API RP 581 PART 5

RISK-BASED INSPECTION METHODOLOGY

FOR SPECIAL EQUIPMENT

PART 5 CONTENTS

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Risk-Based Inspection Methodology
Part 5—Special Equipment

1 SCOPE

1.1 Purpose

This recommended practice, API 581, *Risk-Based Inspection Methodology*, provides semi-quantitative procedures to establish an inspection program using risk-based methods for pressurized fixed equipment including pressure vessel, piping, tankage, pressure-relief devices (PRDs), and heat exchanger tube bundles. API 580, *Risk-Based Inspection* provides guidance for developing risk-based inspection (RBI) programs on fixed equipment in refining, petrochemical, chemical process plants, and oil and gas production facilities. The intent is for API 580 to introduce the principles and present minimum general guidelines for RBI, while the API 581 recommended practice provides semi-quantitative calculation methods to calculate risk and develop an inspection plan.

1.2 Introduction

The calculation of risk outlined in API 581 involves the determination of a probability of failure (POF) combined with the consequence of failure (COF). Failure is defined as a loss of containment from the pressure boundary resulting in leakage to the atmosphere or rupture of a pressurized component. Risk increases as damage accumulates during in-service operation as the risk tolerance or risk target is approached and an inspection is recommended of sufficient effectiveness to better quantify the damage state of the component. The inspection action itself does not reduce the risk; however, it does reduce uncertainty and therefore allows more accurate quantification of the damage present in the component.

1.3 Risk Management

In most situations, once risks have been identified, alternate opportunities are available to reduce them. However, nearly all major commercial losses are the result of a failure to understand or manage risk. In the past, the focus of a risk assessment has been on-site safety-related issues. Presently, there is an increased awareness of the need to assess risk resulting from:

a) on-site risk to employees,

b) off-site risk to the community,

c) business interruption risks, and

d) risk of damage to the environment.

Any combination of these types of risks may be factored into decisions concerning when, where, and how to inspect equipment.

The overall risk of a plant may be managed by focusing inspection efforts on the process equipment with higher risk. API 581 provides a basis for managing risk by making an informed decision on inspection frequency, level of detail, and types of nondestructive examination (NDE). It is a consensus document containing methodology that owner–~~user~~operators may apply to their RBI programs. In most plants, a large percent of the total unit risk will be concentrated in a relatively small percent of the equipment items. These potential higher risk components may require greater attention, perhaps through a revised inspection plan. The cost of the increased inspection effort can sometimes be offset by reducing excessive inspection efforts in the areas identified as having lower risk. Inspection will continue to be conducted as defined in existing working documents, but priorities, scope, and frequencies can be guided by the methodology contained in API 581.

This approach can be made cost-effective by integration with industry initiatives and government regulations, such as Process Safety Management of Highly Hazardous Chemicals (OSHA 29 CFR 1910.119), or the EPA risk management programs for chemical accident release prevention.

2 Normative References

The following referenced documents are indispensable for the application of this document. For dated references, only the edition cited applies. For undated references, the latest edition of the referenced document (including any amendments) applies.

API Recommended Practice 580 Recommended Practice for Risk-Based Inspection, American Petroleum Institute, Washington, D.C.

API Recommended Practice 581, *Risk-Based Inspection Methodology*, *Part 1—Inspection Planning Methodology*, American Petroleum Institute, Washington, DC.

API Recommended Practice 581, *Risk-Based Inspection Methodology*, *Part 2—Probability of Failure Methodology*, American Petroleum Institute, Washington, DC.

API Recommended Practice 581, *Risk-Based Inspection Methodology*, *Part 3—Consequence of Failure Methodology*, American Petroleum Institute, Washington, DC.

3 Pressure Vessels and Piping

3.1 POF

The procedures for POF calculations to be used are provided in Part 2. The POF as a function of time and inspection effectiveness is determined using a GFF, a management systems factor, and DFs for the applicable active damage mechanisms as described in Section 4.1.

3.2 COF

COF calculation procedures for two levels of consequence analysis are provided in Part 3, as described in Section 4.2. In both methods, the consequence analysis may be determined in consequence area or in financial consequence. Consequences from flammable and explosive events, toxic releases, and nonflammable and nontoxic events are considered based on the process fluid and operating conditions.

3.3 Risk Analysis

Risk as a function of time is calculated in accordance with Section 4.3.1. The distribution of risks for different components may be plotted on a risk matrix or iso-risk plot, as described in Section 4.3.2 and Section 4.3.2.3, respectively.

3.4 Inspection Planning Based on Risk Analysis

The procedure to determine an inspection plan is provided in Section **Error! Reference source not found.**. This procedure may be used to determine both the time and type of inspection to be performed based on the process fluid and design conditions, component type and materials of construction, and the active damage mechanisms.

4 Storage Tanks

The calculation of the consequence of a leak or rupture of an API 620 low pressure and API 650 atmospheric storage tanks bottom, edge and course components are covered in this section. The primary liquid container should be evaluated for risk with the secondary container purpose as leak isolation for API 620 double-walled tanks (tank-in-tank systems). The DF and POF calculation use a methodology similar to the approach outlined in Part 2. The methodology for consequence analysis specialized for storage tanks are provided for the COF calculation. The ~~B~~background on the generic failure frequencies for tank bottoms and courses are provided in Part 3, Section 3.A.5.3.1.

4.1 Probability of Failure

POF calculation procedures for storage tank bottom components are provided in this section. Follow calculating procedures outlined in Part 2 for tank course POF. The tank bottom component POF as a function of time and inspection effectiveness is determined using a generic failure frequency, a management systems factor, and DFs for the applicable active damage mechanisms.

The soil-side plates of the tank bottom edge (under-shell) may have a different corrosive environment and foundation conditions than the remainder of the bottom component in tanks with annular rings. Product side corrosion in the perimeter area of the tank may be different than the remainder of the tank bottom due to the as-built or settled profile, edge sump(s), mixers or other appurtenances. In addition, the stresses in the tank bottom edge differs from the tank bottom and the t_{min} calculation in the critical zone are calculated using API 620 and API 650.

4.2 Determination of the Tank Bottom Damage Factor

The calculation procedure for the tank bottom component thinning DF calculation is provided in this section. DFs for other active damage mechanisms are calculated using Part 2, Section 5 through Section 24.

4.2.1 Determination of the Tank Bottom Thinning Damage Factor

a) STEP 1.1 – Determine the furnished thickness, t, and age, age, for the tank component from the installation date. If the tank has an internal liner, determine the liner age, age_{liner} from the liner installation date.

b) STEP 1.2 – Determine the corrosion rate for the base material, $C_{r,bm}$, based on the material of construction and process environment, using guidance from Part 2, Section 4.5.2 and examples in Part 2, Annex 2.B for establishing corrosion rates.

c) STEP 1.3—Determine the inspection effectiveness, N_A^{Thin}, N_B^{Thin}, N_C^{Thin}, and N_D^{Thin}, for the last inspection performed using Part 2, Section 4.5.6 for guidance.

d) STEP 1.4 – Determine the time in service, age_{tk}, since the last inspection known thickness, t_{rdi} where t_{rdi} is the starting thickness with respect to wall loss associated with internal corrosion (see Part 2, Section 4.5.5).

   1) Determine the date of the last inspection with a measured thickness and calculate the service age since the inspection, age_{tk}, and the measured thickness, t_{rdi}. If no measured thickness is available, set $t_{rdi} = t$ and $age_{tk} = age$.

   2) For tank components with internal liners, determine the lining type and age using Table 4.1 or using the remaining life of the internal liner, condition of liner, F_{LC}, at last inspection using Table 4.2, online monitoring factor, F_{OM}, using Equation (5.1). If component does not contain an internal liner, $age_{rc} = 0$.

$$age_{f,rc} = \frac{age_{rc}}{F_{LC}} \cdot F_{OM} \tag{5.1}$$

Online monitoring adjustment factor, F_{OM}—Some lined components have monitoring to allow early detection of a leak or other failure of the lining. The monitoring allows orderly shutdown of the component before failure occurs. If on-line monitoring is used, and it is known to be effective at detecting lining deterioration, F_{OM} = 0.1; otherwise F_{OM} = 1.0. Examples of monitoring systems include thermography or heat sensitive paint (refractory linings), weep holes with detection devices (loose alloy linings), and electrical resistance detection (glass linings).

e) STEP 1.5 – Determine $t_{\min}$ using one of the following methods:

1) For the API STD 620 and API STD 650 tank courses, determine the allowable stress, S, weld joint efficiency, E, and calculate the minimum required thickness, $t_{\min}$, using component type in Part 2, Table 4.2, geometry type in Part 2, Table 4.3 and per the original construction code or API 579-1/ASME FFS-1 [1] or API STD 620, as applicable.
2) API STD 650 Tank bottoms can be modeled with two components. If the component type is Tank650 TANKBOTTOM, use $t_{\min} = 0.1\ in$ if the storage tank does not have a release prevention barrier or $t_{\min} = 0.05\ in$ if the storage tank has a release prevention barrier, in accordance with API STD 653 [2]. If the component is a Tank650 TANKBOTEDGE, use the minimum thickness for an annular ring or the critical zone (for tanks without annular rings), whichever is applicable, in accordance with API STD 653.
3) API STD 620 Tank bottom $t_{\min}$ is determined by using API STD 620. If the component is a Tank620 TANKBOTEDGE, use the minimum thickness for an annular ring or the critical zone (for tanks without annular rings), whichever is applicable, in accordance with API STD 653.
4) A specific $t_{\min}$ calculated by another method and documented in the asset management program may be used at the owner-~~user~~operator's discretion.

f) STEP 1.6 - Determine the tank bottom component A_{rt} parameter using Equation (5.2) based on t from STEP 1, $C_{r,bm}$ from STEP 1.2, age_{tk} and t_{rdi} from STEP 1.4.

NOTE: the age parameter in these equations is equal to age_{tk} from STEP 1.4.

1) For tank courses, go to STEPs 7 through 15 in Part 2, Section 4.5.7 and skip to STEP 1.8.
2) For tank bottom components, calculate the A_{rt} parameter using Equation (5.2).

$$A_{rt} = max\left[\left(1 - \frac{t_{rdi} - (C_{r,bm} \cdot (age_{tk} - age_{f,rc}))}{t_{min} + CA}\right),\ \ 0.0\right] \tag{5.2}$$

g) STEP 1.7 – For tank bottom components, determine the base damage factor for thinning, D_{fB}^{thin}, using Table 4.3 and based on the A_{rt} parameter from STEP 1.6 and inspection effectiveness from STEP 1.3.

h) STEP 1.8 – Determine the DF for thinning, $D_f^{Tank,Thin}$, using Equation (5.3).

$$D_f^{AST,Thin} = max\left[\left(D_{fB}^{Thin} \cdot F_{WD} \cdot F_{AM} \cdot F_{SM}\right),\ \ 0.1\right] \tag{5.3}$$

The adjustment factors in are determined as described below.

1) Adjustment for Welded Construction, F_{WD} – If the component is welded (i.e. not riveted), then $F_{WD} = 1$; otherwise, $F_{WD} = 10$.

2) Adjustment for Maintenance in Accordance with API STD 653, F_{AM} – If the storage tank is maintained in accordance with API STD 653, then $F_{AM} = 1$; otherwise, $F_{AM} = 5$.

3) Adjustment for Settlement, F_{SM} – It is determined based on the following criteria:

   - Recorded settlement exceeds API STD 653 criteria – $F_{SM} = 2$
   - Recorded settlement meets API STD 653 criteria – $F_{SM} = 1$
   - Settlement never evaluated – $F_{SM} = 1.5$
   - Concrete foundation, no settlement – $F_{SM} = 1$

4.2.2 Determination of the SCC Damage Factors

Follow calculating procedures outlined in Part 2, Section 5 through Section 14 for SCC of storage tank courses, if applicable.

4.2.3 Determination of the External Damage Factors

Follow calculating procedures outlined in Part 2, Section 15 through Section 18 for external damage of storage tank courses, if applicable.

4.2.4 Determination of the Brittle Fracture Damage Factors

Follow calculating procedures outlined in Part 2, Section 21 for brittle fracture of storage tank courses, if applicable.

4.2.5 Damage Factor Combination for Multiple Damage Mechanisms

Follow calculating procedures outlined in Part 2, Section 3.4.2 for combining DFs or multiple damage mechanisms of storage tank courses.

4.3 Consequence of Failure

The COF is calculated in terms of affected area or in financial consequence. Consequences from flammable and explosive events, toxic releases, and nonflammable/nontoxic events are considered in both methods based on the process fluid and operating conditions. Financial consequences from component damage, product loss, financial impact, and environmental penalties are considered.

The COF methodology is performed to aid in establishing a ranking of equipment items on the basis of risk. The consequence measures are intended to be used for establishing priorities for inspection programs. Methodologies for two levels of analysis are provided. A special COF methodology is provided for low pressure and atmospheric storage tanks and is covered in this section.

4.4 Consequence of Failure Methodology for Storage Tank Courses

The COF associated with storage tanks is concerned primarily with the financial losses due to leakage and/or rupture of a storage tank course. Safety/area based consequences are addressed for the courses following the Level 1 or Level 2 consequence analysis methods provided in Part 3, Section 4.0 or Section 5.0. Detailed procedures for calculating the financial COF for courses are provided in Section 2.5 through Section 2.16.

The procedure for determining COF of storage tank course components includes calculations for both area and financial-based methods.

4.5 Required Properties at Storage Conditions

Fluid properties should be determined for the COF calculation. When calculating the safety COF area for tank courses, see Part 3, Section 5.1.2 Level 1 or 2 Consequence of Failure methodology. See Part 3, Section 5.1.2 for detailed description of required properties at storage conditions. The financial COF for fluids other than those in Table 4.5 may be modeled if the stored as liquid data required in Table 4.5 and Part 3, Table 4.2 are provided by the user.

NOTE: ~~T~~the flammable COF would be calculated based on the equation constants in Part 3, Table 4.8 and Part 3, Table 4.9 for the fluid closest matching the molecular weight (MW) and normal boiling point (NBP).

Fluid properties at storage conditions are necessary to calculate the financial and area-based Level 1 and Level 2 COF. Refer to the following paragraphs for a detailed description of the required properties at storage conditions for tank course components:

a) Level 1 COF methodology, see Part 3, Section 4.1.2

b) Level 2 COF methodology see Part 3, Section 5.1.2

4.5.1 Required Properties at Flashed Conditions

Fluid properties are determined for a safety based COF for use in the Level 1 or 2 Consequence of Failure methodology. See Part 3, Section 5.1.3 for detailed description of required properties at flashed conditions.

4.6 Release Hole Size Selection

A discrete set of release events or release hole sizes are used for consequence analysis as outlined in Table 4.4.

4.6.1 Calculation of Release Hole Sizes

The following procedure may be used to determine the release hole size and the associated generic failure frequencies.

a) STEP 2.1 – Determine the release hole size, d_n, from Table 4.4 for storage tank courses.

b) STEP 2.2 – Determine the generic failure frequency, gff_n, for the d_n release hole size and the total generic failure frequency from Part 2, Table 3.1 or from Equation (5.4).

$$gff_{tot} = \sum_{n=1}^{4} gff_n \tag{5.4}$$

4.7 Release Rate Calculation

Release rate calculations are provided for a leak in a storage tank course. The liquid head of the product is assumed to be constant over time, and the leak is to atmospheric pressure for a course leak.

4.7.1 Storage Tank Course

The discharge of a liquid through a sharp-edged orifice in a storage tank course with a liquid height above the orifices may be calculated using Equation (5.5).

$$W_n = C_{32} \cdot C_d \cdot A_n \sqrt{2 \cdot g \cdot LHT_{above,i}} \tag{5.5}$$

In Equation (5.5), the discharge coefficient, C_d , for fully turbulent liquid flow from sharp-edged orifices is in the range of $0.60 \le C_d \le 0.65$. A value of $C_d = 0.61$ is recommended.

4.7.2 Calculation of Storage Tank Course Release Rate

a) STEP 3.1 –Determine the height of the liquid, h_{liq} , above the release hole size, d_n for each hole size.

b) STEP 3.2 – Determine the hole area, A_n , for each hole size using Equation (5.6).

$$A_n = \frac{\pi d_n^2}{4} \tag{5.6}$$

c) STEP 3.3 - Determine the liquid height above the i^{th} course where h_{liq} is the maximum fill height in the tank and CHT is the height of each course.

$$LHT_{above,i} = \left[h_{liq} - (i-1) \cdot CHT \right] \tag{5.7}$$

d) STEP 3.3 – Determine the flow rate, W_n , for each hole size using Equation (5.5) based on h_{liq} from STEP 3.1 and A_n from STEP 3.2.

4.8 Estimate the Inventory Volume and Mass Available for Release

The inventory in the storage tank available for release depends on the component being evaluated. The available inventory for courses is a function of the location of the release hole and is calculated as the volume of fluid above the release hole.

