going to be functional at all times. Considering that this work is based on limited data, and given the challenges faced, I have adopted a very conservative approach, and my recommended inspection and testing interval suggested later on in this thesis is subject to change as more genuine and helpful data for performing a leak analysis and consequence evaluation becomes available.

Chapter 4

Determination of testing and inspection interval

In this section, I have tried to establish a very good inspection program for PSVs, using a Risk Based Inspection Assessment (RBI assessment), and also tried to analyse the state of the PSVs using a Corrosion Rate method and Remaining Life Calculation as required by the API 510 standard. The RBI assessment method would be used to estimate the inspection intervals of the PSVs, while the corrosion rate and remaining life methods would help add more credence to the RBI assessment results. The Pressure Test results would help ascertain the again pattern of the PSV by studying the OP/SP ratios of the PSVs. From the risk analysis and pressure test results, I would attempt to give a safe estimation of an appropriate inspection and testing interval. It is also possible to trend the testing intervals and how this is done would be treated in full details as well.

4.1 Risk based Inspection

"The increasing pace of mechanisation and automation and increased focuses on quality and availability of plants, factories, and systems, has created a need for new management techniques in field maintenance engineering."(Uday Kumar, 2002). The aim of any maintenance strategy is to reduce business risk, hence there is the need to incorporate a formal reliability and risk assessment into any system design process which includes the operational stage of any equipment.

RBI assessment uses risk analysis of the results of inspections, testing and monitoring of the PSV. In this work, Risk has being defined as the product of the likelihood of failure (LOF) and the consequence of failure (COF). Hence the risk value of each PSV based on a well performed RBI assessment would be mathematically written as:

$$\text{Risk} = \text{Likelihood of failure (LOF)} \times \text{Consequence of failure (COF)}$$

So the most important aspect of this assessment is the recognition of the LOF and COF associated with PSVs because the RBI methodology depends on both a probability and a consequence evaluation.

The latest version of the API 510 allows the use of RBI assessment in setting intervals for pressure safety valves as long as the safety valves are tested at intervals that are frequent enough to verify that valves function reliably, and that the intervals are governed by the performance of the devices in a given service location.

There are two ways in which this can be done. The first approach would be to find the values of the **LOF** and **COF** values and then calculate the risk value associated with a given valve [3]. The risk results can then be used to rank the risk levels associated with a particular group of safety valves, and subsequently an inspection and testing interval is assigned to each group. The other method involves using the **API RBI 581** approach which involves doing a risk evaluation for a group of safety valves using plant data or the default values stated in the API 581 standard. I would explain both methods in full and then use the API RBI approach in estimating the inspection and testing intervals of safety valves.

4.1.1 First Approach

This method was developed by Chein et al (2009) and it involves calculating the values of the likelihood of failure and a consequence of failure, which can then be used in calculating the risk value for each group of PSVs. I have made a few modifications to the original approach so that it suits the problem being addressed in this project.

4.1.1.1 Likelihood of Failure (LOF)

There are some parameters that affect the LOF apart from the usual PSV parameters, and they have being set according to their discharging capacities. This is because factors like the process operating conditions could also influence the health and aging of the PSVs. Hence I have divided the LOF into two assessment groups namely the likelihood factor ($f_{likelihood}$) and the generic failure condition (f_{gfc}) based on the PSV parameters and the process operating conditions. Each of the above mentioned assessment group should then subdivided into various sub-likelihood factors A_i (for PSV parameters like fluid category, service duration, operating temperature and so on), and also a sub generic failure condition B_i (representing the actual operating conditions like light lifting or heavy lifting, frequency of use of a given section of a plant, etc). It also important that various weighing factors be used for each PSV, since they are normally utilized at different locations and in different service areas. Hence I have brought in the weighing factors W_{Ai} and W_{Bi}, which correspond to the weighing factors for A_i and B_i respectively. The as-test results which should normally include

I. The various fluid categories, inlet size of PSVs; Service duration, operating temperature; OP/SP, Different process.

II. The type of lifting (Heavy or light), leaking or fouling, and any other reason that could lead to the failure of the PSVs, and the results should be documented, collated and used for the LOF calculations as follows:

$$f_{likelihood} = \Sigma W_{Ai} A_i \text{ ……………………………1}$$

$$f_{gfc} = \Sigma W_{Bi} B_i \text{ …………………………………..2}$$

Then LOF value for each PSV can then be calculated as the product of $f_{likelihood}$ and f_{gfc}, i.e.

$$LOF = f_{likelihood} \times f_{gfc} \text{……………………………3}$$

4.1.1.2 Consequence of Failure (COF)

COF determination would be very difficult to conduct fully in this work as it requires a lot of complex data which can not be covered in this work, and would generally require a risk process specialist to get all the data required. This involves a very rigorous analysis that considers likely and historical demand rate on each PSV. Also, COF determination requires specific engineering inputs like original design basis, the likely extent of over pressures, flammability and toxicity, records of management change, and many other factors. The index of toxicity, health and environmental hazards of pressure vessel where the PSVs are mounted have being used in this work based on the assumption that PSV failure would lead to the failure of the Pressure Vessel. This is a very simplified COF analysis and if better results are to be obtained, a specialist must be called in to support the condition of the pressure vessel.

Based on a report prepared by C.H. Chien et al (2009), the COF for PSVs can be calculated using the equation:

$$COF = \Sigma C_i = Index_{toxic} + Index_{health} + Index_{environment} \text{………………………..4}$$

Where C_i is the index of toxic/health/environment provided by a process specialist.

Using the research structure of risk based inspection developed by Chien et al; an appropriate inspection interval for PSVs can be established. This is shown below in fig. 4 below.

Due to time constraints and lack of complete data, no serious COF analysis has being done in this work, neither was a specialist contacted to provide accurate and appropriate data for this RBI assessment to be complete. Chien et al with the help of a risk specialist were able to develop a risk ranking system which they used in estimating an inspection interval for safety valves. This was based of personnel population density, location of process plant (distance from a high density community), corrosive properties of the fluids, risk of plant loss, and risk of environmental pollution. The algorithm for this approach is shown in figure 4.

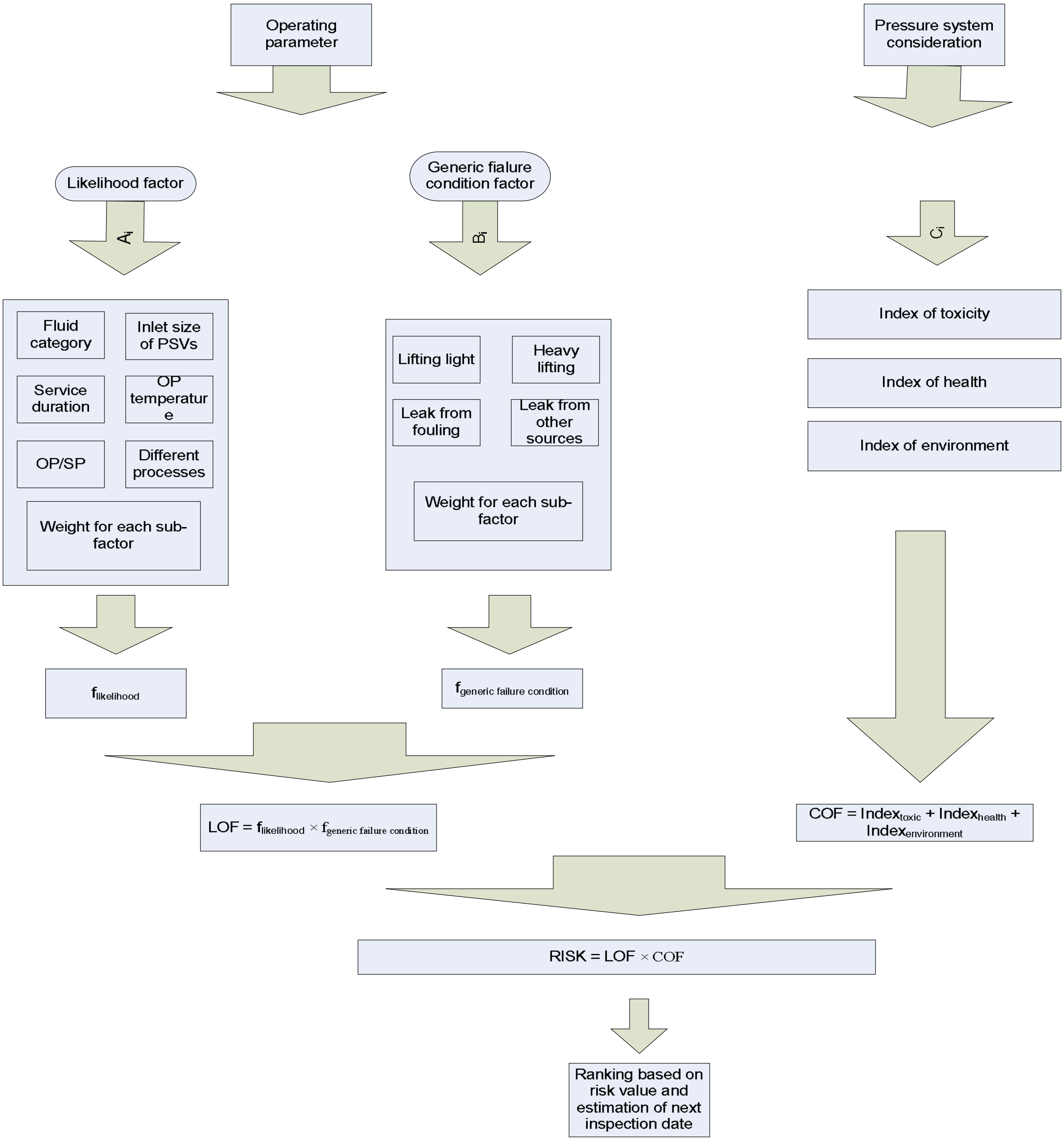

Fig. 4 Research structure of risk based PSVs inspections.

When the LOF and COF values have being determined, the expected risk values for various service conditions can then be determined, and a risk matrix can be established showing different risk categories. Once the risk categories have being set aside, an appropriate inspection interval can be set for each of these risk categories.

However, to be able to set a good testing interval, a more rigorous and statistically based risk evaluation must be done in accordance with API RBI 581 standard. This could either be a qualitative, quantitative or semi-quantitative analysis. How this is done would be described shortly.