4.8.1 Calculation of Storage Tank Course Inventory Mass

The amount of fluid inventory used in the course consequence analysis is the amount of fluid that is above the lower elevation of the course under evaluation.

a) STEP 4.1 – Determine the liquid height above the i^{th} course where h_{liq} is the maximum fill height in the tank and CHT is the height of each course.

$$LHT_{above,i} = \left[h_{liq} - (i-1) \cdot CHT \right] \tag{5.8}$$

b) STEP 4.2 – Determine the volume above the course being evaluated.

$$Lvol_{above,i} = \left(\frac{\pi D_{tank}^2}{4} \right) \cdot LHT_{above,i} \tag{5.9}$$

c) STEP 4.3 – Calculate the available volume of the release.
NOTE: the release hole should be assumed to be at the bottom of the course.

$$Lvol_{avail,n} = Lvol_{above,i} \tag{5.10}$$

d) STEP 4.4 – Calculate the storage tank volume in barrels using Equation (5.11).

$$Bbl_{avail,n} = Lvol_{avail,n} \cdot C_{13} \tag{5.11}$$

e) STEP 4.5 – Calculate the storage tank mass using liquid density, ρ_l, from Table 4.5 and using Equation (5.12).

$$mass_{avail,n} = Lvol_{avail,n} \cdot \rho_l \tag{5.12}$$

4.9 Determine the Type of Release

The type of release for the storage tank is assumed to be continuous.

4.10 Estimate the Impact of Detection and Isolation Systems on Release Magnitude

Detection and isolation systems are not accounted for in the storage tank course consequence analysis.

4.11 Determine the Release Rate and Volume for the Consequence of Failure Analysis

The storage tank course release is assumed to be continuous and the release rate is calculated from Equation (5.13) where W_n is determined in Section 2.7.2.

$$rate_n = W_n \tag{5.13}$$

4.11.1 Calculation for Storage Tank Course Release Volume

A step-by-step methodology for determining the release rate and volume is in accordance with the modeling in Part 3, Section 4 for Level 1 COF and Part 3, Section 5 for Level 2 COF with the following differences:

- The pool fire area should not exceed the area of the dike.
- The release volume should be calculated with the following steps.

a) STEP 5.1 – Determine the release rate, $rate_n$, for each hole size in bbls/day using Equation (5.13) where the release rate, W_n, is from STEP 3.3.

b) STEP 5.2 – Determine the leak detection time, t_{ld}, as follows:

$$t_{ld} = 7\ days\ \ for\ d_n \le 3.17mm\ [0.125\ in]$$, or

$$t_{ld} = 1\ days\ \ for\ d_n > 3.17mm\ [0.125\ in]$$

c) STEP 5.3 – Calculate the leak duration, ld_n, of the release for each hole size using Equation (5.14) based on the release rate, $rate_n$, from STEP 5.1, the leak detection time, t_{ld}, from STEP 5.2, and the storage tank volume, $Bbl_{avail,n}$, from STEP 4.4.

$$ld_n = min\left[\left\{\frac{Bbl_{avail,n}}{rate_n}\right\},\ \ 7\ days\right] \qquad\qquad for\ d_n \le 3.17mm\ [0.125\ in] \tag{5.14}$$

d) STEP 5.4 – Calculate the release volume from leakage, Bbl_n^{leak}, for each hole size using Equation (5.15) based on the release rate, $rate_n$, from STEP 5.1, the leak duration, ld_n, from STEP 5.3, available volume, $Bbl_{avail,n}$, from STEP 4.4.

$$Bbl_n^{leak} = min\left[\left\{rate_n \cdot ld_n\right\},\ \ Bbl_{avail,n}\right] \tag{5.15}$$

e) STEP 5.5 – Calculate the release mass from leakage, $mass_n^{leak}$, for each hole size using Equation (5.16) based on the available volume, Bbl_n^{leak} , from STEP 5.4.

$$mass_n^{leak} = Bbl_n^{leak} \tag{5.16}$$

f) STEP 5.6 – Calculate the release volume from a rupture, $Bbl_n^{rupture}$, for each hole size using Equation (5.17) based on the available volume, $Bbl_{avail,n}$, from STEP 4.4.

$$Bbl_n^{rupture} = Bbl_{avail,n} \tag{5.17}$$

g) STEP 5.7 – Calculate the mass from a rupture, $mass_n^{rupture}$, for each hole size using Equation (5.18) based on the available volume, $Bbl_n^{rupture}$, from STEP 5.6.

$$mass_n^{rupture} = Bbl_n^{rupture} \tag{5.18}$$

4.12 Determine Flammable and Explosive Consequences for Storage Tank Courses

Flammable and explosive consequences for storage tanks courses are determined using a similar approach as implemented for Level 1 and 2 consequence analysis.

4.12.1 Calculation of Flammable and Explosive Consequences

The step-by-step procedure for determining the flammable and explosive consequences are in accordance with the level of consequence analysis, see Part 3, Section 4.8 for Level 1 analysis and Part 3, Section 5.8.9 for Level 2 COF analysis.

4.13 Determine Toxic Consequences for Storage Tank Courses

Toxic consequences for storage tank courses are determined using a similar approach as implemented for Level 1 and 2 consequence analysis.

4.13.1 Calculation of Toxic Consequences for Storage Tank Courses

The step-by-step methodology for determining the toxic consequences are in accordance with the Level 1 and 2 consequence analysis; see Part 3, Section 4.9 and Part 3, Section 5.9.8.

4.14 Determine Non-Flammable, Non-Toxic Consequences

Non-flammable, non-toxic consequences are not determined for storage tanks.

4.15 Determine Component Damage and Personnel Injury Consequences for Storage Tank Courses

Flammable and explosive consequences for storage tank courses are determined using a similar approach as implemented for Level 1 and 2 consequence analysis.

4.15.1 Calculation for Component Damage and Personnel Injury Consequences

The step-by-step procedure for determining the flammable and explosive consequences are in accordance with the Level 1 COF Part 3, Section 4.8 and Level 2 COF in Part 3, Section 5.11.5.

4.16 Determine the Financial Consequences

The financial consequence is determined in accordance with the Level 1 COF in Part 3, Section 4.12.

4.16.1 Calculation of Storage Tank Course Financial Consequence

The step-by-step procedure for estimating the financial consequence is in accordance with Section 4.12.7, except when calculating the environmental financial consequence. The storage tank course financial consequence can be calculated following the approach in Sections defined below using the hole sizes defined in Table 4.8.

- Component Damage Cost in accordance to Section 4.12.2
- Damage cost to surrounding equipment in accordance with Section 4.12.3
- Business interruption costs in accordance to Section 4.12.4
- Potential Injury costs in accordance to Section 4.12.5

The storage tank Environmental financial consequence for courses is calculated following the steps provided below.

a) STEP 6.1 – Determine the following parameters.

   P_{lvdike} – percentage of fluid leaving the dike

   P_{onsite} – percentage of fluid that leaves the dike area but remains on-site

   $P_{offsite}$ – percentage of fluid that leaves the dike area but does not enter nearby water

b) STEP 6.2 – Determine the environmental sensitivity used to establish C_{indike}, $C_{ss-onsite}$, $C_{ss-offsite}$, and C_{water} from Table 4.6.

c) STEP 6.3 – Determine the probability weighted total barrels of fluid released by leakage, $Bbl_{released}$.

$$Bbl_{release}^{leak} = \frac{\sum_{n=1}^{3}\left(Bbl_n^{leak} \cdot gff_n\right)}{gff_{tot}} \tag{5.19}$$

d) STEP 6.4 – Calculate the total barrels of fluid within the dike from leakage, Bbl_{indike}^{leak}, the total barrels of fluid in the on-site surface soil, $Bbl_{ss-onsite}^{leak}$, the total barrels of fluid in the off-site surface soil, $Bbl_{ss-offsite}^{leak}$, and the total barrels of fluid in that reach water, Bbl_{water}^{leak}, using Equation (5.20) through Equation (5.23), respectively.

$$Bbl_{indike}^{leak} = Bbl_{release}^{leak}\left(1 - \frac{P_{lvdike}}{100}\right) \tag{5.20}$$

$$Bbl_{ss-onsite}^{leak} = \frac{P_{onsite}}{100}\left(Bbl_{release}^{leak} - Bbl_{indike}^{leak}\right) \tag{5.21}$$

$$Bbl_{ss-offsite}^{leak} = \frac{P_{offsite}}{100}\left(Bbl_{release}^{leak} - Bbl_{indike}^{leak} - Bbl_{ss-onsite}^{leak}\right) \tag{5.22}$$

$$Bbl_{water}^{leak} = Bbl_{release}^{leak} - \left(Bbl_{indike}^{leak} + Bbl_{ss-onsite}^{leak} + Bbl_{ss-offsite}^{leak}\right) \tag{5.23}$$

e) STEP 6.5 – Calculate the financial environmental cost from leakage, $FC_{environ}^{leakage}$.

$$FC_{environ}^{leak} = Bbl_{indike}^{leak} \cdot C_{indike} + Bbl_{ss-onsite}^{leak} \cdot C_{ss-onite} + Bbl_{ss-offsite}^{leak} \cdot C_{ss-offite} + Bbl_{water}^{leak} \cdot C_{water} \tag{5.24}$$

f) STEP 6.6 – Determine the total barrels of fluid released by a course rupture, $Bbl_{release}^{rupture}$.

$$Bbl_{release}^{rupture} = \frac{Bbl_n^{rupture} \cdot gff_4}{gff_{tot}} \tag{5.25}$$

g) STEP 6.7 – Calculate the total barrels of fluid within the dike from a rupture, $Bbl_{indike}^{rupture}$, the total barrels of fluid in the on-site surface soil that, $Bbl_{ss-onsite}^{rupture}$, the total barrels of fluid in the off-site surface soil that, $Bbl_{ss-offsite}^{rupture}$, and the total barrels of fluid that reach water, Bbl_{water}^{leak}, using Equation (5.26) through Equation (5.29), respectively.

$$Bbl_{indike}^{rupture} = Bbl_{release}^{rupture}\left(1 - \frac{P_{lvdike}}{100}\right) \tag{5.26}$$

$$Bbl_{ss-onsite}^{rupture} = \frac{P_{onsite}}{100}\left(Bbl_{release}^{rupture} - Bbl_{indike}^{rupture}\right) \tag{5.27}$$

$$Bbl_{ss-offsite}^{rupture} = \frac{P_{offsite}}{100}\left(Bbl_{release}^{rupture} - Bbl_{indike}^{rupture} - Bbl_{ss-onsite}^{rupture}\right) \tag{5.28}$$

$$Bbl_{water}^{rupture} = Bbl_{release}^{rupture} - \left(Bbl_{indike}^{rupture} + Bbl_{ss-onsite}^{rupture} + Bbl_{ss-offsite}^{rupture}\right) \tag{5.29}$$

h) STEP 6.8 – Calculate the financial environmental cost for a course rupture, $FC_{environ}^{rupture}$.

$$FC_{environ}^{rupture} = Bbl_{indike}^{rupture} \cdot C_{indike} + Bbl_{ss-onsite}^{rupture} \cdot C_{ss-onite} + Bbl_{ss-offsite}^{rupture} \cdot C_{ss-offite} + Bbl_{water}^{rupture} \cdot C_{water} \tag{5.30}$$

i) STEP 6.9 – Calculate the total financial environmental cost from a leak and a rupture, $FC_{environ}$, where $FC_{environ}^{leak}$ is from STEP 12.5 and $FC_{environ}^{rupture}$ is from STEP 12.8.

$$FC_{environ} = FC_{environ}^{leak} + FC_{environ}^{rupture} \tag{5.31}$$

j) STEP 6.10 – Calculate the total financial COF, FC_{total}, using Equation (5.32).

$$FC_{total} = FC_{environ} + FC_{cmd} + FC_{prod} + FC_{affa} + FC_{inj} \tag{5.32}$$

4.17 Determination of Safety Consequences

Safety consequences, SC_f, for storage tank courses are calculated the approach outlined in Part 3, Section 5.13. The injury area, CA_{inj}, for a course release is outlined in Section 3.15.1.

4.18 Consequence of Failure Methodology for Storage Tank Bottoms

The COF associated with storage tanks is concerned primarily with the financial losses due to loss of containment and leakage through the storage tank bottoms. Area based consequences are not calculated for storage tank bottoms Detailed procedures for calculating the financial COF for bottom plates are provided in this section.

The procedure for determining the COF for storage tank bottom components consists of calculations for financial COF based on environmental consequences, component damage cost and business interruption cost. storage tank consequence analysis for flammable and/or explosive or toxic are not calculated for storage tank bottoms.

4.18.1 Required Properties at Storage Conditions

The tank bottom financial COF is calculated using one of the following approaches:

a) Select the representative fluid from Table 4.5 that most closely matches the stored fluid

b) Determine the dynamic viscosity and density of the stored fluid

4.18.2 Hydraulic Conductivity for Storage Tank Bottom

The amount of and rate of leakage from storage tank bottoms is dependent on the type of soil and its properties as well as whether or not the storage tank bottom has a release prevention barrier (RBP). A list of soil types and properties used in the storage tank consequence analysis routine is shown in Table 4.7

The fundamental soil property required in the analysis is the soil hydraulic conductivity, k_h. The hydraulic conductivity as a function of soil type is provided in Table 4.7 based on water. The hydraulic conductivity for other fluids can be estimated based on the hydraulic conductivity, density, and dynamic viscosity of water, denoted as $k_{h,water}$, ρ_w, and μ_w, respectively, and the density and dynamic viscosity of the actual fluid using Equation (5.33).

$$k_{h,prod} = k_{h,water}\left(\frac{\rho_l}{\rho_w}\right)\left(\frac{\mu_w}{\mu_l}\right) \tag{5.33}$$

4.18.3 Fluid Seepage Velocity for Storage Tank Bottom

The seepage velocity of the fluid in the storage tank bottom or product through the soil is given by Equation (5.34) where k_h is the soil hydraulic conductivity and p_s is the soil porosity.

$$vel_{s,prod} = \frac{k_{h,prod}}{p_s} \tag{5.34}$$

4.18.4 Calculation of Fluid Seepage Velocity for Storage Tank Bottom

a) STEP 7.1 – Determine properties including density, ρ_l, and dynamic viscosity, μ_l, of the stored fluid. If a Level 1 analysis is being performed, select the representative fluid properties from Table 4.5.

b) STEP 7.2 – Calculate the hydraulic conductivity for water by averaging the upper and lower bound hydraulic conductivities provided in Table 4.7 for the soil type selected using Equation (5.35).

$$k_{h,water} = C_{31}\frac{\left(k_{h,water-lb} + k_{h,water-ub}\right)}{2} \tag{5.35}$$

c) STEP 7.3 – Calculate the fluid hydraulic conductivity, $k_{h,prod}$, for the fluid stored in the storage tank using Equation (5.33) based on the density, ρ_l, and dynamic viscosity, μ_l, from STEP 7.1 and the hydraulic conductivity for water, $k_{h,water}$, from STEP 7.2.

d) STEP 7.4 – Calculate the product seepage velocity, $vel_{s,prod}$, for the fluid stored in the storage tank using Equation (5.34) based on fluid hydraulic conductivity, $k_{h,prod}$, from STEP 7.3 and the soil porosity provided in Table 4.7.

4.19 Release Hole Size Selection

A discrete set of release events or release hole sizes are used for consequence analysis as outlined in Table 4.8.

4.19.1 Calculation of Release Hole Sizes

The following procedure may be used to determine the release hole size and the associated generic failure frequencies.

a) STEP 8.1 – Determine the release hole size, d_n, from Table 4.8 for storage tank bottoms.

b) STEP 8.2 – Determine the generic failure frequency, gff_n, for the d_n release hole size and the total generic failure frequency from Part 2, Table 3.1 or from Equation (5.36).

$$gff_{tot} = \sum_{n=1}^{4} gff_n \qquad (5.36)$$

4.20 Release Rate Calculation

Release rate calculations are provided for a leak in a storage tank bottom plate. The liquid head is assumed to be constant in time, and the leak is into the ground that is modeled as a continuous porous media approximated by soil properties typically used for storage tank foundations.

4.20.1 Storage Tank Bottom Release Rate

The product leakage flow rate through a small hole in the storage tank bottom is a function of the soil and fluid properties as well as the liquid head (fill height) above the bottom. The flow rate equations can be found in Rowe [3]. The flow rate through a storage tank bottom into a porous media is calculated using the Bernoulli in Equation (5.37), Giroud in Equation (5.38), or Equation (5.39) based on the hydraulic conductivity, $k_{h,prod}$, and release hole size, d_n.

$$W_n = C_{33} \cdot \pi \cdot d_n^2 \sqrt{2 \cdot g \cdot h_{liq}} \cdot n_{rh,n} \qquad for\ k_{h,prod} > C_{34} \cdot d_n^2 \qquad (5.37)$$

$$W_n = C_{35} \cdot C_{qo} \cdot d_n^{0.2} \cdot h_{liq}^{0.9} \cdot k_{h,prod}^{0.74} \cdot n_{rh,n} \qquad for\ k_{h,prod} \le C_{37} \cdot \left[\frac{d_n^{1.8}}{C_{qo} \cdot h_{liq}^{0.4}} \right]^{\frac{1}{0.74}} \qquad (5.38)$$

$$W_n = C_{38} \cdot 10^{2 \cdot \log(d_n) + 0.5 \cdot \log(h_{liq}) - 0.74 \cdot \left(\frac{C_{39} + 2 \cdot \log(d_n) - \log(k_{h,prod})}{m} \right)^m} \cdot n_{rh,n} \qquad for\ all\ other\ cases \qquad (5.39)$$

$$Where\ m = C_{40} - 0.4324 \cdot \log(d_n) + 0.5405 \cdot \log(h_{liq})$$

In Equation (5.38), the parameter C_{qo} is an adjustment factor for degree of contact with soil and ranges from $C_{qo} = 0.21$ for good contact to $C_{qo} = 1.15$ for poor contact. A value of $C_{qo} = 0.21$ is recommended in the consequence analysis.

If the storage tank bottom has a release prevention barrier (RBP), then the liquid height, h_{liq}, to be used in the flow rate calculations is set to .0762 m (0.25 ft). If the storage tank does not have a release prevention barrier, the liquid height, h_{liq}, to be used in the flow rate calculations is the actual height of the stored product.

The number of release holes, $n_{rh,n}$, for each release hole size is a function of the storage tank diameter and is shown in Table 4.9.

4.20.2 Calculation for Storage Tank Bottom Release Hole Size

a) STEP 9.1 – For each release hole size, determine the number of release holes, $n_{rh,n}$, from Table 4.9.

b) STEP 9.2 – Determine the hole area, A_n, for each hole size from STEP 8.1 using Equation (5.6).

c) STEP 9.3 – Determine the hydraulic conductivity of the stored liquid, $k_{h,prod}$, from STEP 1.4.

d) STEP 9.4 – For each release hole size, determine the flow rate, W_n, using Equation (5.37), Equation (5.38), or Equation (5.39), as applicable. The liquid height, h_{liq}, to use in this calculation is determined as follows:

   1) The storage tank has an RPB: $h_{liq} = 0.25\ ft\ \left(0.0762\ m\right)$

   2) The storage tank does not have an RPB: $h_{liq} = Actual\ \ Product\ \ Height$

4.21 Inventory Volume and Mass Available for Release

The amount of inventory in the storage tank available for release depends on the component being evaluated. The available inventory is the entire contents of the storage tank for bottom components unless the tank has an RPB.

4.21.1 Calculation of Storage Tank Bottom Inventory Mass

The amount of fluid available for release through storage tank bottoms is the fluid level up to the storage tank design fill height or the operating fill height.

a) STEP 10.1 – Calculate liquid volume in the storage tank in m^3 (ft^3) using Equation (5.40).

$$Lvol_{total} = \left(\frac{\pi D_{tank}^2}{4}\right) \cdot h_{liq} \tag{5.40}$$

b) STEP 10.2 – Calculate the total storage tank volume in barrels using Equation (5.41).

$$Bbl_{total} = Lvol_{total} \cdot C_{13} \tag{5.41}$$

c) STEP 10.3 – Calculate the storage tank mass using Equation (5.42).

$$mass_{total} = Lvol_{total} \cdot \rho_l \tag{5.42}$$

4.22 Type of Release

The type of release for the storage tank bottom is assumed to be continuous.

4.23 Impact of Detection and Isolation Systems on Release Magnitude

Detection and isolation systems are not accounted for in the storage tank consequence analysis.

4.24 Release Rate and Volume for the Consequence of Failure Analysis

The release for the storage tank is assumed to be continuous, and the release rate is calculated from Equation (5.43) where W_n is determined in STEP 9.4.

$$rate_n = W_n \tag{5.43}$$

4.24.1 Storage Tank Bottom Release Volume

A step-by-step procedure for determining the release rate and volume is as follows:

a) STEP 11.1 – Determine the release rate, $rate_n$, for each release hole size using Equation (5.43) where the release rate, W_n, is from STEP 9.4.

b) STEP 11.2 – Determine the leak detection time, t_{ld}, as follows:

1) $t_{ld} = 7\ days$ for a storage tank on a concrete or asphalt foundation, or

2) $t_{ld} = 30\ days$ for a storage tank with an RPB, or

3) $t_{ld} = 360\ days$ for a storage tank without an RPB.

c) STEP 11.3 – Calculate the leak duration, ld_n, for each release hole size using Equation (5.44) based on the release rate, $rate_n$, from STEP 11.1, the leak detection time, t_{ld}, from STEP 11.2, and the total volume, Bbl_{total}, from STEP 10.2

$$ld_n = min\left[\left\{\frac{Bbl_{total}}{rate_n}\right\},\ t_{ld}\right] \tag{5.44}$$

d) STEP 11.4 – Calculate the release volume from leakage, Bbl_n^{leak}, for each release hole size using Equation (5.45) based on the release rate, $rate_n$, from STEP 11.1, the leak duration, ld_n, from STEP 11.3, and the total volume, Bbl_{total}, from STEP 10.2.

$$Bbl_n^{leak} = min\left[\left\{rate_n \cdot ld_n\right\},\ \ Bbl_{total}\right] \tag{5.45}$$

e) STEP 11.5 – Calculate the release volume from a rupture, $Bbl_n^{rupture}$, for each release hole size using Equation (5.46) based on the total volume, Bbl_{total}, from STEP 10.2.

$$Bbl_n^{rupture} = Bbl_{total} \tag{5.46}$$

4.25 Determine the Financial Consequences

The step-by-step procedure for estimating the financial consequence is in accordance with Section 4.12.7. The financial consequences for the storage tank bottom are calculated with the steps provided below:

- Damage cost to surrounding equipment in accordance with Section 4.12.3 is not applicable for storage tank bottom component

- Business interruption costs in accordance to Section 4.12.4
- Potential Injury costs in accordance to Section 4.12.5 is not applicable for storage tank bottom component

4.25.1 Calculation of Storage Tank Bottom Financial Consequence

The step-by-step procedure for determining Financial COF is as follows:

a) STEP 12.1 – Determine the following parameters:

1) P_{lvdike} – percentage of fluid leaving the dike
2) $P_{lvdike-onsite}$ – percentage of fluid that leaves the dike area but remains on-site
3) $P_{lvdike-offsite}$ – percentage of fluid that leaves the site area, but does not enter nearby water
4) The storage tank Environmental financial consequence for the bottom can be calculated following the steps provided below.

b) STEP 12.2 – Determine the environmental sensitivity to establish C_{indike}, $C_{ss-onsite}$, $C_{ss-offsite}$, C_{water}, $C_{subsoil}$, and $C_{groundwater}$ from Table 4.6.

c) STEP 12.3 – Determine the seepage velocity of the product, vel_{s-prod}, using Equation (5.34).

d) STEP 12.4 – Determine the total distance to the ground water underneath the storage tank, s_{gw}, and the time to initiate leakage to the ground water, t_{gl}.

$$t_{gl} = \frac{s_{gw}}{vel_{s,prod}} \tag{5.47}$$

e) STEP 12.5 – Determine the volume of the product for each hole size in the subsoil and ground water where the leak detection time, t_{ld}, is determined in STEP 11.2.

$$Bbl_{groundwater,n}^{leak} = Bbl_{n}^{leak}\left(\frac{t_{ld}-t_{gl}}{t_{ld}}\right) \qquad for\ t_{gl} < t_{ld} \tag{5.48}$$

$$Bbl_{groundwater,n}^{leak} = 0 \qquad for\ t_{gl} \geq t_{ld} \tag{5.49}$$

$$Bbl_{subsoil,n}^{leak} = Bbl_{n}^{leak} - Bbl_{groundwater,n}^{leak} \tag{5.50}$$

f) STEP 12.6 – Determine the environmental financial consequence of a leak, $FC_{environ}^{leak}$, for each hole size.

$$FC_{environ}^{leak} = \frac{\sum_{n=1}^{3}\left(Bbl_{groundwater,n}^{leak} \cdot C_{groundwater} + Bbl_{subsoil,n}^{leak} \cdot C_{subsoil}\right) gff_{n}}{gff_{tot}} \tag{5.51}$$

g) STEP 12.7 – Determine the total barrels of fluid released by a storage tank bottom rupture, $Bbl_{release}^{rupture}$.

$$Bbl_{release}^{rupture} = \frac{Bbl_{total} \cdot gff_{4}}{gff_{tot}} \tag{5.52}$$

h) STEP 12.8 – Calculate the total barrels of fluid within the dike from a rupture, $Bbl_{indike}^{rupture}$, the total barrels of fluid in the on-site surface soil, $Bbl_{ss-onsite}^{rupture}$, the total barrels of fluid in the off-site surface soil, $Bbl_{ss-offsite}^{rupture}$, and the total barrels of fluid that reach water, Bbl_{water}^{leak}, using Equation (5.26) through Equation (5.29), respectively.

i) STEP 12.9 – Calculate the financial environmental cost for a storage tank bottom rupture, $FC_{environ}^{rupture}$, using Equation (5.30) where $Bbl_{indike}^{rupture}$, $Bbl_{ss-onsite}^{rupture}$, $Bbl_{ss-offsite}^{rupture}$, and Bbl_{water}^{leak} are from STEP 12.8.

j) STEP 12.10 – Calculate the total financial environmental cost from a leak and a rupture, $FC_{environ}$, using Equation (5.51) where $FC_{environ}^{leak}$ is from STEP 12.6 and $FC_{environ}^{rupture}$ is from STEP 12.9.

k) STEP 12.11 – Calculate the component damage cost, FC_{cmd}, using Equation (5.53) with the release hole size damage costs from Part 3, Table 4.15 and generic failure frequencies for the release hole sizes from STEP 2.3. The material cost factor, $matcost$, is obtained from Part 3, Table 4.16.

$$FC_{cmd} = \left(\frac{\sum_{n=1}^{3} gff_n \cdot holecost_n + gff_4 \cdot holecost_4 \cdot \left(\frac{D_{tank}}{C_{36}} \right)^2}{gff_{total}} \right) \cdot matcost \tag{5.53}$$

The parameter, $\left(\frac{D_{tank}}{C_{36}} \right)^2$, is a cost adjustment factor for a storage tank bottom replacement. The cost factor included in Part 3, Table 4.15 is normalized for a storage tank with a diameter of 30.5 m (100 ft), and this factor corrects the cost for other storage tank diameters.

l) STEP 12.12 – Calculate the total financial COF, FC_{total}, using Equation (5.54).

$$FC_{total} = FC_{environ} + FC_{cmd} + FC_{prod} \tag{5.54}$$

4.26 Nomenclature

The following lists the nomenclature used in Section 2.0. The coefficients C_1 through C_{36} which provide the metric and U.S conversion factors for the equations are provided in Part 3, Annex 3.B.

age is the in-service time that the damage is applied, years

age_{rc} is the remaining life of the internal liner associated with the date of the starting thickness, years

$age_{f,rc}$ is the final remaining life of the internal liner after adjusting for liner age factors, years

age_{tk} is the component in-service time since the last inspection thickness measurement or service start date, years

A_n is the hole area associated with the n^{th} release hole size, inch^2 (mm^2)

A_{rt} is the component wall loss fraction since last inspection thickness measurement or service start date

Bbl_{total} is the product volume in the storage tank, barrels

$Bbl_{avail,n}$ is the available product volume for the n^{th} release hole size due to a leak, barrels

$Bbl_{groundwater,n}^{leak}$ is the product volume for the n^{th} release hole size due to a leak in the groundwater, barrels

$Bbl_{subsoil,n}^{leak}$ is the product volume for the n^{th} release hole size due to a leak in the subsoil, barrels

Bbl_{n}^{leak} is the product volume for the n^{th} release hole size due to a leak, barrels

$Bbl_{groundwater}^{leak}$ is the total product volume in the groundwater due to a leak, barrels

Bbl_{indike}^{leak} is the total product volume in the dike due to a leak, barrels

$Bbl_{release}^{leak}$ is the total product volume released due to a leak, barrels

$Bbl_{ssoffsite}^{leak}$ is the total product volume released on the surface located on-site due to a leak, barrels

$Bbl_{ssonsite}^{leak}$ is the total product volume released on the surface located off-site due to a leak, barrels

$Bbl_{subsoil}^{leak}$ is the total product volume in the subsoil due to a leak, barrels

Bbl_{water}^{leak} is the total product volume in the water due to a leak, barrels

$Bbl_{n}^{rupture}$ is the product volume for the n^{th} release hole size due to a rupture, barrels

$Bbl_{indike}^{rupture}$ is the product volume in the dike due to a rupture, barrels

$Bbl_{release}^{rupture}$ is the product volume in released due to a rupture, barrels

$Bbl_{ssonsite}^{rupture}$ is the product volume on the surface located on-site due to a rupture, barrels

$Bbl_{ssoffsite}^{rupture}$ is the product volume on the surface located off-site due to a rupture, barrels

$Bbl_{water}^{rupture}$ is the total product volume in the water due to a rupture, barrels

CHT is the course height of the storage tank, m (ft)

C_d is the discharge coefficient

C_{indike} is the environmental cost for product in the dike area, \$/bbl

$C_{ss-onsite}$ is the environmental cost for product on the surface located on-site, \$/bbl

$C_{ss-offsite}$ is the environmental cost for product on the surface located off-site, \$/bbl

C_{water} is the environmental cost for product in water, \$/bbl

$C_{subsoil}$ is the environmental cost for product in the subsoil, \$/bbl

$C_{groundwater}$ is the environmental cost for product in the groundwater, \$/bbl

C_{qo}	is the adjustment factor for degree of contact with soil
CA	is the corrosion allowance, in (mm)
$C_{r,bm}$	is the corrosion rate for the base material, inch/year (mm/y)
$D_f^{Tank,Thin}$	is the DF for thinning
d_n	is the diameter of the n^{th} release hole, in (mm)
D_{tank}	is the storage tank diameter, ft (m)
D_{fB}^{Thin}	is the base value of the DF for thinning
E	is the weld joint efficiency or quality code from the original construction code
F_{AM}	is the DF adjustment for AST maintenance per API 653
F_{LC}	is the DF adjustment for lining condition
F_{OM}	is the DF adjustment for online monitoring
F_{SM}	is the DF adjustment for settlement
F_{WD}	is the DF adjustment for welded construction
$FC_{environ}$	is the financial consequence of environmental clean-up, $
FC_{cmd}	is the financial consequence of component damage, $
FC_{prod}	is the financial consequence of lost production on the unit, $
FC_{total}	is the total financial consequence, $
$FC_{environ}^{leak}$	is the financial consequence of environmental cleanup for leakage, $
$FC_{environ}^{rupture}$	is the financial consequence of environmental cleanup for leakage, $
g	is the acceleration due to gravity on earth at sea level = 32.2 ft/s^2 (9.81 m/s^2)
gff_n	are the generic failure frequencies for each of the n release hole sizes selected for the type of equipment being evaluated
gff_{total}	is the sum of the individual release hole size generic frequencies
h_{liq}	is the maximum fill height in the storage tank, ft (m)
k_h	is the soil hydraulic conductivity, ft/day (m/day)
$k_{h,prod}$	is the soil hydraulic conductivity based on the storage tank product, ft/day (m/day)
$k_{h,water}$	is the soil hydraulic conductivity based on water, ft/day (m/day)
$k_{h,water-lb}$	is the lower bound soil hydraulic conductivity based on water, in/s (cm/s)
$k_{h,water-ub}$	is the upper bound soil hydraulic conductivity based on water, in/s (cm/s)
ld_n	is the actual leak duration of the release based on the available mass and the calculated release rate, associated with the n^{th} release hole size, day
$Lvol_{above,n}$	is the total liquid volume for the n^{th} release hole size, ft^3 (m^3)
$Lvol_{avail,n}$	is the available liquid volume for the n^{th} release hole size, ft^3 (m^3)
$Lvol_{above,i}$	is the total liquid volume above the i^{th} storage tank course, v
$Lvol_{total}$	is the total liquid volume in the storage tank, ft^3 (m^3)
$LHT_{above,i}$	is the liquid height above the i^{th} storage tank course, ft (m)

$matcost$	is the material cost factor
$mass_{total}$	is the available mass for release. barrels
N_c	is the total number of storage tank courses
n^{th}	is the representative holes sizes
$n_{rh,n}$	is the number of release holes for each release hole size as a function of the storage tank diameter
N_A^{Thin}	is the number of A level inspections
N_B^{Thin}	is the number of B level inspections
N_C^{Thin}	is the number of C level inspections
N_D^{Thin}	is the number of D level inspections
μ_l	is the dynamic viscosity, $(lb_f\text{-}s)/ft^2$ $((N\text{-}s)/m^2)$
μ_w	is the dynamic viscosity of water at storage or normal operating, $(lb_f\text{-}s)/ft^2$ $((N\text{-}s)/m^2)$
$Outage_{affa}$	is the numbers of days of downtime required to repair damage to the surrounding equipment, days
$Outage_n$	is the number of downtime days to repair damage associated with the n^{th} release hole size, days
p_s	is the soil porosity
P_{lvdike}	is the percentage of fluid leaving the dike
P_{onsite}	is the percentage of fluid that leaves the dike area but remains on-site
$P_{offsite}$	is the percentage of fluid that leaves the dike area, remains off-site and remains out of nearby water
ρ_l	is the liquid density at storage or normal operating conditions, lb/ft^3 (kg/m^3)
ρ_w	is the density of water at storage or normal operating conditions, lb/ft^3 (kg/m^3)
$rate_n$	is the adjusted or mitigated discharge rate used in the consequence calculation associated with the n^{th} release hole size, bbl/day
S	is the allowable stress, psi (MPa)
s_{gw}	is the distance to the groundwater underneath the storage tank, ft (m)
t	is the furnished thickness of the component calculated as the sum of the base material and cladding/weld overlay thickness, as applicable, in (mm)
t_c	is the minimum structural thickness of the component base material, in (mm)
t_{gl}	is the time required for the product to reach the groundwater through a leak in the storage tank bottom, days
t_{ld}	is the leak detection time, days
t_{min}	is the minimum required thickness based on the applicable construction code, in (mm)
t_{rdi}	the furnished thickness, t, or measured thickness reading from previous inspection, only if there is a high level of confidence in its accuracy, with respect to wall loss associated with internal corrosion, in (mm)
$vel_{s,prod}$	is the seepage velocity, ft/day (m/day)
W_n	is the discharge rate of the storage tank product through a hole in the course, bbl/day