4.1.2 API RBI 581 Risk-based inspection Technology

This approach is basically based on the use of a demand rate for the device in combination with a probability of failure on demand that is determined from the plant specific data if it is available; in this work a default data has being used. The data inputs are used to establish the probability of failure as a function of time (Weibull or Exponential approach). A consequence evaluation is also required here and it is done to include overpressure demand scenarios, amount of expected overpressures upon PSV failure and the added consequences associated with device leakage (see section 7 of the API RBI 581 standard). Leakage evaluation has not being considered in this work due to lack of leakage data.

The consequence evaluation is normally performed for the protected equipment on which the PSVs are mounted, and to be able to carry out a good consequence analysis, it is paramount that the various failure modes are identified.

The failure modes of any significance needed in evaluating the risks associated with PSVs identified in the API RBI 581 are:

1. Failure to Open (FAIL)
    a. Stuck or fails to open (FTO)
    b. Valve partially opens (VPO)
    c. Opens above set pressure (OASP)
2. Leakage Failure (LEAK)
    a. Leakage past valve (LPV)
    b. Spurious/premature opening (SPO)
    c. Valve stuck Open (VSO)

4.1.2.1 Use of a Weibull Curve

To be able to predict good test intervals, it is very important to find a good way to trend historical data, and one such way it to get a time dependent curve. Expressing the probability of failure as a function of time is one way to get around this bottle neck. The Weibull distribution curve is the most ideal for this situation as it can be used for very large data set or population. The cumulative failure density function (otherwise known as unreliability) for a Weibull distribution is written as

$$F(t) = 1 - R(t) = 1 - \exp\left[-\left(\frac{t}{h}\right)^{b}\right] \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 5$$

where

F(t) = unreliability

R(t) = reliability

t = time interval

η = Weibull parameter characteristic life. It is equivalent to mean time to failure (MTTF)

β = shape factor.

When β=1, then η becomes the mean time between failures (MBTF) for an exponential distribution.

Adjustments made to η parameter to increase or decrease the probability of failure on demand can be seen as an adjustment on the MTTF for the PSVs. Also, all PSVs are assumed to have similar Probabilities of failure on demand, POF, if they have similar services. Thus industrial failure rate data can be used in the determination of a default probability data. How the probabilities are calculated would be discussed next.

4.1.2.2 Calculation of probability of failure (POF)

The probability of failure to open for the safety valve is defined as the product of probability of the safety valve failing to open on demand (POFOD), the demand rate (DR) and the probability of failure leading to loss of containment (PF). We shall see how each of these constituents or multipliers is obtained.

That is,

$$\text{POF} = \text{POFOD} \times \text{DR} \times \text{PF} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 6$$

PF is both a function of time and potential overpressure, and API RBI recognises that there would be an increase in probability of loss of containment from protected equipment due to elevated overpressures.

4.1.2.2.1 Demand Rate calculation

The demand rate DR for a safety valve is defined as the product of the initiating event frequency, EF, and the demand rate reduction factor, DRRF.

$$DR = EF \times DRRF \quad \ldots\ldots 7$$

API RBI provides estimates for EF of safety valves based on given relief overpressure demand case that the device is providing protection for. See Appendix 3 for a background on the default EF provided.

Also, the safety valves actual demand rate would most likely be less than the initiating event frequency. Hence the need to use a reduction factor becomes necessary. This is the demand rate reduction factor DRRF used in the above equation.

Since safety valves protect equipment with different overpressure demand cases with their unique demand rate, then a total demand rate is evaluated as:

$$DR_{total} = \sum_{j=1}^{ndc} DR_j \quad \ldots\ldots 8$$

4.1.2.2.2 Probability of Failure on Demand

Once the demand rate has being determined, the next step would be to find the probability of failure on demand POFOD. API RBI already provides us with default failure on demand failure rates developed from industrial data that are expressed as default Weibull curves and modified according to the following procedures.

1. Determine default Weibull parameters, β and η_{def}, based on capacity of service severity and type of safety valve. By severity, mean that different fluid categories and temperature effects on safety valves must be considered. Normally, three severity categories of MILD, MODERATE, and SEVERE, are used with the default Weibull cumulative distribution curves. How these different severity parameters are applied is seen in Table 4 of the Appendix section of this work.

2. Different adjustment factors namely; F_c, F_{op}, and F_{env} which are the adjustment factors for convectional valves, for overpressures above 1.3, and for environment factors respectively, are applied to the characteristic life to obtain a modified characteristic life η_{mod} given as

$$\eta_{mod} = F_c \cdot F_{op} \cdot F_{env} \cdot \eta_{def} \quad \ldots\ldots 9$$

The modified characteristic life is then updated on the safety valve inspection history, and used to calculate the probability of failure on demand for a specific safety valve as

$$POFOD = 1 - \exp\left[-\left(\frac{t}{h_{upd}} \right)^{b} \right] \quad \ldots\ldots 10$$

Since a trending analysis is going to be done as well, it is very paramount that the inspection data are updated constantly as more data are collected. A Bayesian approach is used. The adjustment procedure stipulated in API RBI is to start of with a prior probability of failure of failure $P_{f\text{-}prior}$ given as

$$P_{f-prior} = 1 - \exp\left[-\left(\frac{t}{h_{\text{mod}}}\right)^{b}\right] \text{………………………}11$$

Thus the prior probability that the valve will pass on demand $P_{p\text{-}prior}$ would be

$$P_{p-prior} = 1 - P_{f-prior} \text{………………………………...}12$$

After the inspections, a second probability of failure based on the confidence interval for the inspection, is calculated. If the confidence interval is CF, then the conditional probability of failure on demand $P_{f\text{-}cond}$ would be calculated as

$$P_{f-cond} = (1 - CFpass).P_{p-prior} \text{……………………..}13$$

and with a failed inspection, it is calculated as

$$P_{f-cond} = CF_{fail} \bullet P_{f-prior} + (1 - CFpass).P_{p-prior} \text{…………}14$$

A weighted probability of failure $P_{f\text{-}wgt}$ is hen calculated when all weighing factors have being put into consideration so as to add more credence to the results after which the posterior probability of failure on demand is then calculated using the equations provided in table 9 of the Appendix section.

The characteristic life can then be adjusted according using the equation below

$$h_{upd} = \frac{t}{\left(-\ln\left[1 - P_{f-wgt}\right]\right)^{\frac{1}{b}}} \text{………………………}15$$

The complete procedures used for the calculation of the probability of failure to open at a give inspection interval can be seen on section 7.2.6 of the API RBI 581 standard.

It is very important that a leak test be conducted but I will not be treating the probability of leakage in this work. Section 7.3 of the API RBI 581 standard describes how this probability can be obtained in cases where there is a complete set of data.

4.1.2.3 Consequence Evaluation

It is very difficult to perform a quantitative consequence analysis, and for outcomes such as fires, explosions and toxic exposures, the consequence calculations are done in accordance with section 3 of the API RBI standard. The API RBI methodology calculates the consequence of each safety valve failure at a much higher overpressure than normal operating pressure levels. The safety valve consequence calculations are closely linked to the protected equipment so that the existing damage state can easily be determined. The calculation procedures needed for determining the consequence of safety valve failure to open according to the API RBI methodology are as follows:

1. Determine the list of overpressure scenarios applicable to the piece of equipment being protected by the valve. Table 2 of the appendix provides the list of all overpressure demand cases.
2. Estimate the amount of overpressure P_o for each demand case, which is likely to occur given that a given safety valve fails to open. Table 3 of appendix provides all the needed guidance here.
3. For installations that have multiple pressure safety valves, the overpressure adjustment factor F_a should be calculated using the equation

$$F_a = \sqrt{\frac{A^{prod}}{A^{prod}_{total}}} \quad \text{........................16}$$

where

A^{prod} = orifice area of valve

A^{prod}_{total} = total installed orifice area of multiple valve installation

4. Reduce the overpressure determined in step 2 by using the overpressure adjustment factor in step 3 in the equation below

$$P_o = F_a \cdot P_o \quad \text{..........................17}$$

5. For each overpressure demand case, calculate the financial consequences of loss of containment from the protected equipment. All cost evaluations to be done with the owners work order costs, however, API RBI 581 has some equations that can be used to calculate the cost of all consequences resulting from a failure to close and from a leakage.

4.1.2.4 Risk Analysis

In this section I shall only be talking about the risk associated with failure to open since I only have pressure test results to work with. The Risk for a safety valve failing to open at a specified inspection interval, t_{insp}, is determined for each overpressure demand case using the POF of the PSV and the total consequence of failure C_f in the equation below as:

$$Risk_f^{psv} = P_f^{psv} \cdot C_f^{psv} \quad \text{...........................18}$$

The overall risk will then be the sum of the risk from all overpressure demand cases and this is expressed as

$$Risk_f^{psv} = \sum_{j=1}^{ndc} P_{f.j}^{psv} \cdot C_{f.j}^{psv} = \sum_{j=1}^{ndc} Risk_{f.j}^{psv} \ldots\ldots\ldots\ldots..19$$

When there is sufficient leakage data, then the risk would be the sum of risk to fail to open and the risk of leakage.

4.1.3 Inspection Planning and Test intervals based on Risk Analysis

Using risks results to determine inspection and testing intervals is dependent on a probability outcome and a consequence outcome. Since the probability of failure of the safety valves increases with time, and the consequences associated with valve failure also increases over time, it is correct to imply that the risk increases as a function of time too. Hence if we manage to establish a risk curve dependent on time and pick a risk target, we can manage to recommend an inspection and testing interval for safety valves. This would be equal to the time at which the risk value equals the risk target. Figure 7.7 in the appendix section shows a risk curve, and the effect of testing, inspection and repair of the pressure safety valves. The risk curve is based on the assumption that once a valve is repaired at the time of testing, the risk of failures drops to zero, and almost seen to be like new valves.

4.2 Corrosion Rate Determination and Remaining Life Calculation

Apart from a risk based inspection, another tool that can help add credence to the risk methodology described in section 4.1 above, is an examination of the thickness of the inlet of the valve. This can be done in two ways and how this is done is well documented in the API 510 standard.