4.27 Tables

Table 4.1 – Internal Liner Types

Internal Liner	Lining Resistance	Expected Age
Alloy Strip Liner	Subject to failure at seams	5-15 years
Organic Coating - Low Quality Immersion Grade Coating (Spray Applied, to 40 mils)	Limited life	1-3 years
Organic Coating - Medium Quality Immersion Grade Coating (Filled, Trowel Applied, to 80 mils)	Limited life	3-5 years
Organic Coating - High Quality Immersion Grade Coating (Reinforced, Trowel Applied, ≥ 80 mils)	Limited life	5-10 years
Thermal Resistance Service: Castable Refractory Plastic Refractory Refractory Brick Ceramic Fiber Refractory Refractory/Alloy Combination	Subject to occasional spalling or collapse	1-5 years
Thermal Resistance Service: Castable Refractory Ceramic Tile	Limited life in highly abrasive service	1-5 years
Glass Liners	Complete protection, subject to failure due to thermal or mechanical shock	5-10 years
Acid Brick	Partial protection. The brick provides thermal protection, but is not intended to keep the fluid away from the base material	10-20 years

Table 4.2 – Lining Condition Adjustment

Qualitative Condition	Description	Adjustment Multiplier – F_{LC}
Poor	The lining has either had previous failures or exhibits conditions that may lead to failure in the near future. Repairs to previous failures are not successful or are of poor quality.	3
Average	The lining is not showing signs of excessive attack by any damage mechanisms. Local repairs may have been performed, but they are of good quality and have successfully corrected the lining condition.	2
Good	The lining is in “like new” condition with no signs of attack by any damage mechanisms. There has been no need for any repairs to the lining.	1

Table 4.3 – Thinning Damage Factors for Storage Tank Bottom Components

A_{rt}	Inspection Effectiveness				
	E	1 Inspection			
		D	C	B	A
0.00	0.1	0.1	0.1	0.1	0.1
0.05	4	1	0.5	0.4	0.3
0.10	14	3	1	0.7	0.5
0.15	32	8	2	1	0.7
0.20	56	18	6	2	1
0.25	87	32	11	4	3
0.30	125	53	21	9	6
0.35	170	80	36	16	12
0.40	222	115	57	29	21
0.45	281	158	86	47	36
0.50	347	211	124	73	58
0.55	420	273	173	109	89
0.60	500	346	234	158	133
0.65	587	430	309	222	192
0.70	681	527	401	305	270
0.75	782	635	510	409	370
0.80	890	757	638	538	498
0.85	1,005	893	789	696	658
0.90	1,126	1,044	963	888	856
0.95	1,255	1,209	1,163	1,118	1,098
1.00	1,390	1,390	1,390	1,390	1,390

Table 4.4 – Release Hole Sizes and Areas – Storage Tank Courses

Release Hole Number	Release Hole Size	Range of Hole Diameters (inch)	Release Hole Diameter (inch)
1	Small	0 – 1/8	$d_1 = 0.125$
2	Medium	> 1/8 – ¼	$d_2 = 0.25$
3	Large	> ¼ – 2	$d_3 = 2$
4	Rupture	> 2	$d_4 = 12\left(\frac{D_{tank}}{4}\right)$

Table 4.4M – Release Hole Sizes and Areas – Storage Tank Courses

Release Hole Number	Release Hole Size	Range of Hole Diameters (mm)	Release Hole Diameter (mm)
1	Small	0 – 3.175	$d_1 = 3.175$
2	Medium	> 3.175 – 6.35	$d_2 = 6.35$
3	Large	> 6.35 – 50.8	$d_3 = 50.8$
4	Rupture	> 50.8	$d_4 = 1000\left(\frac{D_{tank}}{4}\right)$

Table 4.5 – Fluids and Fluid Properties for Storage Tank Consequence Analysis

Fluid	Level 1 Consequence Analysis Representative Fluid	Molecular Weight	Liquid Density (lb/ft^3)	Liquid Dynamic Viscosity ($lb_f\text{-}s/ft^2$)
Gasoline	C_6-C_8	100	42.702	8.383E-05
Light Diesel Oil	C_9-C_{12}	149	45.823	2.169E-05
Heavy Diesel Oil	C_{13}-C_{16}	205	47.728	5.129E-05
Fuel Oil	C_{17}-C_{25}	280	48.383	7.706E-04
Crude Oil	C_{17}-C_{25}	280	48.383	7.706E-04
Heavy Fuel Oil	C_{25+}	422	56.187	9.600E-04
Heavy Crude Oil	C_{25+}	422	56.187	9.600E-04

Table 4.5M – Fluids and Fluid Properties for Storage Tank Consequence Analysis

Fluid	Level 1 Consequence Analysis Representative Fluid	Molecular Weight	Liquid Density (kg/m^3)	Liquid Dynamic Viscosity ($N\text{-}s/m^2$)
Gasoline	C_6-C_8	100	684.018	4.01E-03
Light Diesel Oil	C_9-C_{12}	149	734.011	1.04E-03
Heavy Diesel Oil	C_{13}-C_{16}	205	764.527	2.46E-03
Fuel Oil	C_{17}-C_{25}	280	775.019	3.69E-02
Crude Oil	C_{17}-C_{25}	280	775.019	3.69E-02
Heavy Fuel Oil	C_{25+}	422	900.026	4.60E-02
Heavy Crude Oil	C_{25+}	422	900.026	4.60E-02

Table 4.6 – Cost Parameters Based on Environmental Sensitivity

Location (1)	Description	Environmental Sensitivity		
		Low (US$/bbl)	Medium (US$/bbl)	High (US$/bbl)
1	C_{indike} – Environmental cost for product located in the dike area	10	10	10
2	$C_{ss-onsite}$ – Environmental cost for product located in surface soil located on-site	50	50	50
3	$C_{ss-offsite}$ – Environmental cost for product located in surface soil located off-site	100	250	500
4	$C_{subsoil}$ – Environmental cost for product located in subsoil	500	1500	3000
5	$C_{groundwater}$ – Environmental cost for product located in groundwater	1,000	5,000	10,000
6	C_{water} – Environmental cost for product in surface water	500	1,500	5,000

NOTE 1:See Figure 4.1

NOTE 2: The values shown above are estimates. The end user should decide if these values are appropriate for the specific application.

Table 4.7 – Soil Types and Properties for Storage Tank Consequence Analysis

Soil Type	Hydraulic Conductivity for Water Lower Bound (in/s)	Hydraulic Conductivity for Water Upper Bound (in/s)	Soil Porosity
Gravel	3.94E-01	3.94	0.40
Coarse Sand	3.94E-03	3.94E-02	0.33
Fine Sand	3.94E-04	3.94E-03	0.33
Very Fine Sand	3.94E-06	3.94E-04	0.33
Silt	3.94E-07	3.94E-06	0.41
Sandy Clay	3.94E-08	3.94E-07	0.45
Clay	3.94E-09	3.94E-08	0.50
Concrete-Asphalt	3.94E-12	3.94E-11	0.3

Table 4.7M – Soil Types and Properties for Storage Tank Consequence Analysis

Soil Type	Hydraulic Conductivity for Water Lower Bound (cm/s)	Hydraulic Conductivity for Water Upper Bound (cm/s)	Soil Porosity
Gravel	1.00E00	1.00E01	0.40
Coarse Sand	1.00E-02	1.00E-01	0.33
Fine Sand	1.00E-03	1.00E-02	0.33
Very Fine Sand	1.00E-05	1.00E-03	0.33
Silt	1.00E-06	1.00E-05	0.41
Sandy Clay	1.00E-07	1.00E-06	0.45
Clay	1.00E-08	1.00E-07	0.50
Concrete-Asphalt	1.00E-11	1.00E-10	0.3

Table 4.8 – Release Hole Sizes and Areas – Storage Tank Bottoms

Release Hole Number	Release Hole Size	Release Prevention Barrier?	Range of Hole Diameters (inch)	Release Hole Diameter (inch)
1	Small	Yes	0 – 1/8	$d_1 = 0.125$
		No	0 – ½	$d_1 = 0.50$
2	Medium	NA	0	$d_2 = 0$
		NA	0	
3	Large	NA	0	$d_3 = 0$
		NA	0	
4	Rupture	Yes	> 1/8	$d_4 = 12\left(\frac{D_{tank}}{4}\right)$
		No	> 1/2	

Table 4.8M – Release Hole Sizes and Areas – Storage Tank Bottoms

Release Hole Number	Release Hole Size	Release Prevention Barrier?	Range of Hole Diameters (mm)	Release Hole Diameter (mm)
1	Small	Yes	0 – 3.175	$d_1 = 3.175$
		No	0 – 12.7	$d_1 = 12.7$
2	Medium	NA	0	$d_2 = 0$
		NA	0	
3	Large	NA	0	$d_3 = 0$
		NA	0	
4	Rupture	Yes	> 3.175	$d_4 = 1000\left(\frac{D_{tank}}{4}\right)$
		No	> 12.7	

Table 4.9 – Number of Release Holes as a Function of Storage Tank Diameter

Storage Tank Diameter (m (ft))	Number of Release Holes With or Without a Release Prevention Barrier		
	Small	Medium	Large
30.5 (100)	1	0	0
61.0 (200)	4	0	0
91.4 (300)	9	0	0

NOTE: For intermediate storage tank diameters, the number of small release holes may be calculated using the following equation where the function nint() is defined as the nearest integer. For example, nint(3.2)=3, nint(3.5)=4, and nint(3.7)=4.

$$n_{rh,1} = \max\left[\text{nint}\left[\left(\frac{\mathrm{D}}{\mathrm{C}_{36}}\right)^2\right],\ 1\right]$$

4.28 Figures

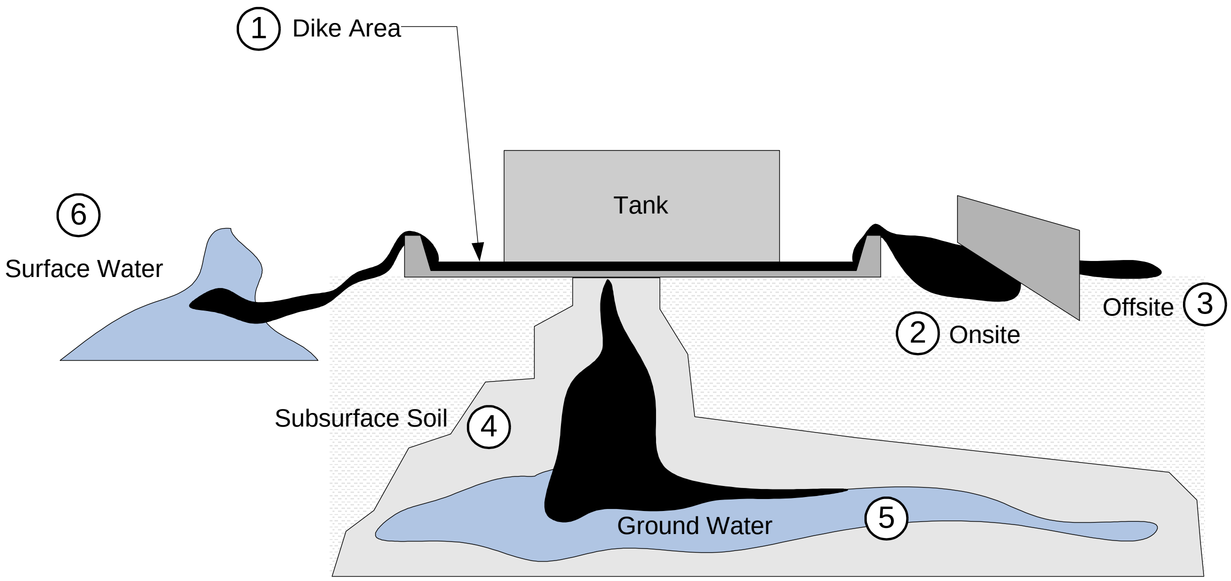


Figure 4.1 – Storage Tank Consequence

5 HEAT EXCHANGER TUBE BUNDLES

5.1 Overview

This section describes a methodology to assess the reliability and remaining life of heat exchanger bundles. It also provides a methodology for performing cost benefit analysis to assist in making risk-based inspection and replacement decisions and to determine the optimal replacement frequency of heat exchanger bundles.

The purpose of the module is to manage heat exchanger bundle inspection and replacement cycles and to reduce annual operating and maintenance costs of heat exchanger bundles. The costs considered include bundle fabrication and installation costs, environmental impact costs, and lost opportunity costs due to unit and plant unplanned shutdowns or unit rate reductions as a result of an unplanned bundle failure.

5.2 Background

Analyzing each heat exchanger bundle service history generally does not consider the financial consequences associated with a bundle failure. Many exchangers experience few or no bundle failures while some failures may not have occurred at the current operating conditions or practices. In addition, statistically significant data may not exist in order to make an accurate prediction of future performance or POF for the heat exchanger bundle.

5.3 Basis of Model

The application of risk principles to exchanger bundle inspection allows decisions to be made based on the consequences of bundle failure, including costs associated with lost production and environmental impact costs associated with leakage into utility systems and the replacement and maintenance costs associated with bundle replacement.

The combined experience of heat exchanger bundles of similar design and service is combined and statistically analyzed to provide a prediction of future performance. The exchanger bundle under evaluation is matched to similar bundles and statistically analyzed using a Weibayes or similar analysis to estimate the POF of the bundle. The results from the analysis are used to determine if the exchanger bundle will operate safely and reliably until the next scheduled maintenance opportunity. See Annex 5.A for an explanation on how to determine Weibull parameters.

5.4 Required Data

The data listed in Table 5.1 shows the minimum data requirements for each heat exchanger bundle.

5.4.1 Methodology Overview

Calculations for the risk and inspection for heat exchanger bundles are performed following the flow chart shown in Figure 5.1.

An overview of the steps for calculating Risk are shown in the following steps:

a) STEP 1.1 – Gather and input the bundle basic input data required as defined in Table 5.1.
b) STEP 1.2 – Gather and input inspection historical data, if available.
c) STEP 1.3 – Determine the maximum acceptable POF, P_{tgt}^{tube} based on the calculated C_f^{tube} and the risk target, $Risk_{tgt}$.
d) STEP 1.4 – Provide $MTTF$ or Weibull parameters for the bundle failure rate curve based on historical bundle failures in the same or similar service using Weibull analysis or some other statistical approach.
e) STEP 1.5 – Calculate the POF at the current date, RBI Date, Plan Date, Turnaround Date 1 and Turnaround Date 2 using the Weibull data.
f) STEP 1.6 – Determine consequences of bundle failure, C_f^{tube}.
g) STEP 1.7 – Calculate the risk at the plan date with and without inspection.
h) STEP 1.8 – Define a recommended inspection plan.
i) STEP 1.9 – Calculate a bundle replacement frequency.

5.5 Probability of Failure

5.5.1 Definition of Bundle Failure

A definition of bundle failure was established to determine the bundle life (failure) to predict the point in time at which an existing bundle will reach its end of life. A failure is defined as a tube leak for the purposes of RBI.

The current condition or remaining life of a bundle is quantified and expressed as a percent of the original wall thickness when the controlling damage mechanism for the bundle is general corrosion. Inspection data may be used to determine when failure occurred or to predict when a failure is likely to occur if inspection records documenting average remaining wall thickness are available. Other damage mechanisms, such as local corrosion, erosion, or vibration damage are not easily predicted based on inspection measurements. In these cases, a predicted bundle life is based on a remaining life estimate.

Bundles are often replaced or repaired prior to failure due to a deteriorated condition. An assumed remaining bundle life (25% remaining life is recommended) for the degraded condition provides an adjustment for a bundle replaced prior to a failure. A degraded condition is a bundle that would not be expected to make another operating cycle without expected tube failures.

5.5.2 Probability of Failure Using Weibull Distribution

a) The POF for a heat exchanger bundle is expressed using a two parameter Weibull distribution is Equation (5.55) [4].

$$P_f^{tube} = 1 - R(t) = 1 - \exp\left[-\left(\frac{t}{\eta}\right)^{\beta}\right] \tag{5.55}$$

Where P_f^{tube} is the POF as a function of time or the fraction of bundles that have failed at time t, β is the Weibull shape factor that is unitless, η is the Weibull characteristic life in years, and t is the independent variable time in years.