4.2.1 Determination of Corrosion rate

According to the API 510 standard, corrosion rate is used to monitor thinning damage mechanisms and it is usually calculated as the difference between thickness readings, divided by a certain time interval. It could either be a long term corrosion rate determination or a short term corrosion rate determination. Short term corrosion rate determinations uses thickness results obtained from recent thickness readings, and are usually the most recent thickness readings, while uses one recent thickness reading and one that must have being taken earlier on in the life of the equipment. This helps us differentiate between current corrosion mechanisms from those acting over the long term. Mathematically, the corrosion rates are written as:

$$\text{Corrosion Rate (LT)} = (t_{initial} - t_{actual}) / (\text{time between } t_{initial} \text{ and } t_{actual}) \ldots\ldots \text{(20)}$$

$$\text{Corrosion Rate (ST)} = (t_{previous} - t_{actual}) / (\text{time between } t_{previous} \text{ and } t_{actual}) \ldots \text{(21)}$$

Where

$t_{initial}$ = the initial thickness as the same CML as t_{actual}.

t_{actual} = the actual thickness of a CML, measured during the most recent inspection

$t_{previous}$ = the previous thickness measured during the prior inspections.

Long-term and Short-term corrosion rates should be compared as part of the data assessment. API 510 advices any inspector to select the corrosion rate that best reflects the current conditions of the safety valves.

4.2.2 Remaining Life Calculations

The remaining life of the vessel (in years) is usually calculated as:

$$remaining\ life = \frac{t_{actual} - t_{required}}{corrosion\ rate} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(22)}$$

Where

t_{actual} = the actual thickness of a CML, measured during the most recent inspection

$t_{required}$ = the required thickness at the same CML or component as the t_{actual} measurement.

Note: When a statistical analysis is to be done, it must be done in such a way that it reflects the true condition of the section vessel section where the relief valves or PSVs are positioned. The corrosion analysis has only being explained in this work because if it is done properly as a supplementary tool to the RBI assessment treated earlier on, it adds credence to any inspection interval that is proposed, and ultimately plays a part in the determination of an appropriate testing interval. However, the biggest use of corrosion analysis would be in the trending of the aging patterns of safety valves. Only pressure test outcomes shall be considered in the trend analysis in this report.

4.3 Examination of pressure test results of PSVs from IKM Laboratorium AS

I have looked at the pressure test results for 110 safety valves being tested at IKM Laboratorium AS, and I have only considered those results for which the ratio of the pop up pressure (OP) to set pressure (SP) is between 0.7 and 1.7. The remaining results for the other seven valves have being disregarded in this sense as their pressure ratios fall outside the range I have worked with. I have only considered the reports and results of the testing personnel and I have not done any of the tests myself too. The test results and a cumulative frequency curve for the data I collected are shown below.

Table I

Pressure test results for accepted safety valves.

	OP/SP	frequency	% freq	cumulative frequency
	0,85	1	0,97 %	0,97 %
	0,9	1	0,97 %	1,94 %
	0,95	7	6,80 %	8,74 %
	1	69	66,99 %	75,73 %
	1,05	10	9,71 %	85,44 %
	1,1	5	4,85 %	90,29 %
	1,15	2	1,94 %	92,23 %
	1,3	1	0,97 %	93,20 %
	1,4	1	0,97 %	94,17 %
	1,5	4	3,88 %	98,06 %
	1,7	2	1,94 %	100,00 %
Sum		**103**	**100,00 %**	

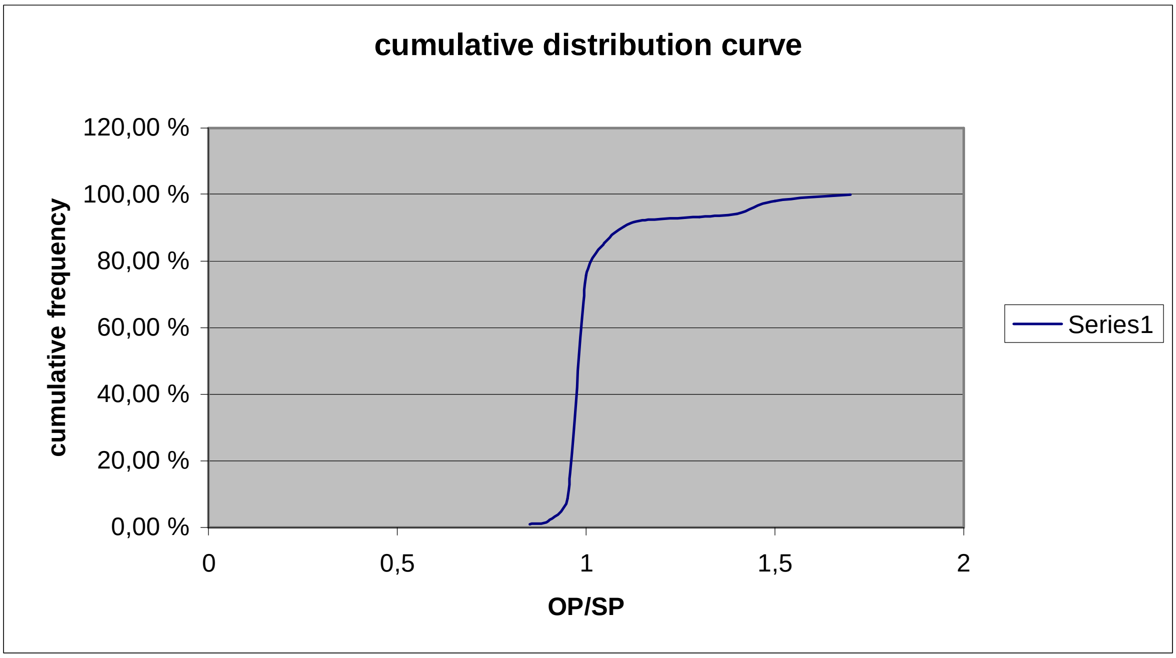

Figure 5: Cumulative frequency curve of test results

I have used the pressure test results and cumulative frequency to set a confidence level of 93.2% at which OP/SP = 1.3. This is the confidence level I have utilised in estimating an appropriate inspection and testing interval for the pressure safety valves.

This work involves calculating the probabilities of failures for safety valves working under both mild and severe service conditions, and then trying to find an estimated time interval for such failures to occur using a Weibull probability distribution. In a more detailed work, the cost of the consequences for each valve ought to be computed and then a risk value found for each service location which allows for risk ranking. However, I would be using a more conservative approach in this work by only relying on the results for the probability of failure on demand for the valves, and try to make my estimates from those results.

4.4 Estimation of Optimum inspection and testing interval for PSVs

Case 1:

I will first consider a valve operating in a MILD service environment. By MILD condition, I mean the best case situation with operating temperatures less than 500°F and 90% <OP/SP<150%. I also want to make it known that all the valves have being treated like conventional and balanced PSVs. Hence, from table 5 of the appendix, I have used η = 50.5, and β = 1.8. In the event that a leak test result is available, the probability of leakage result should be adjusted by a factor of 0.6.

I have said I would be using a confident level, CF = 0.932 at OP/SP = 1.3.

From figure 4 of the appendix section, the failure rate associated with this confidence level is approximately 3%. This would then give us a prior probability of failure on demand calculated using equation 11. However, we must first calculate the valve of η_{mod} using equation 9. That is:

$$\eta_{mod} = F_c \cdot F_{op} \cdot F_{env} \cdot \eta_{def}$$

Using F_c = 0.75, F_{op} = 1, and F_{env} = 1 (since I have assumed mild conditions), then

$$\eta_{mod} = 0.75 \times 1 \times 1 \times 50.5 = 37.875$$

So that,

$$P_{f-prior} = 1 - \exp\left[-\left(\frac{t}{h_{\text{mod}}}\right)^b\right]$$ = 0.01036 if we assume an inspection interval of 3 years.

We can go on to get the prior probability of pass on demand given as

$$P_{p-prior} = 1 - P_{f-prior} = 1 - 0.01036 = 0.98964$$

Since we have deleted data from the sample examined, we would use equation 13 (neglecting the effect of the deleted data since they are so few). This would give us a conditional probability of failure on demand computed as

$$P_{f-cond} = (1 - CFpass).P_{p-prior} = (1 - 0.932) \cdot 0.98964 = 0.0673$$

This enables us to compute the weighted probability of failure on demand using the appropriate equation from table 2 of the appendix

$$P_{f-wgt} = P_{f-prior} - 0.2P_{f-prior}\left(\frac{t}{h_{\text{mod}}}\right) + 0.2P_{f-cond}\left(\frac{t}{h_{\text{mod}}}\right) = 0.01126$$

From the weighted value, the characteristic lives of the valves are updated using equation 15 so that the value is updated to

$$h_{upd} = \frac{t}{\left(-\ln\left[1 - P_{f-wgt}\right]\right)^{\frac{1}{b}}} = 36.16$$

At this point I can now moving on to determine the probability of failure to open for the valves and this is computed using equation 6, POF = POFOD × DR × PF.

Where,

DR = EF×DRRF = 0.5 × 0.5 = 0.25 (assuming the average values for all overpressure conditions)

The POFOD value is obtained from equation 10 as

$$POFOD = 1 - \exp\left[-\left(\frac{t}{h_{upd}}\right)^{b}\right] = 0.01126$$

Since I have used a pressure ratios of 1.3 in selecting my confidence level, I have selected a value of P_f = 0.2 using the curve in figure 4 of the appendix.

This

POF = 0.25 × 0.01126 × 0.2 = 5.63E-4

Using the Weibull function, the time interval in years can be estimated from the equation, $F(t) = 1 - \exp\left[-\left(\frac{t}{h_{upd}}\right)^{b}\right]$, and solving for t gives a value of t = 0.567 years or approximately once in seven months.

Case 2:

Now I will look at those valves operating in a SEVERE service environment. By Severe condition, I mean a worst case situation with operating temperatures more than 500^{o}F and 80% <OP/SP<170%. From table 5 of the appendix, I have used η = 17.6, and β = 1.8 to represent the service conditions in a severe case. Just like in the mild case, if leak test results are available, the probability of valve leakage should also be adjusted by a factor of 0.6.

I have also used a confident level, CF = 0.932 at OP/SP = 1.3 and from figure 4 of the appendix section, the failure rate associated with this confidence level is approximately 3%. This would then give us a prior probability of failure on demand calculated using equation 11. That is:

$$\eta_{mod} = F_c \cdot F_{op} \cdot F_{env} \cdot \eta_{def}$$

And using F_c = 0.75, F_{op} = 1, and F_{env} = 0.8 (since the environment factors do shift the POFOD curve to the left as seen from figure 5 of the appendix), then

$$\eta_{mod} = 0.75 \times 1 \times 0.8 \times 17.6 = 10.56$$

So that,

$$P_{f-prior} = 1 - \exp\left[-\left(\frac{t}{\eta_{mod}}\right)^{\beta}\right]$$ = 0.0986 if we assume an inspection interval of 3 years.