The time to reach a specified POF is calculated by using Equation (5.55) and solving for t, as shown in Equation (5.56).

$$t = \eta \cdot \left(-\ln\left[1 - P_f^{tube}\right]\right)^{\frac{1}{\beta}} \tag{5.56}$$

b) POF is calculated as a function of in-service duration using one of the methods below:

   a) Method 1, Specified Weibull Parameters (see Part 2, Section 5.5.3.1) – The Weibull β and η parameters for the exchanger bundle are provided and used for the POF calculation. A statistical analysis such as Weibayes or other statistical analysis is used to establish the Weibull β and η parameters from an exchanger bundle reliability library or available bundle failure data. Annex 5.A shows an example of calculating Weibull parameters from an exchanger bundle reliability library.

   b) Method 2, Specified $MTTF$ (see Part 2, Section 5.5.3.2) – An $MTTF$ for the bundle is provided for the POF calculation. This approach uses the $MTTF$ to calculate a Weibull η parameter using a β value of 3.0. As an option, the Weibull β parameter in addition to the $MTTF$ is specified.

   c) Method 3, Specific Bundle Inspection History (see Part 2, Section 5.5.3.3) – Statistical approaches are outlined to calculate the η parameter for the bundle, if sufficient inspection history is available.

5.5.3 POF Calculation

5.5.3.1 POF Using the Supplied Weibull Parameters

The β and η parameters for the exchanger bundle are provided from a statistical analysis and used in Equation (5.55) to determine the POF for the bundle as a function of time.

5.5.3.2 POF using the Supplied MTTF

An MTTF is calculated if sufficient inspection information exists for a bundle using a Weibull distribution with a known β parameter (default to 3.0 if unknown), and η parameter using the gamma function in Equation (5.57).

$$MTTF = \eta \cdot \Gamma\left[1+\frac{1}{\beta}\right] \tag{5.57}$$

POF is calculated using Equation (5.55) for β and η.

5.5.3.3 POF Calculated using Specific Bundle History

Information gained from inspection of the tube bundle is used to assess the actual condition of the bundle and adjust the POF rate. Inspection provides two benefits:

A reduction in uncertainty due to the effectiveness of the inspection providing a more accurate assessment of the bundle condition and failure rate.

d) Improved knowledge of the true condition of the bundle by using measured tube wall thicknesses to make an estimate of the remaining life.

Inaccuracies and biases are addressed with uncertainty, as shown in API RP 581, Part 2, Annex C, Table 2.C.4.1. Uncertainty is reduced and the POF decreases through bundle inspection. The level of uncertainty decreases as more effective inspection techniques are used and risk reduction through inspection results in more rigorous inspection techniques as the bundle reaches end of life. Inspection effectiveness is discussed in more details in API RP 581, Part 2, Annex 2.C. The bundle may reach a time in life when inspection (more data) does little or nothing to lower the risk and repair, replace, coat, or other recommendations are more appropriate. This is typically because it is actually at or near end of life.

5.5.3.3.1 Specific Bundle History

Inspection provides knowledge of the current condition of the bundle. Inspection determines if the bundle is in better or worse condition than predicted by using data from similar service bundles.

If general corrosion is the primary damage mechanism, average measured tube thickness data is used to predict the bundle failure date. When other damage mechanisms (such as vibration or tube end thinning) or when measured thickness data does not exist, a qualitative estimate of the remaining life is used to predict the bundle failure date. Two methods are provided for inspection data use in adjusting the POF calculation.

Calculated Failure Data Based on Measured Thickness Data

The thinning rate of the tube bundle, t_{rate}, is calculated using the average furnished wall thickness, $\overline{t}_{orig}$, and average measured wall thickness, $\overline{t}_{insp}$, from inspection, and the time in service, t_{dur}, using Equation (5.58):

$$t_{rate} = \frac{\overline{t}_{orig} - \overline{t}_{insp}}{t_{dur}} \tag{5.58}$$

The calculated rate is adjusted, $t_{rate,adj}$, in Equation (5.59) uses the probabilities and damage state factors used in the thinning damage factor calculation in API RP 581, Part 2, Section 4.5.7.

$$t_{rate,adj} = \left(t_{rate1} \cdot D_1^{Bundle}\right) + \left(t_{rate2} \cdot D_2^{Bundle}\right) + \left(t_{rate3} \cdot D_3^{Bundle}\right) \quad (5.59)$$

Where t_{rate1}, t_{rate2} *and* t_{rate3} are the thinning states based on the measured corrosion rate from inspection and D_1^{Bundle}, D_2^{Bundle} *and* D_3^{Bundle} are the probabilities Part 2, Table 4.5.

The calculated bundle life, PBL_{adj}, is adjusted for inspection using Equation (5.60).

$$PBL_{adj} = \frac{RWT_f \cdot \bar{t}_{orig}}{t_{rate,adj}} \quad (5.60)$$

Where the failure point is defined as a fraction of remaining wall thickness, RWT_f.

e) Calculated Failure Data Based on Estimated Remaining Life

The estimated remaining life, ERL, of the bundle is used to calculate bundle life if tube wall thickness data is not available for calculation of a bundle tube thinning rate, or when the damage mechanism is not general corrosion. The ERL is calculated using inspection data combined with accepted Fitness-For-Service calculations based on the damage mechanism known or anticipated and the time in service, t_{dur}.

$$PBL_{adj} = t_{dur} + ERL \quad (5.61)$$

5.5.3.3.2 Adjustment to Failure Rate Based on Condition of Bundle

A bundle with a recommended two or more life cycles with inspection data is used to calculate a β parameter for the matching bundle criteria (default to 3.0 if unknown) with a Weibayes analysis. The η parameter is calculated using Equation (5.62).

$$\eta = \left(\sum_{i=1}^{N} \frac{t_{dur,i}^{\beta}}{r}\right)^{\frac{1}{\beta}} \quad (5.62)$$

Where N is the number of past bundles, $t_{dur,i}$ is the time in service for each bundle in years, r is the number of failed bundles, and β is the Weibull slope parameter. This method assumes that the current operating conditions for the bundle have not changed including changes in metallurgy, process conditions, or bundle design. POF is calculated using Equation (5.55) for β and η.

A modified characteristic life, η_{mod}, for the bundle is calculated using Equation (5.63) if the bundle life is calculated based on the last inspection using Equation (5.60) or Equation (5.63).

$$\eta_{mod} = \left(\frac{1}{r}\sum_{i=1}^{N} t_{i,dur}^{\beta}\right)^{\frac{1}{\beta}} \quad (5.63)$$

Where N is the number of past bundles, $t_{dur,i}$ is the time in service for each bundle in years, r is the number of failed bundles, and β is the Weibull slope parameter.

NOTE: If the bundle was replaced before failure, a factor of 1.25 is applied to the service duration. This method assumes that the current operating conditions for the bundle have not changed including changes in metallurgy, process conditions, or bundle design. POF is calculated using Equation (5.55) for β and η_{mod}.

The recommended inspection interval at the target POF for the bundle is calculated using Equation (5.64):

$$t_{insp} = \eta_{mod} \cdot \left(-\ln\left[1 - P_{f,tgt}^{tube}\right]\right)^{\frac{1}{\beta}} \tag{5.64}$$

The adjusted characteristic life, and adjusted POF, $P_{f,adj}^{tube}$,of the bundle is calculated using η_{mod} from Equation (5.63) using Equation (5.65).

$$P_{f,adj}^{tube} = 1 - \exp\left[-\left(\frac{t}{\eta_{mod}}\right)^{\beta}\right] \tag{5.65}$$

5.5.3.3.3 Effects of Bundle Life Extension Efforts

Minor repairs and cleaning operations performed on bundles prior to inspection do not affect the life of the bundle. However, there are life extension methods that are often implemented during shutdowns that return the bundle to service in an improved condition. An adjustment is made to the inspection interval based on Table 4.2 for life extension methods and by determining a life extension factor, LEF . The adjusted service duration, $tadj_{dur}$, is calculated with the LEF using Equation (5.66).

$$tadj_{dur} = \left(1 + LEF\right) \cdot t_{dur} \tag{5.66}$$

NOTE: The actual service duration, t_{dur} , is the time period in years between the bundle installation date and the inspection date that the life extension method was performed, as shown in Equation (5.67).

$$t_{dur} = Inspect\ Date - Install\ Date \tag{5.67}$$

The Effective Installation Date, $Bundle\ Installation\ Date_{adj}$, is calculated using $tadj_{dur}$, as shown in Equation (5.68).

$$Bundle\ Installation\ Date_{adj} = Inspect\ Date - tadj_{dur} \tag{5.68}$$

5.6 Consequence of Failure

Bundle failure is defined as a tube leak. Financial consequence of failure (COF) is determined based on the bundle criticality which includes costs associated with lost opportunity due to production downtime, environmental impact costs, and costs associated with maintenance and replacement of the bundle. The consequence of an unplanned shutdown due to a bundle tube leak is determined using Equation (5.69).

$$C_f^{tube} = \left(Unit_{prod} \cdot \frac{Rate_{red}}{100} \cdot D_{sd}\right) \cdot Outage_{mult} + Cost_{env} + \left(Cost_{bundle} \cdot matcost\right) + Cost_{maint} \tag{5.69}$$

Where D_{sd} is the time in days for a planned or unplanned shutdown and $matcost$ factor is from Table 4.3.

5.7 Risk Analysis

Risk over time is calculated to determine what inspection is required to manage risk. Uncertainty exists when relevant, credible data is lacking. More relevant data reduces the amount of uncertainty in the risk calculation. Information from inspection is often needed to improve confidence in the damage states and damage rates associated with bundles. Risk for bundles is a function of time is the product of the POF and the COF in financial terms, as shown in Equation (5.70).

$$Risk_f^{tube} = P_f^{tube} \cdot C_f^{tube} \tag{5.70}$$

5.7.1 Risk Matrix

A risk matrix is a valuable visual tool for identifying high risk bundles. The risk of each bundle is characterized by the POF and COF categories, shown in Part 1, Section 4.3.2.2 and enables each bundle to be plotted on the risk matrix as shown in Part 1, Figure 4.2 and Figure 4.3.

The risk matrix is grouped into four areas: high risk, medium high risk, medium risk, and low risk. If an exchanger has been identified as high risk prior to the turnaround, it would require a more rigorous inspection than has been used on that bundle in the past. For example, if the bundle were determined to be a HIGH risk on the risk matrix and past inspections for that bundle were *usually effective*, it is very likely that a *highly effective inspection* would be required at the upcoming shutdown. The benefits of the different levels of inspection are discussed in Section 5.8.

5.8 Inspection Planning Based on Risk Analysis

The inspection target date is the date at which the calculated risk using Equation (5.55) exceeds the risk target, $Risk_{tgt}$. An inspection is required prior to the target date to maintain a risk level below the risk target. The target date for the next inspection is calculated using the inspection adjusted Weibull parameters.

5.8.1 Use of Risk Target in Inspection Planning

The risk target is a function of the owner-~~user~~operator's corporate philosophy for making risk decisions. Some companies are more risk adverse than others, and this will have a direct impact on the inspection planning results.

Equation (5.71) is used to calculate the target POF for a bundle as a function of the COF and using the target risk:

$$P_{f,tgt}^{tube} = \frac{Risk_{tgt}}{C_f^{tube}} \tag{5.71}$$

A target inspection date is calculated using Equation (5.56). The target date is the date when the bundle risk reaches the target risk.

A user defined $P_{f,tgt}^{tube}$ is used in place of the calculated $P_{f,tgt}^{tube}$ if a lower risk or probability of bundle failure is required for inspection planning.

The target inspection time is calculated using Equation (5.72). The target time is the number of years from the installation date when the bundle risk reaches the target risk.

$$t_{insp} = \eta_{tgt} \cdot \left(-\ln\left[1 - P_{f,tgt}^{tube}\right]\right)^{\frac{1}{\beta}} \tag{5.72}$$

The target inspection date is calculated using Equation (5.73) using t_{insp} and the installation date. The target date is the date when the bundle risk reaches the target risk.

$$Target\ Inspection\ Date = Bundle\ Installation\ Date + t_{insp} \tag{5.73}$$

Bundle target characteristic life, η_{tgt}, is calculated using the $P_{f,tgt}^{tube}$ and the bundle age at the plan date as shown in Equation (5.74).

$$\eta_{tgt} = \frac{t_{plan}}{-\ln\left[1 - P_{f,tgt}^{tube}\right]^{\frac{1}{\beta}}} \tag{5.74}$$

5.8.2 Determine Inspection Recommendation

Once a decision has been made to inspect per Equation (5.74), an economic decision can be made as to the appropriate level of inspection with similar techniques as described in Section 5.9.1 by comparing the cost of the various inspection techniques to the reduction in risk expected for the level of inspection.

NOTE: ~~n~~No inspection is required if $P_{f,plan}^{tube} \le P_{f,tgt}^{tube}$.

The target Uncertainty, $AU_{tgt}\%$ is the level of uncertainty associated with an inspection required to remain below the $P_{f,tgt}^{tube}$ at the Plan Date from Equation (5.75).

$$AU_{tgt}\% = \frac{\eta_{tgt}}{\eta_{mod}} \tag{5.75}$$

The $AU_{tgt}\%$ is used with Table 4.5 to determine the level of inspection required to achieve target $P_{f,tgt}^{tube}$ at the Plan Date. The inspection plan is defined by using the target inspection date from Equation (5.71) and the recommended inspection from Equation (5.75).

5.8.3 Calculate Characteristic Life at Plan Date

The recommended inspection uncertainty is used calculate the characteristic life at the plan date after inspection using Equation (5.76).

$$\eta_{insp} = \eta_{mod} \cdot \left(\frac{1 - AU_{w/insp}\%}{1 - AU_{noinsp}\%} \right) \tag{5.76}$$

Where η_{mod} is defined in Equation (5.62).

5.8.4 Calculation of Risk

The POF at the plan date, $P_{f,w/insp}^{tube}$, with inspection is calculated with Equation (5.55) using t_{plan} for time at the plan date, η_{insp} from Equation (5.76) and the original β value.

5.8.5 Calculation of Risk

The Risk at the plan date is calculated usng Equation (5.70) using $P_{f,w/insp}^{tube}$ and $C_{f,plan}^{tube}$.

5.9 Bundle Inspect/Replacement Decisions using Cost Benefit Analysis

Weibull parameters are used to predict the optimal replacement frequency for a bundle and determine whether it makes economic sense to inspect or replace a bundle at an upcoming shutdown.

5.9.1 Decision to Inspect or Replace at Upcoming Shutdown

Risk reduction cost benefit is calculated from mitigating actions including various levels of inspection or bundle replacement. The cost benefit calculation includes the cost of the mitigating action to inspect or replace). An optional rate of return, ROR (or hurdle cost), is added to the cost of a bundle replacement to prevent excessive bundle replacement for borderline risk determinations that require action. The decision to perform a risk mitigating bundle inspection or bundle replacement at an upcoming turnaround is determined by comparing the incremental risk ($) associated with deferring the inspection or replacement to the cost associated with the performing the inspection or replacement.

Expected incremental risk, EIR_{t1}^{t2}, associated with deferring the inspection or replacement of a bundle to a later date is calculated using Equation (5.77).

$$EIR_{t1}^{t2} = C_f^{tube} \cdot \left(1 - \left[\frac{1 - P_f^{tube}(t_2)}{1 - P_f^{tube}(t_1)} \right] \right) \tag{5.77}$$

Where t_1 is the service duration of the bundle at the next shutdown (Turnaround Date 1) and t_2 is the service duration of the bundle at the subsequent shutdown (Turnaround Date 2).

If the cost to inspect or replace is greater than the expected incremental risk using Equation (5.78) and Equation (5.79), that action is recommended.

$$\begin{aligned} &If\ \left(Cost_{insp}+Cost_{maint}\right)\cdot\left(1+ROR\right)<EIR_{t1}^{t2}\ \ then\ inspect \\ &If\ \left(Cost_{insp}+Cost_{maint}+Hurdle\ Cost\right)<EIR_{t1}^{t2}\ \ then\ inspect \end{aligned} \tag{5.78}$$

$$\begin{aligned} &If\ \left(Cost_{bundle}+Cost_{maint}\right)\cdot\left(1+ROR\right)<EIR_{t1}^{t2}\ \ then\ replace\ the\ bundle \\ &If\ \left(Cost_{bundle}+Cost_{maint}+Hurdle\ Cost\right)<EIR_{t1}^{t2}\ \ then\ replace\ the\ bundle \end{aligned} \tag{5.79}$$

The actual inspection costs should be used when available. Maintenance costs to pull the bundles for inspection should be included in the total inspection costs when using Equation (5.78) and Equation (5.79).

5.9.2 Optimal Bundle Replacement Frequency

Maintenance optimization helps to strike a balance between cost and reliability. The cost per day of a "run to failure" strategy shows low costs early in the life of the equipment and increasing costs as reliability decreases. By overlaying the costs of an associated preventative maintenance to address the failure mode, initial costs are high, but costs per unit time decrease as time progresses. This optimization occurs at a point where the total cost function (sum of the two cost functions) is at a minimum. The time at which the minimum occurs is the optimum time to perform maintenance [5].

The optimum replacement frequency is calculated comparing the cost associated with a bundle failure (increasing with increasing replacement frequency) to the replacement cost associated with periodic planned shutdowns to replace the bundle (decreasing with increasing replacement frequency). The point where the two costs reach a minimum value is the optimum replacement frequency.