We can go on to get the prior probability of pass on demand given as

$$P_{p-prior} = 1 - P_{f-prior} = 1 - 0.0986 = 0.9014$$

Since we have deleted data from the sample examined, we would use equation 13 (neglecting the effect of the deleted data since they are so few). This would give us a conditional probability of failure on demand computed as

$$P_{f-cond} = (1 - CFpass).P_{p-prior} = (1 - 0.932) \cdot 0.9014 = 0.0613$$

This enables us to compute the weighted probability of failure on demand using the appropriate equation from table 2 of the appendix

$$P_{f-wgt} = P_{f-prior} - 0.2P_{f-prior}\left(\frac{t}{\eta_{mod}}\right) + 0.2P_{f-cond}\left(\frac{t}{\eta_{mod}}\right) = 0.0509$$

From the weighted value, the characteristic lives of the valves are updated using equation 15 so that the value is updated to

$$\eta_{upd} = \frac{t}{\left(-\ln\left[1 - P_{f-wgt}\right]\right)^{\frac{1}{\beta}}} = 15.465$$

At this point I can now moving on to determine the probability of failure to open for the valves and this is computed using equation 6, POF = POFOD × DR × PF.

Where,

DR = EF×DRRF = 0.5 × 0.5 = 0.25 (assuming the average values for all overpressure conditions)

The POFOD value is obtained from equation 10 as

$$POFOD = 1 - \exp\left[-\left(\frac{t}{h_{upd}} \right)^{b} \right] = 0.051$$

Since I have used a pressure ratios of 1.3 in selecting my confidence level, I have selected a value of P_f = 0.2 using the curve in figure 4 of the appendix.

This

POF = 0.25 × 0.051 × 0.2 = 2,25E-3

Using the Weibull function, the time interval in years can be estimated from the equation, $F(t) = 1 - \exp\left[-\left(\frac{t}{h_{upd}} \right)^{b} \right]$, and solving for t gives a value of t = 0.07 years or approximately twice in three months or to be on a safe side, once every two months.

Chapter 5

5.1 Discussion and Recommendation

The time intervals arrived at in chapter 4 of this project is very conservative due to insufficient data plus the fact that I neglected the probability of leakage and consequence calculations, which means the solutions for the time intervals for both the testing and inspections of PSVs would be shorter than necessary in a real plant. The results of the calculations show that for a worst case situation which is likely to be for those PSVs operating under very adverse environmental conditions, the time interval needed to limit the probability of failure to open to 2.25E-3 would be at most once in 2 months. On the other hand, we would need an interval of about 6 to 7 months to limit the probability of failure to open for a PSV to 5.63E-4.

API 510 states that PSVs shall be tested and inspected at intervals that are as frequent enough to verify and guaranty that valves perform reliably in a particular service condition when they are called upon. Its also suggests an interval of 5 years for PSVs used for typical process services and 10 years for PSVs in clean and non corrosive services. The problem with this time line is that PSVs have no indicators that can enable operators carry out any kind of meaningful condition monitoring, so it would not advice any firm to leave their PSVs in service for so long without determining how well they are functioning. The risk to personnel and plant is far too great for such a chance to be taken. However, API 510 also states that the intervals could be moved according to test results. This implies that the test intervals could be far shorter or longer than the stated times mentioned in the standard.

Also the NORSOK standard P-001, states that the test intervals shall be anything from a one year period to any length of time. It also states that any pressure safety device that requires testing in an interval lesser than a one year period should be considered to be not robust enough and should not be used in the industry.

In trying to make an appropriate adjustment to my calculated values in the previous chapter, I have also used the curve in figure 5 of the appendix section as a guide. For the very low probability of failure on demand values (POFOD) obtained, the years in service obtained from the curve can be seen to be around 1 to 3 years at best. Hence it is not advisable to have a testing interval greater than 5 years for any PSV no matter how mild the service condition is. Also, the risk curve in figure 7 is used as a good guide too. I have considered the high safety level that is usually demanded by the authorities here in Norway and pegged the maximum risk that can be tolerated at 18 000 Norwegian Kroner per year. This would give an updating time of about 2 years on the average for most valves.

Taking all factors into account, including the neglected data and the various service conditions under which thee PSVs operate, it is recommended that;

1. PSVs operating in very severe service conditions and utilised in plants that are heavily manned by personnel should be inspected every two months and tested every six months.
2. PSVs operating in very mild service conditions and used in plants with little or no personnel, should be inspected every six months and tested every year.
3. Leak test results should be taken for PSVs that show significant amount of fouling or malfunction, and the test results should be used to update the maintenance programme for the PSVs.

The above recommendations can be used in any part of the world, but in Norway, no safety valves can be allowed in any operating plant if such valves require a test interval lesser than a one year period, hence it is advisable to have PSVs that are very robust operating in this region, and they should be tested once every year.

5.2 Challenges

The bulk of this work was done with very limited data (basically on pressure test results) obtained from IKM Laboratorium AS. However, the pressure test results still remains the most important indicator of the aging trend of the PSVs.

The biggest challenge encountered in this work is that there is no common standard guiding the design, installation and maintenance of PSVs. I have had to go through a lot of standards in trying to get all the information needed for testing PSVs. Also, I have not being able to visit any plant where PSVs are being utilised, and the response I received from some of the client representatives were not so comprehensive as they do not have complete records for their PSVs. All data used as field results obtained from companies database.

Despite these challenges faced while trying to obtain the test and inspection interval for PSVs, IKM Laboratorium provided me with invaluable advice and materials needed in writing this masters thesis in other to give it more credibility and objectivity.

Chapter 6

Conclusion and Further Studies

6.1 Conclusion

PSVs are very intricate and vital devices in any plant and it is imperative that the state of the PSVs is known at all times to guaranty the safety of plant personnel and protect capital investments. The only way to know if a valve is still functionally active and okay is to carry out proper inspections at the right times for any damages and fouling (like leakage) and to test these valves regularly to ascertain if they would act when called upon.

Generally, PSVs that are inspected at least once every six months and tested within six months to one year would most likely be very reliable. These time intervals are however subject to change and they can be lengthened or shortened when sufficient data becomes available, and the maintenance data base for PSVs should be updated accordingly.

6.2 Further Studies

In this work, I have done a general assessment for all PSVs. I would have loved to do this work for the different types of PSVs and for various other factors like valve inlet size and plant locations which also affect the OP/SP ratios. This way I would have being able to do a more in depth analysis using tools like Analysis of Variance (ANOVA). A good consequence analysis should be done in the future to be able to determine the risk associated with the various kinds of PSVs in various locations. This way an optimum inspection and testing for all types of PSV for each specific plant can be determined and a maintenance plan suited for a given plant can be established.

Appendix

Table 1 Basic data needed for a PSV or Pressure relief devices (PRD) module

Data	Description	Data Source
PRD Type	Type of Pressure Relief Device, (Drop Down Menu). • Conventional Spring Loaded PRV (Default) • Balanced Bellows PRV • Pilot-Operated PRV • PRV with Rupture Disk • Rupture Disk only	User Specified
Fluid Composition	Process fluid mixture components, either mass or mole fraction. Limit of 10 components in mixture definition	API RBI Fixed Equipment
Service Severity	Severity of process fluid. Choices are Mild, Moderate and Severe. The service severity provides the basis for the selection of the default probability of failure on demand and probability of leakage curves. <u>FAIL TO OPEN</u> • Mild • Moderate (Default) • Severe <u>LEAKAGE</u> • Mild • Moderate (Default) • Severe	User Specified
Overpressure Scenarios	Provide a listing of the applicable overpressure scenarios for each PRD. For each overpressure scenario, default values for the initiating event frequency and the PRD demand rate reduction factor (DRRF) are provided in Table 7.2. These two parameters when multiplied together provide an estimate of the demand rate on the PRD installation.	User Specified
PRD Discharge Location	• Atmosphere • Flare (Default) • Closed Process	User Specified
PRD Inspection History	• Date of Testing • Install Date • Type of Test (Effectiveness) • Results of Test/Inspection • Overhauled ? (Yes/No) • Piping Condition	User Specified
Protected Equipment Details	Operating conditions, Design conditions, dimensions, damage mechanisms, generic failure frequency and damage factors	RBI Fixed Equipment
Fluid Inventory	Fluid inventory associated with the protected equipment, (lbm) May be less than the RBI calculated inventory due to shut-in conditions, e.g. reactor discharge valve fails closed	RBI Fixed Equipment
Injury Costs	Cost of serious injury, $	RBI Fixed Equipment
Environmental Costs	Environmental fines and costs associated with PRD leakage or loss of equipment containment, $/event	RBI Fixed Equipment
Production Costs	Cost of Lost Production, $	RBI Fixed Equipment
Unit Costs	Cost to replace unit, $/ft2	RBI Fixed Equipment

Table 2 Default initiating event frequencies

Overpressure Demand Case	**Event Frequency**	EF_j (events/year)	$DRRF_f$ (See notes 2 and 3)	Reference
Fire	1 per 250 years	0.004	0.1	[6]
Blocked Discharge with Administrative Controls in Place (see Note 1)	1 per 100 Years	0.01	1.0	[16]
Blocked Discharge without Administrative Controls (see Note 1)	1 per 10 years	0.1	1.0	[16]
Loss of Cooling Water Utility	1 per 10 years	0.1	1.0	[6]
Thermal Relief with Administrative Controls in Place(see Note 1)	1 per 100 Years	0.01	1.0	Assumed same as Blocked Discharge
Thermal Relief without Administrative Controls (see Note 1)	1 per 10 years	0.1	1.0	Assumed same as Blocked Discharge
Electrical Power Supply failure	1 per 12.5 years	0.08	1.0	[6]
Control Valve Failure, Initiating event is same direction as CV normal fail position (i.e. Fail safe)	1 per 10 years	0.1	1.0	[17]
Control Valve Failure, Initiating event is opposite direction as CV normal fail position (i.e. Fail opposite)	1 per 50 years	0.02	1.0	[17]
Tower P/A or Reflux Pump Failures	1 per 5 years	0.2	1.0	
Runaway Chemical Reaction	1 per year	1.0	1.0	
Liquid Overfilling with Administrative Controls in Place (see Note 1)	1 per 100 years	0.01	0.1	[6]
Liquid Overfilling without Administrative Controls (see Note 1)	1 per 10 years	0.01	0.1	[6]
Heat Exchanger Tube Rupture	1 per 1000 years	0.001	1.0	[18]