The methodology in Shultz, 2001[6] described below is recommended to determine the optimum bundle replacement frequency.

a) Increasing Risk Cost of Unplanned Outage

A planned replacement time frequency is defined by the variable, tr_n, and the risk cost associated with an unplanned failure to replace the bundle (including business interruption and bundle replacement costs) is calculated using Equation (5.80).

$$Risk_f^{tube}\left(tr_n\right)=C_{f,unplan}^{tube}\cdot P_f^{tube}\left(tr_n\right) \tag{5.80}$$

Where $C_{f,plan}^{tube}$ is defined in Equation (5.81).

$$C_{f,unplan}^{tube}=\left(Unit_{prod}\cdot\frac{Rate_{red}}{100}\cdot D_{sd,unplan}\right)\cdot Outage_{mult}+Cost_{env}+\left(Cost_{bundle}\cdot matcost\right)+Cost_{maint} \tag{5.81}$$

NOTE: Equation (5.81) is similar to Equation (5.69) but uses the unplanned outage time, $D_{sd,unplan}$. The consequence of an unplanned frequency due to a tube bundle failure, C_f^{tube}, includes business interruption, the number of days required for bundle replacement during an unplanned outage, $D_{sd,unplan}$ and environmental impact, $Cost_{env}$. The risk cost due to bundle failure increases with time since the POF, $P_f^{tube}\left(tr_n\right)$, increases with time.

b) Decreasing Cost of Bundle Replacement

The bundle replacement costs as a function of planned replacement frequency, tr, is calculated using Equation (5.82).

$$Cost_{pbr}\left(tr_n\right)=C_{f,plan}^{tube}\cdot\left[1-P_f^{tube}\left(tr_n\right)\right] \tag{5.82}$$

Where $C_{f,plan}^{tube}$ is defined in Equation (5.83).

$$C_{f,plan}^{tube} = \left(Unit_{prod} \cdot \frac{Rate_{red}}{100} \cdot D_{sd,plan} \right) \cdot Outage_{mult} + Cost_{env} + \left(Cost_{bundle} \cdot matcost \right) + Cost_{maint}$$

$$C_{f,plan}^{tube} = Cost_{env} + \left(Cost_{bundle} \cdot matcost \right) + Cost_{maint} \tag{5.83}$$

c) Optimization of Total Cost

The total cost as a function of replacement time frequency averaged over the service bundle life is calculated using Equation (5.84).

$$Cost_{total}\left(tr_n\right) = \frac{Risk_f^{tube}\left(tr_n\right) + Cost_{pbr}\left(tr_n\right)}{365.25 \cdot ESL_n} \tag{5.84}$$

The estimated service life as a function of replacement time interval may be approximated using an integration technique using Equation (5.85).

$$ESL_n = ESL_{f,n} + ESL_{p,n} \tag{5.85}$$

Where the average life of the bundles that would have been expected to fail prior to the planned replacement time, $ESL_{f,n}$, and the average life of the bundles that would not have been expected to fail prior to the planned replacement time, $ESL_{p,n}$ are summed, ESL_n.

The average life of the bundles that would have been expected to fail prior to the planned replacement time is:

$$ESL_{f,n} = ESL_{f,n-1} + tr_n \cdot \left(P_{f,n}^{tube} - P_{f,n-1}^{tube} \right) \tag{5.86}$$

The average life of the bundles that would have been expected to <u>not</u> fail prior to the planned replacement time is:

$$ESL_{p,n} = tr_n \cdot \left(1 - P_{f,n}^{tube} \right) \tag{5.87}$$

A planned replacement frequency is selected and the costs associated with the frequency is calculated to allow optimization of the total cost. The frequency is incrementally increased and the costs are calculated for each incremental step, n ($n = n+1$). The point where the costs reach a minimum is the optimum replacement frequency:

1) STEP 2.1 – Select an appropriate time step, t_s, in days. (A value for t_s of 7 to 14 days should be sufficient) and an increment of $n = 1$. Subsequent calculations will increase the increment by 1 ($n = n+1$).
2) STEP 2.2 – Calculate the planned replacement frequency, tr_n, by multiplying the increment number, n, by the time step, t_s as follows:

$$tr_n = n \cdot \frac{t_s}{365.25} \tag{5.88}$$

3) STEP 2.3 – Calculate the POF at the planned replacement frequency at increment n, $P_{f,n}^{tube}\left(tr_n\right)$, using Equation (5.55), the updated Weibull parameters based on the latest inspection of the bundle and the time value to use in Equation (5.56) is tr_n obtained in STEP 2.2.

NOTE: the time unit is in years.

4) STEP 2.4 – Calculate the average life of the bundles that would have been expected to fail prior to the planned replacement time, $ESL_{f,n}$, using Equation (5.86).

5) STEP 2.5 – Calculate the average life of the bundles that would have not been expected to fail prior to the planned replacement time, $ESL_{p,n}$, using Equation (5.87).

6) STEP 2.6 – Calculate the estimated service life, ESL_n, using Equation (5.83).

7) STEP 2.7 – Calculate the risk cost associated with bundle failure at the replacement frequency, $Risk_f\left(tr_n\right)$, using Equation (5.80).

8) STEP 2.8 – Calculate the bundle replacement cost at the replacement frequency, $Cost_{pbr}\left(tr_n\right)$, using Equation (5.82).

9) STEP 2.9 – Calculate the total costs at the replacement frequency averaged over the expected life of the bundle, $Cost_{total}\left(tr_n\right)$, using Equation (5.84).

10) STEP 2.10 – Increase the increment number by 1 ($n = n+1$) and repeat STEPs 2.2 through 2.9 until a minimum value of $Cost_{total}\left(tr_n\right)$ in STEP 2.9 is obtained.

11) STEP 2.11 – The optimal bundle replacement frequency, t_{opt}, is where the tr_n is at the minimum $Cost_{total}\left(tr_n\right)$.

5.10 Nomenclature

β	is the Weibull shape parameter that represents the slope of the line on a POF vs. time plot
Γ	is the Gamma function
η	is the Weibull characteristic life parameter that represents the time at which 62.3% of the bundles are expected to fail, years
η_{insp}	is the Weibull characteristic life parameter at the plan date after inspection, years
η_{mod}	is the Weibull modified characteristic life parameter modified with inspection history, years
η_{tgt}	is the Weibull target characteristic life parameter based on the risk target, years
$AU\%$	is the percent additional uncertainty, %
$AU_{w/insp}\%$	is the additional inspection uncertainty at the plan date after inspection, %
$AU_{w/outinsp}\%$	is the additional inspection uncertainty at the plan date before inspection, %
$AU_{tgt}\%$	is the additional inspection uncertainty required to remain below the $P_{f,tgt}^{tube}$ at the plan date, %
C_f^{tube}	is the consequence of bundle failure, $
$C_{f,plan}^{tube}$	is the consequence of bundle failure based on a planned bundle replacement, $
$C_{f,unplan}^{tube}$	is the consequence of bundle failure during an unplanned bundle replacement, $
$Cost_{bundle}$	is the replacement cost of the tube bundle, $
$Cost_{env}$	is the environmental costs due to a bundle leak, $

$Cost_{insp}$ is the cost to perform the inspection, $

$Cost_{maint}$ is the cost of maintenance for bundle inspection or replacement, $

$Cost_{pbr}(tr_n)$ is the cost per year of bundle replacement at a planned frequency, tr_n, $/year

$Cost_{prod}$ is the production losses as a result of shutting down to repair or replace a tube bundle, $

$Cost_{total}(tr_n)$ is the total cost of a bundle replacement program at a planned frequency, tr_n, $/year

D_{sd} is the number of days required to shut a unit down to repair a bundle during an unplanned shutdown, days

$D_{sd,plan}$ is the number of days required to shut a unit down to repair a bundle during a planned shutdown, days

$D_{sd,unplan}$ is the number of days required to shut a unit down to repair a bundle during an unplanned shutdown, days

D_1^{Bundle} is the probability adjustment for t_{rate1}

D_2^{Bundle} is the probability adjustment for t_{rate2}

D_3^{Bundle} is the probability adjustment for t_{rate3}

EIR_{t1}^{t2} is the expected incremental risk between turnaround dates T1 and T2, $/year

ERL is the estimated remaining life of the bundle, years

ESL_n is the estimated service life of a bundle as a function of replacement time interval, years

$ESL_{f,n}$ is the average life of bundles that would have failed at the replacement time interval, years

$ESL_{f,n-1}$ is the average life of bundles that would have failed at the previous replacement time interval (n-1), years

$ESL_{p,n}$ is the average life of bundles that would not have failed at the replacement time interval, years

LEF is the bundle life extension factor

$matcost$ is the material cost factor for the tube bundle material of construction

$MTTF$ is the mean time to failure, years

N is the number of bundles in a heat exchangers past history

$Outage_{mult}$ is the outage multiplier factor of the unit

P_f^{tube} is the probability of the bundle failure, failures/year

$P_{f,n}^{tube}$ is the probability of bundle failure calculated for the current (n) increment of the optimization procedure, failures/year

$P_{f,n-1}^{tube}$ is the probability of bundle failure calculated for the previous (n-1) increment of the optimization procedure, failures/year

$P_{f,w/insp}^{tube}$ is the probability of bundle failure at the plan date with inspection, failures/year

$P_{f,tgt}^{tube}$	is the maximum acceptable probability of bundle failure based on the owner-~~user~~operator's risk target, failures/year
PBL_{adj}	is the predicted bundle life adjusted based on inspection, years
r	is the number of failed bundles in a heat exchangers past history
$R(t)$	is the risk as a function of time, m^2/year (ft^2/year) or \$/year
$Risk_f^{tube}$	is the risk of failure of the tube bundle, \$/year
$Risk_f^{tube}(tr_n)$	is the risk of failure of the tube bundle at a planned bundle replacement frequency, tr_n, \$/year
$Rate_{red}$	is the production rate reduction on a unit as a result of a bundle being out of service, %
$Risk_{tgt}$	is the risk target, \$/year
ROR	is the fractional rate of return or hurdle rate
RWT_f	is the failure point defined as a fraction of remaining wall thickness
t	is time, years
t_1	is the service duration of the bundle at the upcoming turnaround (Turnaround Date1), years
t_2	is the service duration of the bundle at the subsequent turnaround (Turnaround Date2), years
t_{dur}	is the bundle duration or time in service, years
$t_{dur,i}^{\beta}$	is the time in service for the i^{th} bundle in a heat exchanger, years
t_{insp}	is the inspection interval, years
t_{plan}	is the time from the bundle installation date to the plan date, years
t_{rate}	is the thinning rate for the tube bundle, inch/year (mm/year)
t_{rate1}	is the corrosion rate for damage state 1, in/year (mm/year)
t_{rate2}	is the corrosion rate for damage state 2, in/year (mm/year)
t_{rate3}	is the corrosion rate for damage state 3, in/year (mm/year)
$t_{rate,adj}$	is the probability adjusted corrosion rate
t_s	is the time step used in the optimization routine for bundle replacement frequency, days
$tadj_{dur}$	is the bundle duration or time in service adjusted for life extension activities, years
tr_n	is the bundle planned replacement frequency, year
$\bar{t}_{insp}$	is the average measured tube wall thickness, in (mm)
$\bar{t}_{orig}$	is the average furnished tube wall thickness, in (mm)
$Unit_{prod}$	is the daily production margin on the unit, \$/day

5.11 Tables

Table 5.1 – Basic Data for Exchanger Bundle Risk Analysis

Bundle Remaining Life Methodology	
Specified MTTF	User specified MTTF for bundle, years to be used in calculation
Specified Weibull η	User specified Weibull characteristic life (years) to be used in calculations (β should also be provided)
Specified Weibull β	User specified Weibull slope parameter to be used in calculations (η should also be provided)
Bundle Life	The life of the bundle under evaluation, years (required for inactive bundles)
Consequences of Bundle Failure	
Financial Risk Target	User risk target, $/year
Tube Wall Failure Fraction	Wall thickness fraction that constitutes bundle failure (0 and 1.0)
Production Cost	Unit production costs, $/day (should be equal to the production rate, bbls/day x margin ($/bbl)
Production Impact	Production impact, e.g. none, bypass, bypass with rate reduction, shutdown
Rate Reduction	Rate Reduction, % (required if Production Impact is Bypass with Rate Reduction")
Planned Shutdown Days	Number of days required to repair or replace failed exchanger bundle when the shutdown is planned, days
Unplanned Shutdown Days	Number of days required to repair or replace failed exchanger bundle when the shutdown is unplanned, days (should be a longer duration than a planned shutdown to allow for lead time to mobilize or to purchase a replacement bundle)
Environmental Impact	Environmental costs associated with bundle failure that includes damage to cooling water system and towers
Lost Opportunity Cost	Additional cost beyond production losses or environmental costs as a result of bundle failure, $
Bundle Cost	Cost of replacement bundle, $
Bundle Installation Cost	Cost of maintenance required to remove, clean, and re-install exchanger bundle, $
Hurdle Cost	Additional cost above the economic breakeven point at which a decision to inspect or replace a bundle is made, $
Turnaround Date 1	The date for the next scheduled turnaround from the RBI date (used as Plan Date for calculating risk)
Turnaround Date 2	The date for the second scheduled turnaround from the RBI date (used in the cost benefit analysis to make inspection or replacement decisions)

Table 5.2 – Effects of Bundle Life Extension Methods

Life Extension Method	Life Extension Factor (LEF)[3]
Plug Tubes	0.10
180° Bundle Rotation	0.50
Partial Re-tube	0.50
Total Re-tube	0.90
Install Spare Bundle[2]	0.50
Install Tube Ferrules[1]	0.75

NOTE 1: This LEF is only valid when the tube ferrules are installed for protection against localized, tube-end damage due to erosion, corrosion or impingement.
NOTE 2: The spare bundle condition is known to be good through prior inspection. If the condition of the spare bundle is known to be excellent, a higher LEF can be used.
NOTE 3: LEFs provided in this table are suggestions. It is the responsibility of the owner-~~user~~operator to define life extensions for use for the bundle life extension methods.

Table 5.3 – Bundle Material Cost Factors

Bundle Generic Material	Tube Material Cost Factor, M_f
Carbon Steel	1.0
C-1/2 Mo	2.0
1-1/4 Cr	2.0
2-1/4 Cr	2.8
5 Cr	3.2
9 Cr	3.3
12 Cr	3.4
70/30 CuNi	3.5
90/10 CuNi	3.5
Monel 400	7.0
Nickel 200	8.5
304/309/310 SS	2.6
304L/321/347 SS	2.8
316 SS	3.0
316L SS	3.0
317L SS	4.2
410/439 SS	2.8
444 SS	3.2
904L	7.0
2205 Duplex SS	3.0
2304 Duplex SS	2.8

Table 5.3 – Bundle Material Cost Factors

Bundle Generic Material	Tube Material Cost Factor, M_f
2507 Duplex SS	4.0
AL6XN/254 SMO	7.0
Seacure/E-Brite	6.0
Admiralty Brass/Aluminum Brass/Red Brass/Muntz	2.5
Aluminum Alloy	3.0
Alloy 20 Cb3	6.5
Alloy 600	9.5
Alloy 625	11.0
Alloy 800	7.0
Alloy 825	8.0
Alloy C276	11.0
Ferralium 255	7.0
Bimetallic	4.5
Ceramic	1.0
Plastic	1.0
Titanium Gr. 2	6.0
Titanium Gr. 12	10.0
Titanium Gr. 16	14.0
Zeron 100	4.0
Zirconium Alloy	15.0
NOTE: The tube material cost factors are generic data and the user is encouraged to set values based on current material cost factors.	

Table 5.4 – Numerical Values Associated with POF and Financial-Based COF Categories for Exchanger Bundles

Probability Category (1)		Consequence Category (2)	
Category	**Range**	**Category**	**Range ($)**
1	$POF \le 0.1$	A	$COF \le \$10,000$
2	$0.1 < POF \le 0.2$	B	$\$10,000 < COF \le \$50,000$
3	$0.2 < POF \le 0.3$	C	$\$50,000 < COF \le \$150,000$
4	$0.3 < POF \le 0.5$	D	$\$150,000 < COF \le \$1,000,000$
5	$0.5 < POF \le 1.0$	E	$COF > \$1,000,000$
NOTE: In terms of the total DF, see Part 2, Section 2.3. In terms of consequence area, see Part 3, Section 4.11.4.			

Table 5.5 – Inspection Effectiveness and Uncertainty

Inspection Category	Inspection Effectiveness Category	Inspection Confidence	Inspection Uncertainty
A	Highly Effective	> 90%	< 10%
B	Usually Effective	> 70 to 90%	< 30% to 10%
C	Fairly Effective	> 50% to 70%	< 50% to 30%
D	Poorly Effective	> 40% to 50%	< 60% to 50%
E	Ineffective	≤ 40%	≥ 60%
NOTE 1: Inspection cost numbers are not provided in this table but may be used in the methodology regarding a 'repair or replace' strategy. It is the responsibility of the operator-~~user~~operator to determine the cost numbers unique to their particular operation and strategy. NOTE 2: Refer to Part 2, Annex C, Section 2.C.4 for more information. NOTE 3: The ~~operator~~owner-operator~~user~~ should consider applying confidence / uncertainty based upon the relationship between the following variables: a) Amount of the bundle inspected (percentage whole or percentage per pass) b) Examination method(s) used and degree of cleanliness c) Metallurgy of the bundle d) Damage mechanism(s) expected / found			

5.12 Figures

Figure 5.1 – Flow Chart of Bundle Calculation Approach

6 Pressure-Relief Devices (PRDs)

6.1 General

6.1.1 Overview

Pressure-Relief Devices (PRDs) are routinely inspected and tested to assure that the PRDs will relieve properly in overpressure events to prevent a loss of containment of the protected component. The impact of leaks associated with PRDs is also considered.

A risk-based approach to evaluate PRD criticality is covered in this section to set inspection and test intervals. All spring-loaded and pilot-operated Pressure-Relief Valves (PRVs) with and without rupture disks are covered. AST pressure/vacuum vents (P/Vs) and explosion hatches may be analyzed using this methodology provided Weibull parameter reliability data are available.

This methodology is not intended to be used to perform or check PRD design or capacity calculations. It is assumed that the owner–operator has completed due diligence and the PRDs have been designed in accordance with API 521 [7] and sized, selected, and installed in accordance with API 520 [8]. It is also assumed that minimum inspection practices in accordance with API 576 [9] are in place.

The methodology outlined uses a demand rate for the PRD combined with a Probability of Failure to Open On Demand (POFOD) determined from plant-specific data, if available, or using conservative default data provided. These inputs are used to generate POF as a function of time with a Weibull statistical approach. The protected component COF if the PRD fails to operate on demand is based on the methodology outlined in Part 3 substituting the operating pressure with the overpressure of each demand case combined with the consequences associated with PRD leakage. The combination of COF with a time-based POF results in an increasing risk value with time between inspection and test, allowing test intervals to be set based on risk targets.

The flow chart shown in Figure 6.1 illustrates the basic methodology required for the determination of an RBI inspection and test schedule. The basic data required for the evaluation are listed in Table 6.1.

6.1.2 Failure Modes

There are several failure modes of significance when evaluating the risks associated with PRD failure. For the PRD, the failure modes are grouped into two categories.

a) Fails to open as designed (FAIL):

  1) stuck or fails to open (FTO),

  2) device partially open (DPO),

  3) opens above set pressure (OASP).

b) Leakage failure (LEAK):

  1) leakage past device (LPD),

  2) spurious or premature opening (SPO),

  3) device stuck open (DSO).

FAIL modes generally causes the potential for the protected equipment overpressure resulting in a loss of containment. Included in FAIL modes is the case of a DPO causing a less severe equipment overpressure. ~~API 581 conservatively considers data indicating a DPO as a failure to open as discussed in Section 6.1.2.~~

A PRD OASP is included in the FAIL failure mode. The POF curves are based on bench test data where a failure is defined as any test requiring a pressure greater than 1.3 times the set pressure. A value that opens above set pressure during a test but does not exceed 1.3 times the set pressure is considered a successful test and is not included in the FAIL case.

Consequences associated with the FAIL failure mode include the effects of fires and explosions on personnel and equipment and the exposure of personnel to toxic fluids as a result of loss of containment. These consequences and their effect on personnel and equipment are further described in Part 3.

A secondary concern is a failure of the PRD to contain the process while operating at normal conditions. The API 581 methodology groups the remaining three failure modes together into the LEAK category. LPD, SPO, and DSO failures will not result in overpressure or loss of containment from the protected equipment but represent potentially unacceptable leakage from the process system. The consequences of leakage through a PRD can range from a minor nuisance, causing some loss of product, to a more severe consequence resulting in a process shutdown to repair or replace the PRD. If the PRD discharges to the atmosphere, additional consequences may be environmental impact and potential for fires, explosions, and toxic exposure.

6.1.3 Use of Weibull Curves

POFOD and the Probability of Leakage (POL) is expressed as a function of time for risk-based planning of inspections and tests. Weibull functions are suitable for this task with the added advantage that they may be used to evaluate large populations of data points to seek trends. In the absence of large sets of failure data, the functions are still useful as a starting point since the parameters involved describe both the manner of failure and the time to failure.

Using a two-parameter Weibull distribution [4], The cumulative failure density function, $F(t)$, sometimes referred to as Unreliability, is using a two-parameter Weibull distribution is shown in Equation (5.90) and discussed in Section 6.1.2.

The Weibull characteristic life parameter, η, is equivalent to the MTTF when the Weibull β parameter is equal to 1.0. Adjustments to the η parameter are made to increase or decrease the POFOD and POF as a result of environmental factors, PRD types, or available inspection data the PRD. Adjustments to the η parameter may be viewed as an adjustment to the PRD MTTF.

Determination of the default Weibull parameters assumes that PRDs in similar services will have a similar POFOD, P_{fod}, and POL, P_l and industry failure rate data may be used as a basis to establish the initial or default PRD POF. The POFOD is evaluated for the process and installation conditions, such as process temperature, process corrosivity, and the tendency of the process to foul, polymerize, block the PRD inlet, or prevent the PRD from reseating during operation. Rough handling during transportation and installation, and excessive piping vibration are also associated with failures. In addition, increased demand rates and improper installations that result in chatter may also increase the POFOD and POL.

6.1.4 PRD Testing, Inspection, and Repair

Inspection, testing, reconditioning, or replacement of PRDs are recognized safe practices and serve to reduce the POFOD and leakage. This PRD methodology assumes that a bench test performed on a PRD in the as-received condition from a process unit will result in a true determination of the performance of the PRD on the unit.

An effective inspection program for PRDs will track the history of inspection and test of each PRD. The outlined PRD methodology adjusts the POF data for each PRD based on historical data and allowing for the varying degrees of inspection effectiveness. Pass/fail test data are given the highest confidence inspection effectiveness level when a shop bench pre-pop test is performed. A lower confidence inspection effectiveness level is associated with the inspection if a PRD is inspected or overhauled without a pre-test.

6.1.5 PRD Shop Inspection/Overhaul or Replacement Start Date [9]

The PRD is assumed to be returned to service in the as-new condition when the PRD is overhauled in the shop. The original install date for the PRD is retained with a last inspection date reflecting the PRD overhaul date or the installation date after overhaul. As a result, the calculated inspection interval and subsequent new due date for the PRD is based on the date the PRD was most recent overhaul or the date the PRD was installed following the most recent overhaul.

The installation date and last inspection date will be the same for a PRD that is replaced rather than overhauled and the calculated inspection interval and next test due date is based on the installation date. The replaced PRD does not necessarily need to be a brand new valve, it could be a spare or overhauled valve.

It is important to note that there may be a delay between the time of overhaul and the time the PRD is placed in service. This time delay may occur because the PRD has a spare and is currently installed or there is a delay in recommissioning. If there is a delay in between the time of the overhaul and the time the PRD returns to service, consider using the date of return to service date to calculate the inspection interval and subsequent new due date. For example, if the last inspection date with shop inspection/overhaul was in the year 2015, and the PRD was not put back into service until 2018, then consider using 2018 to calculate the next inspection due date.

The impact on the PRDs reliability of a time delay between the testing and installation dates should ~~to~~ be considered. Where the reliability may have adversely been affected, retesting prior to the installation should be considered. The owner-operator should specify the maximum delay time after which:

a) the install date is to be used rather than the last overhaul date to calculate the next inspection date, and

~~a)~~b) a retest should be considered before installation.

Often PRDs are pop-tested either in the field or in the shop without overhauling the PRD. When a PRD is pop-tested in the field or in the shop without performing an overhaul, the PRD has not been returned to service in an as-new condition. The PRD is assumed to remain in the prior test condition and the POFOD may be adjusted based on the results of the field test (i.e. credit for inspection to reduce uncertainty). In this case, ~~is~~ the last overhaul date remains unchanged and the PRD will not get the full benefit of an overhaul. In this case, the due date is determined by adding the recommended inspection interval to the last overhaul date (not the last inspection date). For example, if PRD was pop-tested and overhauled in 2005, and then pop-tested, but not overhauled in 2010, and put back into service, the next inspection date is determined by adding the recommended inspection interval (say, 7 years) to 2005, the date of the last overhaul. The next inspection due date is therefore 2012.

6.1.6 Risk Ranking of PRDs

The PRD methodology provides individual PRD risk ranking as well as risk ranking between PRDs and other fixed equipment being evaluated.

The two key drivers for effectively risk ranking between PRDs is the:

1. Specific PRD reliability for each PRD by selecting a severity of service for the PRDs, establishing a default POF, and modifying the POFOD using the inspection and test history.

2. Relative importance or criticality of each PRD by defining the relief system design basis and knowledge of the overpressure demand cases applicable for each PRD. The PRD risk rank will increase based on the criticality of and the demand placed the PRD.

6.1.7 Link to Fixed or Protected Equipment

To effectively characterize the risk associated with PRD failure, the consequence associated with the failure of a PRD to open upon demand should be tied directly to the equipment that the PRD protects. This is accomplished using direct links to the fixed equipment RBI analysis as covered in Part 2 and Part 3 of this document. The risk of loss of containment from fixed equipment increases proportionately with the amount of overpressure that occurs as a result of the PRD failing to open on demand. In addition, the calculated risk associated with damaged fixed equipment will be greater than that for undamaged equipment since the actual damage states (i.e. damage factor, D_f, see Part 2) are used in the calculations.

Although consequences associated with PRD overpressure cases are greater than those associated with the fixed equipment operating at normal pressure, this is tempered by the fact that the use of realistic PRD demand rates and accurate PRD failure rate data results in a low frequency of occurrence.

6.2 Overpressure Potential for Overpressure Demand Cases

The PRD analysis should consider the overpressure demand cases applicable for each PRD. The overpressure demand cases are the potential process upsets that the PRD is designed protect against and the criticality of the protected equipment if a failure on demand occurs. The importance of the criticality of the protected equipment in addition to the PRD failure is demonstrated by the following examples.

Example 1: A PRD that protects equipment and piping for the blocked discharge demand case downstream of a pump is less critical than a PRD that is protecting a reactor from a runaway chemical reaction. In the former case, a lower overpressure as a result of a PRDs failure to open upon demand would be expected.

Example 2: A PRD protecting piping against thermal relief is less critical than a PRD protecting low-pressure equipment from a high-pressure gas breakthrough due to a control valve failure.

The potential overpressure resulting from a PRDs failure to open upon demand may be calculated for most of the overpressure demand cases. The logic for determining the potential overpressure for each of the overpressure demand cases is provided in Table 6.2. The potential overpressure approaches the burst pressure (defined as design margin times MAWP) of the protected equipment in cases where the overpressure demand case is not self-limiting. In other cases, such as a blocked discharge downstream of a centrifugal pump, the potential overpressure is self-limiting as it deadheads at the pump pressure of typically 1.3 times the normal discharge pressure of the pump.

Defining demand case overpressure scenarios for each PRD requires a thorough review of the unit pressure-relief study in conjunction with the piping and instrumentation diagrams (P&IDs). The review should be performed by qualified personnel with experience in the design and installation of pressure-relief systems.

The determination of the potential overpressure, P_o, due to a PRDs failure to open upon demand is generally a function of the following.

a) Type of Upstream Overpressure Source—Centrifugal pumps, steam supply headers, upstream pressure vessels, etc.

b) Upstream Source Pressures—Steam supply pressure, control valve upstream pressure, pressure from the high-pressure side of a heat exchanger, and deadhead pressure for centrifugal rotating equipment. Additionally, credit for PRDs on upstream equipment can be assumed to be available to limit overpressure.

c) Heat Sources, Types, and Temperatures—Blocked-in equipment, the heat source supplying energy to the system has a significant impact on the potential overpressure. Examples, a solar heat/energy supplied in a thermal relief scenario may result in flange leaks, limiting the overpressure to the normal operating pressure of the system. Alternatively, the overpressure may increase until a rupture occurs if the heat source is a fired heater (i.e. overpressure exceeding the burst pressure of the protected equipment). Other heat sources include steam reboilers to towers and the hot side of heat exchangers.

d) Fluid Bubble Point Pressure—Pressure increase is limited by the bubble point pressure of the contained process fluid at the temperature of the heat/energy source being supplied to the process.

6.2.1 Multiple Relief PRD Installations

The probability is reduced when multiple PRDs are used to manage the relief capacity required since the likelihood that multiple failures would occur is unlikely. In this case, the component POF is lower due to the expectation that some of the PRD capacity will be available on demand and minimize the overpressure experienced. When a component is protected by multiple PRDs, the calculated POFOD of each PRD in the multiple installation does not change. This multiple PRD installation adjustment factor, F_a, adjusts the overpressure that the component is likely to experience with a multiple PRD installation to minimize the potential overpressure.

$$F_a = \sqrt{\frac{A^{prd}}{A^{prd}_{total}}} \tag{5.89}$$

The F_a is a ratio of the area of a single PRD to the total area considering all PRDs in the multiple set up. The multiple PRD installation adjustment factor has a minimum reduction value of 0.25 since PRDs in a multiple PRD installation may have common failure modes. The final component overpressure is reduced by using Equation (5.90):

$$P^{comp}_{o,j} = F_a \cdot P_{o,j} \tag{5.90}$$

The reduced overpressure should be used when determining the protected component POF but is not used for calculating the overpressure factor, F_{op}.

6.2.2 Calculation Procedure

The following procedure is used to identify the potential PRD overpressure demand case scenarios.

a) STEP 1.1—Determine the list of overpressure scenarios applicable to the piece of equipment being protected by the PRD under evaluation. Table 5.3 provides a list of overpressure demand cases specifically covered. Additional guidance on overpressure demand cases and pressure-relieving system design is provided in API 521 [7].

b) STEP 1.2—Determine the Design Margin, DM, for the protected component material of construction.

c) STEP 1.3—For each overpressure demand case, estimate the amount of overpressure, $P_{o,j}$, likely to occur during the overpressure event if the PRD were to fail to open.

d) STEP 1.4—Calculate the total PRD orifice area, A^{prd}_{total}, for all PRDs in a multiple PRD installation.

e) STEP 1.5—Calculate the overpressure adjustment factor, F_a using Equation (5.89).

f) STEP 1.6—Calculate the final component overpressures determined in STEP 1.4 using Equation (5.90).

6.3 PRD POF

6.3.1 Definition

The POF calculations are performed for each overpressure demand case identified for the PRD according to Section 5.2. Failure of a PRD is defined as:

1. Failure to open during emergency or upset condition causing an overpressure of the protected component and resulting in loss of containment in failures/year.

2. Leakage through a PRD (Section 6.4).

6.3.2 Failure to Open

The calculation for the POF of a PRD failing to open is the product of an estimated overpressure demand case frequency (failures/demand), the probability of the PRD failing to open on demand (failures/demand), and the POF of the protected component at the overpressures.

A PRD protects equipment components from multiple overpressure scenarios. Guidance on overpressure demand cases and pressure relieving system design is provided in API 521 [7]. Each of these scenarios (fire, blocked discharge, etc.) may result in a multiple possible overpressure demand case scenarios, $P_{o,j}$. In addition, each overpressure demand case scenario has an associated demand rate, DR_j. Demand cases are discussed in more detail in Section 6.2, Table 6.2, and Table 6.3. The POF of the PRD failing to open for each overpressure demand case scenario is defined in Equation (5.91).

$$P_{f,j}^{prd} = P_{fod,j} \cdot DR_j \cdot P_{f,j} \tag{5.91}$$

Where j is the applicable overpressure demand case scenario for the PRD, $P_{f,j}^{prd}$.

The protected component POF, $P_{f,j}$, is a function of time and the potential overpressure. The individual parts for the POF of a PRD failing to open in Equation (5.91) is discussed in more detail in the following sections.

a) Section 5.3.3—PRD Demand Rate, DR_f

b) Section 5.3.4—PRD POFOD, $P_{fod,j}$

c) Section 5.3.5—POF of Protected Component as a Result of Overpressure, $P_{f,j}$

6.3.3 PRD Demand Rate

The first step in evaluating the POF of a PRD failing to open is to determine the expected demand rate (demands/year) placed on the PRD.

a) Default Initiating Event Frequencies

Estimated initiating event frequencies, EF_j, are provided based on the types of overpressure demand case scenario assigned. Examples of the initiating event frequencies are provided in Table 6.3 and the background on the default initiating event frequencies is provided in Table 6.2.

b) Credit for Other Layers of Protection

The actual PRD demand rate is not necessarily equal to the initiating event frequency. A Demand Rate Reduction Factor, $DRRF_j$, accounts for the difference in the overpressure demand case event frequency and the PRD overpressure demand rate.

Pressure vessels often contain control systems, high integrity protective instrumentation, shutdown systems, and other layers of protection to reduce the PRD demand rate. Credit can be taken for additional layers of protection, $DRRF_j$, or operator intervention for by the to reduce the probability of overpressure. The $DRRF_j$ may be determined rigorously for the installation as a result of a Layer of Protection Analysis (LOPA) or use the estimated value provided in Table 6.3.