Notes:
1. Administrative Controls for isolation valves are procedures intended to ensure that personnel actions do not comprise the overpressure protection of the equipment.
2. The DRRF recognizes the fact that demand rate on the PRD is often less than the initiating event frequency. As an example, PRDs rarely lift during a fire since the time to overpressure may be quite long and fire fighting efforts are usually taken to minimize overpressure.
3. The DRRF can also be used to take credit for other layers of overpressure protection such as control and trip systems that reduce the likelihood of reaching PRD set pressure

Table 3 Overpressures scenario logic

Initiating Event Frequency	Equipment Type	PRD Demand Rate Reduction Factor	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario – Blocked Discharge, Manual Valve					
1 per 100 years (admin controls) 1 per 10 years (w/o admin controls Multiply Event Frequency times the # of applicable block valves located in process flow path. Lees, 1983 [16] suggests an estimated rate of 0.5 to 0.1 events per year for shutting manual valve in error	Exchangers, Fin Fans, Reactors, Piping, Drums or Rotating Equipment	1.0	Downstream of rotating equipment other than positive displacement type	Deadhead Pressure or 1.3 times the normal discharge pressure or Bubble Point pressure of the feed stream at heat source temperature (for cases where the equipment has internal or external heat sources), whichever is greatest	Most centrifugal rotating equipment will deadhead at 30% above the normal operating point. Initiating event frequency should be adjusted if the protected equipment is removed from service for maintenance or operational needs (filter replacement or cyclic process operation) at a frequency greater than the unit turnaround frequency. Equipment with internal or external heat sources may have a significant potential for overpressure as a result of vaporization of the contained fluid stream.
		1.0	Downstream of positive displacement type rotating equipment	4.0 X MAWP (Rupture)	Discharge pressure from positive displacement pumps will continue to increase pressure. Assumption is made that rupture will occur.
		1.0	Downstream of Steam Turbines	Steam Supply Pressure or Bubble Point pressure of the feed stream at steam supply temperature (for cases where the equipment has internal or external heat sources), whichever is greatest	
		1.0	Downstream of Process Units or vessels	1.1 X MAWP of Upstream Vessel Source Pressure	
	Process Tower with Fired Heater heat source	1.0 Consider LOPA or risk reduction analysis associated with loss of flow controls on the fired heater	Heat Source to tower is a fired heater	4.0 X MAWP (Rupture)	Assumption is made that rupture occurs. This applies to the blocked vapor outlet line only, see Liquid Overfilling Case for blocked liquid/bottoms outlet
	Process Tower, all other heat sources	1.0	No Upstream Fired Heater	Bubble Point pressure of the feed stream at heat source temperature	This applies to the blocked vapor outlet line only, see Liquid Overfilling Case for blocked liquid/bottoms outlet
	Heaters	1.0		4.0 X MAWP (Rupture)	Added increase in potential overpressure with fired/radiant heat transfer. Assumption is made that rupture occurs.

Table 3 Overpressure scenario logic

Initiating Event Frequency	Equipment Type	PRD Demand Rate Reduction Factor	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario – Control Valve Fail Close at Outlet					
1 per 10 years [17] for fail-closed Control Valves 1 per 50 years for fail-open Control Valves Multiply Event Frequency times the # of applicable control valves located in process flow path.	Exchangers, Fin Fans, Reactors, Piping or Drums or Rotating Equipment	1.0	Downstream of rotating equipment other than positive displacement type	Deadhead Pressure or 1.3 times the normal discharge pressure or Bubble Point pressure of the feed stream at heat source temperature (for cases where the equipment has internal or external heat sources), whichever is greatest	Most centrifugal rotating equipment will deadhead at 30% above the normal operating point. Initiating event frequency should be adjusted if the protected equipment is removed from service for maintenance or operational needs (filter replacement or cyclic process operation) at a frequency greater than the unit turnaround frequency. Equipment with internal or external heat sources may have a significant potential for overpressure as a result of vaporization of the contained fluid stream.
		1.0	Downstream of positive displacement type rotating equipment	4.0 X MAWP (Rupture)	Discharge pressure from positive displacement pumps will continue to increase pressure. Assumption is made that rupture will occur.
		1.0	Downstream of Steam Turbines	Steam Supply Pressure or Bubble Point pressure of the feed stream at steam supply temperature (for cases where the equipment has internal or external heat sources), whichever is greatest	
	Process Tower with Fired Heater heat source	1.0	Downstream of Process Units or vessels	1.1 X MAWP of Upstream Vessel Source Pressure	
		1.0 Consider LOPA or risk reduction analysis associated with loss of flow controls on the fired heater	Heat Source to tower is a fired heater	4.0 X MAWP (Rupture)	Assumption is made that rupture occurs. This applies to the blocked vapor outlet line only, see Liquid Overfilling Case for blocked liquid/bottoms outlet
	Process Tower, all other heat sources	1.0		Bubble Point pressure of the feed stream at heat source temperature	This applies to the blocked vapor outlet line only, see Liquid Overfilling Case for blocked liquid/bottoms outlet
	Heaters	1.0		4.0 X MAWP (Rupture)	Added increase in potential overpressure with fired/radiant heat transfer. Assumption is made that rupture occurs.

Table 3 Overpressure Scenario logic

Initiating Event Frequency	Equipment Type	PRD Demand Rate Reduction Factor	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario – Control Valve Fail Open at Inlet, including the HP/LP Gas Breakthrough Case					
1 per 10 years [17] for fail-closed Control Valves 1 per 50 years for fail-open Control Valves Multiply Event Frequency times the # of applicable control valves located in process flow path.	All Equipment Types	1.0	N/A	Use the upstream source pressure	Overpressure Potential is a function of the pressure ratio across the control valve
Overpressure Scenario – Fire					
1 per 250 years See Lees [16] page A7-7, states major fire at plant 1 every 10 years	All Equipment Types	0.1 Industry experience justifies this value	N/A	4.0 X MAWP (Rupture)	Modified by industry data which indicates demand rates on the order of 1 per 400 years The DRRF factor of 0.1 recognizes the industry experience that relatively few vessels exposed to a fire will experience a PRD opening. Assumption is made that in those rare cases where a PRD would open during a fire, rupture will occur if the PRD failed to open upon demand

Table 3 Overpressure Scenario Logic

Initiating Event Frequency	Equipment Type	PRD Demand Rate Reduction Factor	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario – Thermal /Hydraulic Expansion Relief					
1 per 100 years (manual valve w/admin controls) 1 per 10 years (manual valve w/o admin controls or control valve Multiply initiating event frequency times the number of applicable block valves located in process flow path.	Piping or other liquid filled equipment	1.0	N/A	Operating Pressure or Bubble Point pressure of contained fluid at 140 °F, whichever is larger	Assumption is made that the probability of a leak is 1.0 (flange leaks), modeled as a ¼ inch hole. The probability of rupture is assumed to be 0.0. For fluids that will not boil, since the pressure is relieved immediately upon leakage, the pressure for the consequence calculation will be the normal operating pressure of the piping. Not likely to result in rupture, likely to cause flange leaks/small leaks, heated only If the fluid can boil due to solar energy, the consequence pressure could be maintained at the bubble point pressure of the contained fluid. Leak and rupture probabilities will be calculated as a function of the bubble point pressure.
	Cold side of Heat Exchangers	1.0	N/A	Operating Pressure or Bubble Point pressure of contained fluid at the hot side fluid inlet temperature, whichever is larger	Added increase in potential overpressure with additional heat transfer from hot side. For liquids that do not boil, the assumption is made that the probability of leak is 1.0 (flange leaks), modeled as a ¼ inch hole, and the probability of rupture is 0.0. If the cold side fluid can boil, the consequence pressure could reach the bubble point pressure of the stored fluid at the hot side fluid inlet temperature. Leak and rupture probabilities will be calculated as a function of the bubble point pressure.
Overpressure Scenario – Loss of Cooling					
1 per 10 years	Process Tower with Fired Heater heat source	1.0 Consider LOPA or risk reduction analysis associated with loss of flow controls on the fired heater	Heat Source to tower is a fired heater	4.0 X MAWP (Rupture)	Assumption is made that rupture occurs.
	All Other Equipment with internal or external heat sources	1.0		Bubble Point pressure of the feed stream at heat source temperature	

Table 3 Overpressure scenario logic

Initiating Event Frequency	Equipment Type	PRD Demand Rate Reduction Factor	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario – Tower P/A or Reflux Pump Failure					
1 per 5 years	Process Tower with Fired Heater heat source	1.0 Consider LOPA or risk reduction analysis associated with loss of flow controls on the fired heater	Heat Source to tower is a fired heater	4.0 X MAWP (Rupture)	Assumption is made that rupture occurs.
	All Other Process Towers	1.0		Bubble Point pressure of the feed stream at heat source temperature	
Overpressure Scenario – Electrical Power Failure					
0.08 per year (1 per 12.5 years) power supply failure per Table on page 9/30 of [16]	Process Tower with Fired Heater heat source	1.0 Consider LOPA or risk reduction analysis associated with loss of flow controls on the fired heater	Heat Source to tower is a fired heater	4.0 X MAWP (Rupture)	Assumption is made that rupture occurs.
	Process Tower and Other Equipment with internal or external (non-fired) heat sources	1.0		Bubble Point pressure of the feed stream at heat source temperature	
Overpressure Scenario – Runaway Chemical Reaction					
1 per year	All Equipment	1.0		4.0 X MAWP (Rupture)	This overpressure scenario should be based on a thorough review of the wide variety of potential initiating events and mitigation measures associated with the reactor system installation. The DRRF and the potential overpressure associated with failure of PRD to open upon demand should be chosen based on a risk assessment. Per Shell study, 50% of all vessel ruptures are attributed to reactive overpressure case.
Overpressure Scenario – Tube Rupture					
1 per 1000 years (9×10^{-4} per exchanger per [18])	Exchangers – HP Gas in Tubes, LP Liquid in Shell	1.0		Normal Maximum Operating Pressure of the high pressure side of the exchanger	Likelihood of shell rupture is increased when high pressure tubeside gas enters low pressure shellside liquid

Table 3 Overpressure scenario logic

Initiating Event Frequency	Equipment Type	PRD Demand Rate Reduction Factor	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario – Liquid Overfilling					
1 per 100 years (admin controls) 1 per 10 years (w/o admin controls Multiply Event Frequency times the number of applicable block valves located in process flow path.	All Equipment including Process Tower (Blocked Outlet of Liquid Bottoms)	1.0	Downstream of rotating equipment other than positive displacement type	Deadhead Pressure or 1.3 times the normal discharge pressure or Bubble Point pressure of the feed stream at heat source temperature (for cases where the equipment has internal or external heat sources), whichever is greatest	Most centrifugal rotating equipment will deadhead at 30% above the normal operating point. Initiating event frequency should be adjusted if the protected equipment is removed from service for maintenance or operational needs (filter replacement or cyclic process operation) at a frequency greater than the unit turnaround frequency. Equipment with internal or external heat sources may have a significant potential for overpressure as a result of vaporization of the contained fluid stream.
		1.0	Downstream of positive displacement type rotating equipment	4.0 X MAWP (Rupture)	Discharge pressure from positive displacement pumps will continue to increase pressure. Assumption is made that rupture will occur.
		1.0	Downstream of Steam Turbines	Steam Supply Pressure or Bubble Point pressure of the feed stream at steam supply temperature (for cases where the equipment has internal or external heat sources), whichever is greatest	
		1.0	Downstream of Process Units or vessels	1.1 X MAWP of Upstream Pressure Source Vessel	

Table 4 Categories of PRD Service Severity

Service	Description
Mild	Clean hydrocarbon products at moderate temperature. No aqueous phase present. Low in sulfur and chlorides. Failure is characterized by a long (25 years) $MTTF$. Failure is strongly characterized as a "wear out" type of failure, in which the failure occurs due to an accumulation of damage over a long period of time. Examples include, product hydrocarbon streams (including lubricating oils), LPG, BFW, low pressure steam and clean gasses such as nitrogen and air.
Moderate	Hydrocarbons that may contain some particulate matter. A separate aqueous phase may be present, but is a minor component; however, clean, filtered and treated water may be included in this category. Some sulfur or chlorides may be present. Temperatures of up to 500°F may exist. Failure occurs at an average (15 years) $MTTF$. Failure is weakly characterized as a "wear out" type of failure, in which the failure occurs due to an accumulation of damage. Examples include, intermediate hydrocarbon streams, in-service lube and seal oils, process water (not cooling water or BFW) and medium to high pressure steam.
Severe	Hydrocarbons that are processed at temperatures above 500°F with significant tendency to foul. Sulfur and chloride concentrations may be high. Monomers processed at any temperature that can polymerize are in this group as well. Sometimes included are aqueous solutions of process water, including cooling water. Failure is characterized as a relatively short (7 years) $MTTF$. Failure is characterized as a "random" type of failure, in which the failure can occur due to a variety of mechanisms such as corrosion or plugging. Examples include; Heavy hydrocarbon streams such as crude, amine services, cooling water, corrosive liquids and vapors, and streams containing H_2S.

Table 5 Default Weibull parameters for probability of failure on demand

Fluid Severity	Conventional and Balanced Bellows PRVs [1]		Pilot-Operated PRVs [2]		Rupture Disks [3]	
	β	η	β	η	β	η
Mild	1.8	50.5	1.8	33.7	1.8	50.5
Moderate	1.8	23.9	1.8	8.0	1.8	50.5
Severe	1.8	17.6	1.8	3.5	1.8	50.5

Notes:
1. The η parameter values for conventional PRVs are reduced by 25% if the discharge is to a closed system or to flare.
2. The η parameter values for pilot-operated valves are currently based on the conventional PRV data; however, reduced by a factor of 1.5, 3, and 5 for Mild, Moderate and Severe services, respectively.
3. Without any failure rate data for rupture disks, the conventional PRV values for Mild services were used. This assumes that the RD material has been selected appropriately for the fluid service.

Table 6 Environmental adjustment factors.

Environment Modifier	Adjustment to POFOD η Parameter	Adjustment to POL η Parameter
Operating Temperature 200<T<500°F	1.0	0.8
Operating Temperature > 500°F	1.0	0.6
Operating Ratio >90% for spring-loaded PRVs or >95% for pilot-operated PRVs	1.0	0.5[1]
Installed Piping Vibration	1.0	0.8
Pulsating or Cyclical service, such as Downstream of Positive Displacement Rotating Equipment	1.0	0.8
History of Excessive Actuation in Service (greater than 5 times per year)	0.5	0.5[2]
History of Chatter	0.5	0.5

Notes:
1. Some Pilot-operated PRVs operate extremely well with operating ratios approaching 98%. In these cases, the environmental factor should not be applied.
2. This factor will not be applied if the environmental factor for operating ratio is already applied.

Table 7 Inspection and Testing Effectiveness

Inspection Effectiveness	Component Type	Description of Inspection
Highly Effective A	Pressure Relief Device	A bench test has been performed on the PRV in the as-received condition from the unit and the initial leak pressure, opening pressure and the reseat pressure has been documented on the test form. The inlet and outlet piping has been examined for signs of excessive plugging or fouling.
	Rupture Disk	None Available.
Usually Effective B	Pressure Relief Device	• A bench test has been performed, however, the PRD was cleaned or steamed out prior to the bench test. Additionally, a visual inspection has been performed where detailed documentation of the condition of the PRD internal components was made. • An in-situ test has been performed using the actual process fluid to pressurize the system.
	Rupture Disk	The rupture disk is removed and visually inspected for damage or deformations.
Fairly Effective C	Pressure Relief Device	• A visual inspection has been performed without a pop test, where detailed documentation of the condition of the PRD internal components was made. • A trevitest or in-situ test has been performed where the actual process fluid was not used to pressurize the system.
	Rupture Disk	The space between the disk and the PRV is monitored for leakage in accordance with the ASME Code and API RP 520 Part 2.
Ineffective D	Pressure Relief Device	No pop test was conducted.
	Rupture Disk	No details of the internal component were documented.

Table 8 Level of inspection Confidence Factors,

Inspection Result	Confidence Factor That Inspection Result Determines the True Damage State, CF			
	Ineffective	Fairly Effective	Usually Effective	Highly Effective
Pass, CF_{pass}	No credit	0.5	0.70	0.9
Fail, CF_{fail}	No Credit	0.70	0.95	0.95
No Leak, CF_{noleak}	No Credit	0.5	0.70	0.9
Leak, CF_{leak}	No Credit	0.70	0.95	0.95

Table 9: Inspection updating equations

Inspection Effectiveness and Result	Equation for Weighted Probability of Failure on Demand
Highly Effective Pass	$P_{f,wgt}^{prd} = P_{f,prior}^{prd} - 0.2 \cdot P_{f,prior}^{prd}\left(\frac{t}{\eta}\right) + 0.2 \cdot P_{f,cond}^{prd}\left(\frac{t}{\eta}\right)$
Usually Effective Pass	$P_{f,wgt}^{prd} = P_{f,prior}^{prd} - 0.2 \cdot P_{f,prior}^{prd}\left(\frac{t}{\eta}\right) + 0.2 \cdot P_{f,cond}^{prd}\left(\frac{t}{\eta}\right)$
Fairly Effective Pass	$P_{f,wgt}^{prd} = P_{f,prior}^{prd} - 0.2 \cdot P_{f,prior}^{prd}\left(\frac{t}{\eta}\right) + 0.2 \cdot P_{f,cond}^{prd}\left(\frac{t}{\eta}\right)$
Highly Effective Fail	$P_{f,wgt}^{prd} = P_{f,cond}^{prd}$
Usually Effective Fail	$P_{f,wgt}^{prd} = P_{f,cond}^{prd}$
Fairly Effective Fail	$P_{f,wgt}^{prd} = 0.5 \cdot P_{f,prior}^{prd} + 0.5 \cdot P_{f,cond}^{prd}$

Table 10: Damage classes for protected equipment

Damage Factor Class	Damage Factor	Description
None	1	New vessel or inspection shows little if any damage.
Minimal	20	Equipment has been in service for a reasonable amount of time and inspection shows evidence of minor damage.
Minor	200	One or more damage mechanisms have been identified, limited inspection data available and fairly moderate evidence of damage. Single damage mechanism identified, recent inspection indicates moderate evidence of damage.
Moderate	750	Moderate damage found during recent inspection. Low susceptible to one or more damage mechanisms, and limited inspection exists.
Severe	2000	One or more active damage mechanisms present without any recent inspection history. Limited inspection indicating high damage susceptibility.

Table 11: Categories of PRD service severity

Service	Description
Mild	Many heavy liquid streams such as crude oil tend not to leak through a PRV. Cooling water and amine services are some examples of a corrosive/fouling fluids that do not leak. Additionally, clean fluids such as LPG, air, and nitrogen are MILD leakage services.
Moderate	Most of the intermediate and product HC streams, most HC vapors, lube, seal and cycle oils and process water (not cooling water or BFW).
Severe	BFW/Condensate, Steam and corrosive liquids such as caustic and acids.

Table 12 Default Weibull parameters for probability of leakage

Fluid Severity	Conventional PRVs [1]		Balanced Bellows PRVs [1]		Pilot-Operated PRVs [2]		Rupture Disks [3]	
	β	η	β	η	β	η	β	η
Mild	1.6	17.5	1.6	16.0	1.6	17.5	1.6	17.5
Moderate	1.6	15.5	1.6	14.0	1.6	15.5	1.6	17.5
Severe	1.6	13.1	1.6	11.5	1.6	13.1	1.6	17.5

Notes:
1. The η parameter values are increased by 25% for Conventional and Balanced PRVs that have soft seats.
2. The η parameter values for pilot-operated valves are currently based on the conventional PRV data, since there is currently no failure rate data to support otherwise.
3. Without any failure rate data for rupture disks, the conventional PRV values for Mild service were used.

Table 13 Potential consequence of pressure vessel over pressure

Accumulation (% over MAWP)	Significance	Potential Consequence
10%	ASME code allowable accumulation for process upset cases (non-fire) protected by a single relief device	No expected consequence at this accumulation level.
16%	ASME code allowable accumulation for process upset cases protected by multiple relief devices	No expected consequence at this accumulation level.
21%	ASME code allowable accumulation for external fire relief cases regardless of the number of relief devices	No expected consequence at this accumulation level.
50%	ASME standard hydrostatic test pressure (may be 30% on new designs)	Possible leaks in associated instrumentation, etc. Medium consequence.
90%	Minimum yield strength (dependent on materials of construction)	Catastrophic vessel rupture, remote possibility. Significant leaks probable. Failure of damaged vessel areas (corrosion, cracks, blisters, etc. likely. High consequence.
300%	Ultimate tensile strength (dependent on materials of construction)	Catastrophic vessel rupture predicted. Highest consequence.

Table 14 Estimated Leakage duration

PRD Inlet Size (inches)	Leak Duration Discharge to Flare or Closed System, D_{leak} (days)	Leak Duration Discharge to Atmosphere, D_{leak} (days)
≤ 3/4 inch	60	8
3/4 < inlet size ≤ 1-1/2	30	4
1-1/2 < inlet size ≤ 3	15	2
3 < inlet size ≤ 6	7	1
Greater than 6	2	0.33

Table 15 Estimated leakage rate from PSVs

Bench Test Leak Description	Leak Categorization	Percent of PRVs Leaking on Bench	Percent of All Leaks	Assumed Leakage (Percent of Capacity)
Leaked between 70 and 90% of set pressure, PRV opened at set pressure	Minor	8.4	50	1
Leakage below 70% of set pressure, PRV opened at set pressure	Moderate	6.6	40	10
Immediate Leakage or PRV leaked too much to open	Spurious Open	2.4	10	25

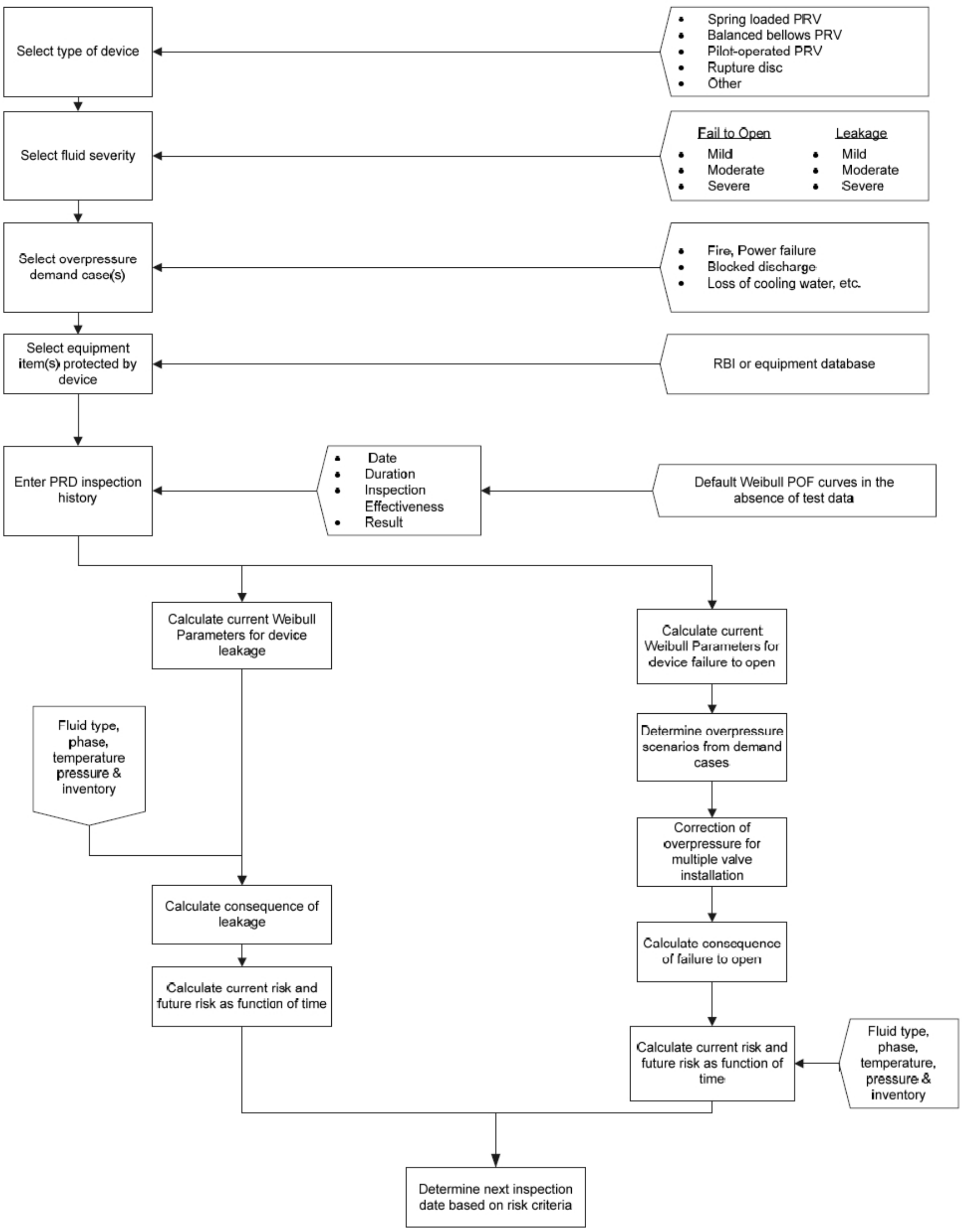

Figure A1: API RBI methodology

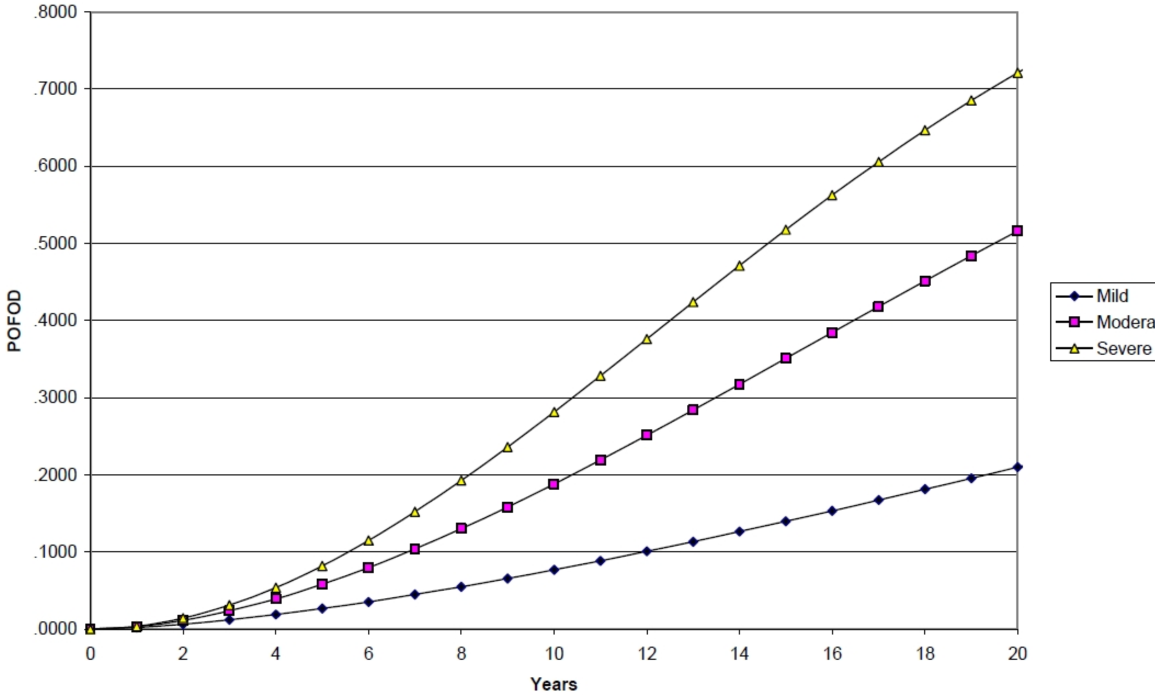

Figure A2 Default conventional PSV failure to open on demand Weibull curves

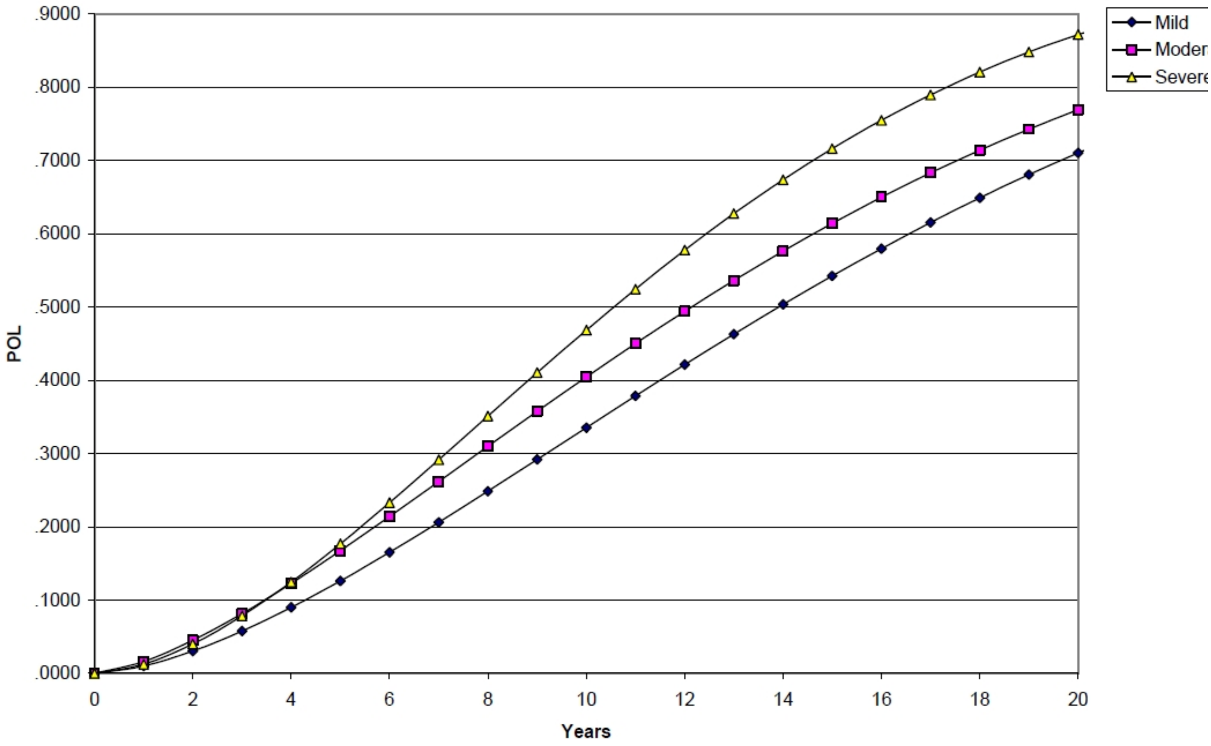

Figure A3: Default leakage failure rate for conventional PSVs

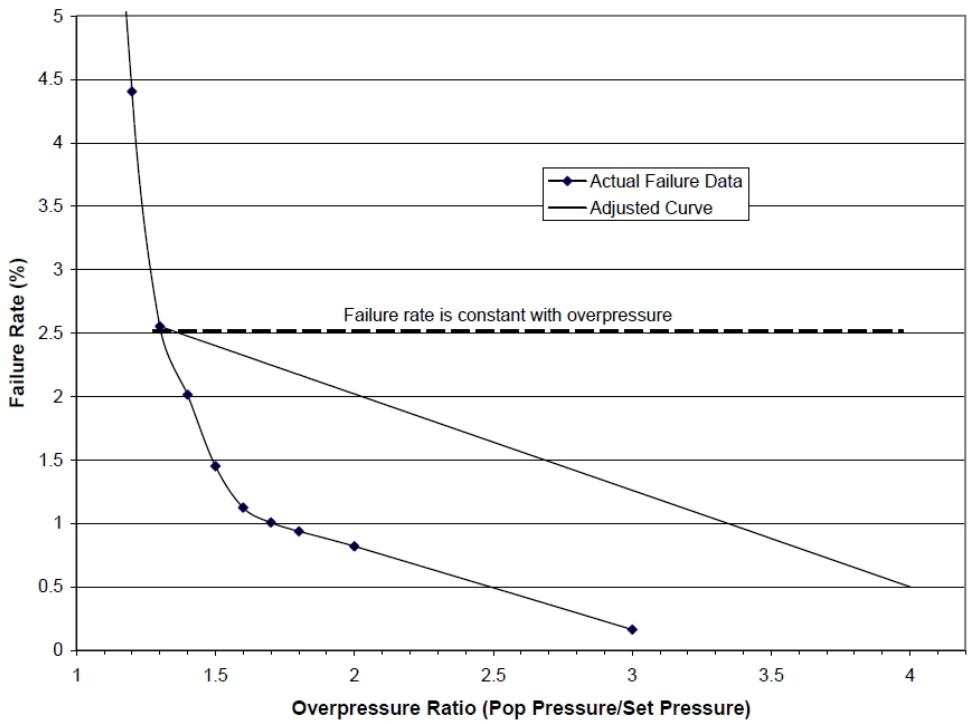

Figure A4: PSV failure rate as a function of overpressure

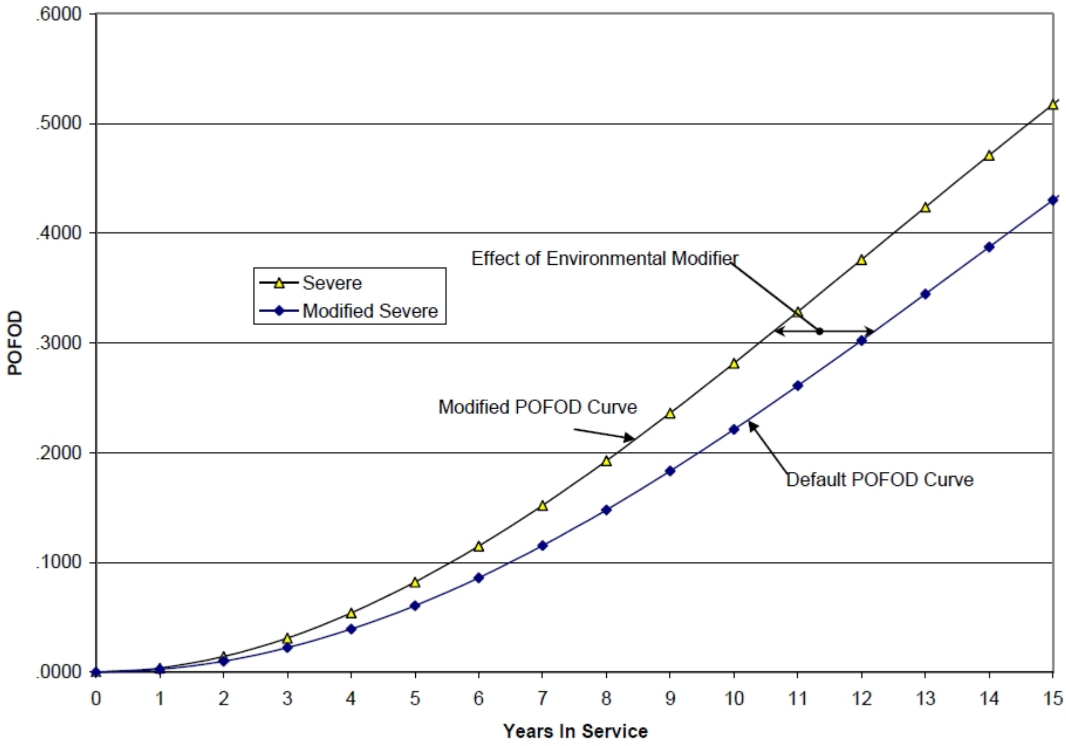

Figure A5: Effect of environmental factors on PSV Weibull curves

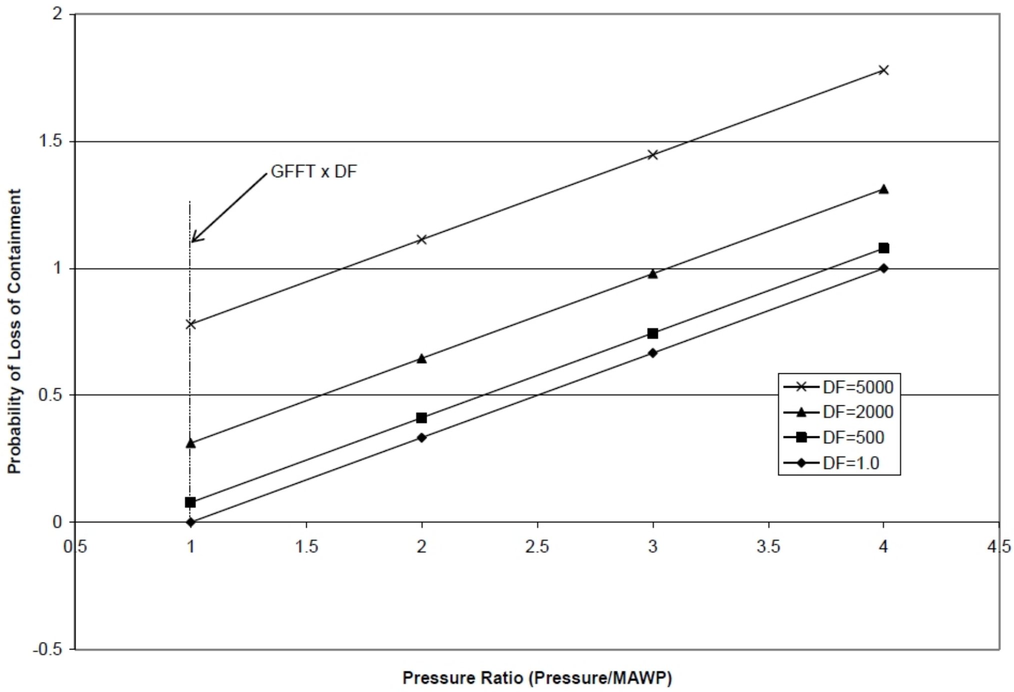

Figure A6: Probability of loss of containment as a function of overpressure

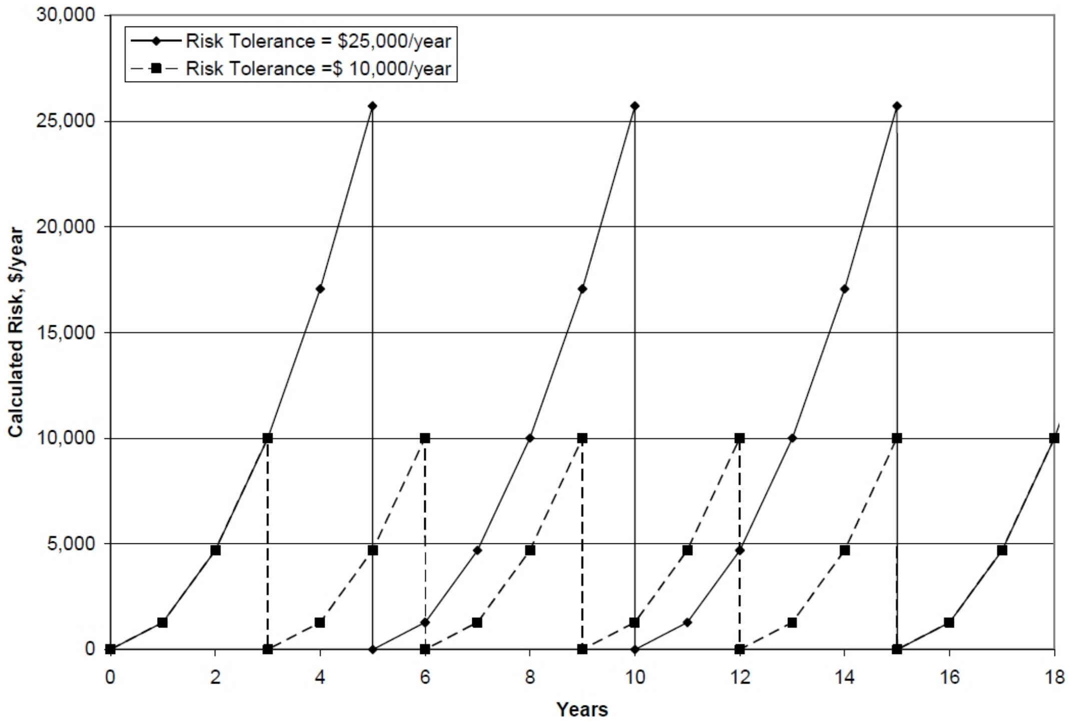

Figure A7 Inspection Test Updating of PSVs

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