An example using the $DRRF_j$ credit is for the fire overpressure demand case with an estimated initiating event frequency of 1 every 250 years (0.004 events/year). However, due to factors such as fire impinging on equipment rarely results a significant pressure increase that would cause the PRD to open. As a result, factors reducing the actual PRD demand rate, such as fire proofing, availability of other escape paths for the process fluid, and fire-fighting efforts at the facility may increase the $DRRF_j$.

c) Calculation of Demand Rate

The PRD demand rate, DR_j, is calculated as the product of the initiating event frequency and the $DRRF_j$, using Equation (5.92):

$$DR_j = EF_j \cdot DRRF_j \tag{5.92}$$

Where j is the applicable overpressure demand case scenario.

A PRD typically protects equipment from several overpressure demand case scenarios and each overpressure demand case has a unique demand rate. Default EF_j values for each of the overpressure cases are provided in Table 5.3. An overall demand rate on the PRD can be calculated in Equation (5.93):

$$DR_{total} = \sum_{j=1}^{ndc} DR_j \tag{5.93}$$

Additional guidance on overpressure demand cases and pressure relieving system design is provided in API 521 [7].

If the relief design basis of the PRD installation has not been completed, the list of applicable overpressure demand cases may not be available, and it may be more appropriate to use a simple overall average value of the demand rate for a PRD. An overall demand rate for a particular PRD may usually be estimated from past operating experience for the PRD.

d) Owner–~~User~~Operator Experience

The EF_j for the overpressure demand cases as shown in Table 6.3 are default values that may not be applicable in all situations. Owner–~~user~~operators may have operating experience with a particular process system that may warrant using other event frequencies. Additionally, a PRD that protects multiple components may experience an increased demand for a particular overpressure scenario. For example, a PRD located on a crude distillation tower may also protect the desalted preheat exchanger train. Since the PRD protects equipment encompassing a much greater area of the unit, an increase in the EF_j for the fire case may be appropriate. In general, where a PRD protects multiple components, the EF_j should be evaluated to determine if an increase is justified.

6.3.4 PRD POFOD

The next step is to determine the PRD POFOD in service.

a) Categories of Service Severity

PRD failure rates are directly related to the process severity of service. Categories of service are established for a PRD based on the process fluid tendency to result in a PRD failure caused by corrosion, fouling, plugging, or other effects. Temperature may also be a factor in determining the severity of service. The categories of service severity (mild, moderate, or severe) are associated with specific failure tendencies and default Weibull cumulative failure distribution curves, as described in Table 6.4.

It is important to note that a process fluid classified as mild service for POFOD is not necessarily a mild service for POL. For example, industry failure data show that cooling water, which is known to be a dirty/scaling service, has one of the highest POFOD rates and therefore may be classified as severe. Conversely, PRDs in cooling water service have not demonstrated a significant amount of POL failures and therefore may be classified as mild service for the POL. Steam service is another example where industry data indicate that steam should be classified as mild for a POFOD failure. Steam is classified as severe for a POL failure since steam is known to cause PRD leaks due to erosion of high-temperature steam.

b) Default POFOD vs Time in Service

1) General

Table 6.5 provides the default Weibull parameters for failure to open for conventional spring-loaded PRVs, balanced bellows PRVs, pilot-operated PRVs, and rupture disks. The Weibull parameters provided in Table 6.5 were determined using industry failure rate data with the majority of the available data from successful performance during the PRD service interval. The successful service test points are referred to as suspensions and were included with the failure data in determination of the Weibull parameters.

Weibull parameters are provided for the three categories of PRD service severity (mild, moderate, or severe), as discussed in Section 6.3.4a. The Weibull parameters provide the default POFOD curves for each of the PRD types listed in Table 6.5 when used in the Weibull cumulative failure density function, $F(t)$, in Equation (5.90). For example, Figure 6.2 provides the default Weibull cumulative failure distribution curves used for spring-loaded conventional PRVs using the Weibull function to describe the three categories of service severity.

NOTE: that the units for the POFOD data presented in Figure 6.2 are failures/demand since the data were established from actual PRD bench test results rather than continuous service data. POFOD should not be confused with POF (failures/year) that includes the demands on the PRD (see Section 6.2) and the probability that the protected component will fail in an overpressure event (see Section 6.3.5).

The cumulative failure distribution curves shown in Figure 6.2 and the Weibull parameters presented in Figure 6.6 are the default values based on the category of service severity of the PRD being evaluated. These base values are defaults and should be replaced with owner~~–user~~operator site-specific data, if available (Section 6.3.4c3).

2) Presence of an Upstream Rupture Disk

Rupture disks are often installed in combination with PRVs to isolate the PRV from process conditions that could cause corrosive or fouling fluids and reduces the POFOD. API 520, Parts 1 and 2 provide additional information related to the use and installation of rupture disks upstream of PRVs.

A mild service for POFOD is recommended for a PRD with an upstream rupture disks, regardless of the process fluid severity. Assigning a mild POFOD service assumes that the space between the rupture disk and the PRV is vented and monitored for leakage, as required by Code and recommended by API 520. If the space is not vented and monitored for leakage, credit for an upstream rupture disk should not be given.

3) Use of Plant-specific Failure Data

Data collected from specific plant testing programs may be used to for POFOD and POL analysis. MTTF or failure per million operating hours may be calculated in the required format using simple conversion routines.

c) Default Data for Balanced Bellows PRVs

Balanced spring-loaded PRVs contain a bellows to isolate the back side of the disk from the effects of superimposed and built-up back pressure. The bellows isolates the PRD internals from a corrosive process fluid in the discharge system. Analysis of industry failure rate data indicates that balanced bellows PRVs have the same POFOD rates as conventional PRDs. Though bellows PRVs typically discharge to dirty/corrosive closed systems, due to the isolation of the PRV internals from the discharge fluid and the effects of corrosion and fouling. As shown in Table 6.6, the η characteristic life for bellows PRVs is the same as for conventional PRVs.

d) Default Weibull Parameters for Pilot-operated PRVs

To date, there is little failure rate data in the industry available for pilot-operated PRVs. One source [10] indicates that pilot-operated PRVs are 20 times more likely to fail than their spring-loaded counterparts. The Weibull parameters for the POFOD curves for conventional PRVs in Table 6.5 are used as the basis for pilot-operated PRVs with adjustment factors applied to the η characteristic life. For mild service, the η characteristic life for pilot-operated PRVs is reduced by a factor of 1.5; for moderate service, the reduction factor is 3.0; and for severe service, the reduction factor is 5.0.

e) Default Weibull Parameters for Rupture Disks

To date, there is little failure rate data in the industry available for rupture disks. Rupture disks are simple, reliable to use. Rupture disks ~~and~~ open at ~~pressures significantly over their~~or near burst pressure unless ~~provided~~ the inlets or outlets are plugged, or the disk is installed improperly. Failure of rupture disks are typically due to burst prematurely. The Weibull parameters for POFOD for rupture disks are based on the mild severity curve for conventional PRVs and assuming that a rupture disk is at least as reliable as a conventional PRV. The default parameters also assume that the rupture disk material is resistant to the process fluid corrosion. If the rupture disk material is resistant to the process fluid corrosion, the disk Weibull parameters should be adjusted accordingly.

f) Adjustment for Conventional PRVs Discharging to Closed System

An adjustment factor is used to modify the base Weibull parameters for conventional PRVs discharging to a closed system or to flare. A conventional PRV η characteristic life is reduced by 25% since no bellows is present to protect the bonnet housing from discharge system corrosion.

$F_c = 0.75$ for conventional valves discharging to closed system or flare

$F_c = 1.0$ for all other cases

g) Adjustment for Environmental Factors

Environmental and installation factors that affect the reliability of PRDs include installed piping vibration, a history of chatter, or pulsing flow or cyclical service (downstream of reciprocating rotating equipment).

Other environmental factors that can significantly affect POL are operating temperature and operating ratio. The PRD operating ratio is the ratio of maximum system operating pressure to the set pressure. When the operating ratio is greater than 90% for spring-loaded PRVs, the system pressure is close to overcoming the closing force provided by the spring on the seating surface and the PRV will be more likely to leak (simmer). The increased potential for leakage is considered by applying an environmental factor to the default leakage curve. Similarly, an environmental factor is applied when the operating margin is greater than 95% for pilot-operated PRVs.

NOTE: some pilot-operated PRVs can function at operating ratios up to 98% (see API 520 for guidance on operation margin).

Analysis of the industry failure rate data shows that PRDs in vibration or cyclical service generally experience higher leakage rates but POFOD rates are not significantly affected.

PRVs in service with a history of chattering should be redesigned or modified to eliminate the chatter, as soon as possible. An adjustment factor of 0.5 is applied to the Weibull η parameters for the POFOD and POL curves of a PRD experiencing chattering in service since the effects of chatter are detrimental to the protection provided by the PRD.

Table 6.6 provides the environmental adjustment factors applied to the default POFOD and POL Weibull curves. The environmental factor, F_{env}, increases the POFOD or POL (shifting the probability curves to the left) by reducing the curve's η characteristic life, as shown in Figure 6.5.

h) Updating POFOD Based on PRD-specific Inspection and Test Data

1) Tracking Historical Inspection and Test Data

An inspection program should track each PRD's testing and inspection history from its initial installation. Adjustments to the PRD POFOD, P_{fod} ,and POL, P_l curves are made to provide credit for information during a PRD inspection and test.

Data obtained from a PRD inspection and test will increase or decrease the POFOD and POL by modifying the Weibull parameters based on the pass/fail and no leak/leak test results for the service duration, $t_{dur,i}$, since the last inspection. An increase or decrease in the POFOD and POL through inspection will increase or decrease the recommended inspection and test interval.

Modifying the POFOD based on test results alone (i.e. bench test) will be nonconservative if the inlet or outlet piping was plugged during operation, affecting the operating of the PRD. The visually inspected condition of the piping should be documented for each inspection and specifically noted if the piping is plugged. Plugged PRD piping should be considered to have failed the inspection and test, regardless of the bench test results or inspection method used. More than 25% of the pipe plugged is considered plugged and the PRD should fail the inspection and test.

2) Effectiveness of Inspection Programs in Confirming Failure Rates

Inspection effectiveness is based on its ability to adequately predict the pass/fail condition of the PRD and detect/quantify damage. Definitions for PRD inspection and test effectiveness are provided in Part 2, Annex 2.C, Table 2.C.3.1.

PRD inspection and test should document the effectiveness of the inspection and test performed. The inspection effectiveness concept is described in Part 2, Section 3.4.3. for fixed equipment is similar for PRDs. In addition, PRD inspection effectiveness measures the confidence in the pass/fail/leak result of the inspection and test.

Table 6.7 provides default conditional probabilities based on expert opinion. The conditional probabilities indicate the ability of the inspection and test to reflect an accurate representation of the PRD performance in an overpressure event. For example, a 90% effectiveness associated with passing a highly effective inspection and test indicates that there is a 90% chance that the PRD would perform as intended in service. Conversely, there is a 10% chance that the PRD would fail to perform as intended in service.

The conditional probabilities in Table 6.7 assign the highest confidence to a PRD passing a bench tested without any prior cleaning (i.e. as-received condition). Bench testing of PRDs that were cleaned prior to testing or testing in situ, as well as visual inspections provide information for expected PRD performance in service but are not considered as reliable as the as-received bench test.

PRDs that fail an inspection and test are treated differently than passed test results. For PRDs that fail a highly effective bench test, the 95% confidence indicates a 95% chance that the PRD would have failed to perform as intended in service. A usually effective ~~in~~ bench test or test in situ after the PRD was steamed is assigned a 95% confidence the PRD will fail to perform as intended in service.

An ineffective test does not provide any additional information about the ability of the PRD to perform as intended in service and receives ~~s~~no inspection and test credit. Credit is provided for an overhauled PRD and is returned to service in like-new condition. In this case the service duration, $t_{dur,i}$, is calculated based on the date of the ineffective inspection and test.

3) Inspection Updating

The initial default Weibull parameters for the listed provided process fluid services are modified as inspection and test data is provided.

The Bayesian updating approach used assumes that the Weibull β shape parameter remains constant based on historical data and modifies the η characteristic life based no available inspection data. This is analogous to evaluating a one-parameter Weibull to update the PRD performance. Bayes' Theorem works best when the error rates for a test are very small; however, test effectiveness in Table 6.8 vary widely from 50% to 90%. As a result, using Bayes' Theorem high levels of uncertainty generates an unrealistically high adjusted POF, particularly for a pass bench test result. A modified inspection updating method was developed to provide more realistic modification approach to characteristic life.

A default POFOD is defined for the PRD based on service duration, $t_{dur,i}$, at the time of inspection to provide a POFOD vs. time. The methodology calculates a prior PRD POFOD (prior to inspection) using Equation (5.94).

$$P_{f,prior}^{prd} = 1 - exp\left[-\left(\frac{t}{\eta_{mod}}\right)^{\beta}\right] \tag{5.94}$$

The prior probability that the PRD will operate on demand (pass) is calculated using Equation (5.95).

$$P_{p,prior}^{prd} = 1 - P_{f,prior}^{prd} \tag{5.95}$$

A PRD POFOD posterior probability is calculated based on the conditional probability, or confidence factor, CF, from Table 6.7 after an inspection of a specific effectiveness is performed. The updated POFOD is the conditional POFOD and is calculated using Equation (5.96) or Equation (5.97) depending on the inspection and test result.

The conditional PRD POFOD, $P_{p,cond}^{prd}$,for a passed inspection is calculated using Equation (5.96).

$$P_{p,cond}^{prd} = \left(1 - CF_{pass}\right) \cdot P_{p,prior}^{prd} \tag{5.96}$$

The conditional PRD POFOD, $P_{f,cond}^{prd}$,for a failed inspection is calculated using Equation (5.97).

$$P_{f,cond}^{prd} = CF_{fail} \cdot P_{f,prior}^{prd} + \left(1 - CF_{pass}\right) \cdot P_{p,prior}^{prd} \tag{5.97}$$

Weighted equations were developed to increase credit for inspection and test conducted later in the characteristic life. The posterior POFOD, $P_{f,wgt}^{prd}$, is calculated using the weighted prior and conditional probability equations provided in Table 6.9.

The updated η characteristic life is calculated using Equation (5.101) based on the service duration, $t_{dur,i}$, of the PRD, the known β shape parameter, and $P_{f,wgt}^{prd}$.

The weighted equations produce a gradual shift from default POFOD data to PRD-specific POFOD data with a gradual increasing η characteristic life. A significantly shorter η characteristic life results if the PRD inspection and test has resulted in repeated failures early in the service.

Additional inspection and test updating guidance are:

i) Tests conducted less than 1 year apart should not be credited.

ii) The η characteristic life cannot decrease after a pass inspection and test result – If the methodology decreases the η characteristic life, the prior probability should be used for the η characteristic life.

iii) The η characteristic life cannot increase after a fail inspection and test result – If the methodology increases the η characteristic life, the prior probability should be used for the η characteristic life.

4) Updating Failure Rates After Modification to the Design of the PRD

Design changes that improve the PRD performance may result in a failure rate change, such as upgrading to a corrosion-resistant material or installation of an upstream rupture disk. Past inspection data are no longer applies after PRD design changes. A new default curve should be selected based on Figure 6.2 or PRD-specific Weibull parameters should be defined based on owner–~~user~~operator experience (generating a unique PRD curve) should be used after PRD design changes.

i) Adjustment for Overpressures Higher Than Set Pressure

As discussed in Section 6.1.2, the POFOD curves are based on bench test data where a failure is defined as any test requiring a pressure greater than 1.3 times the set pressure. Industry failure data supports that as ratio of overpressure increases, the POFOD decreases, as shown in Figure 6.4.

A conservative assumption decreases the operating failure rate, $F_{op,j}$, by a factor of 5 at an overpressure of 4.0 times the set pressure and linearly interpolate between 1.3 and 4.0 at an overpressure times the set pressure, shown in Equation (5.98).

$$F_{op,j} = 1.0 \qquad \text{for } \frac{P_{o,j}}{P_{set}} < 1.3$$

$$F_{op,j} = 0.2 \qquad \text{for } \frac{P_{o,j}}{P_{set}} > 4.0 \tag{5.98}$$

$$F_{op,j} = 1 - \frac{1}{3.375} \cdot \left(\frac{P_{o,j}}{P_{set}} - 1.3 \right) \qquad \text{for all other cases}$$

The $F_{op,j}$ adjustment factor ranges from 0.2 and 1.0.

The overpressure factor, $F_{op,j}$, is an adjustment for overpressure scenarios higher than 1.3 times the set pressure where j is the overpressure demand case scenario.

6.3.5 Protected Equipment Failure Frequency as a Result of Overpressure

A damage adjusted POF for components evaluated with RBI is included in the PRD POF calculation (Section 6.1). The component DF increases as a function of time and is calculated based on the applicable damage mechanisms for the equipment, the inspection history, and condition of the equipment. As the PRD inspection interval is extended, the component damage continues, and the risk increase as well as the risk of the PRD over time.

a) Damage Factor Calculation Procedure for PRD with Fixed Equipment

The damage adjusted POF that is calculated at the normal operating pressure of the component is adjusted when evaluating PRDs. When a PRD fails to open on demand, the protected component pressure exceeds the normal operating pressure and may significantly exceed the MAWP. Equation (5.99) is used to calculate the protected component damage POF based on the expected pressure for each overpressure demand case. The damage adjusted component POF, $P_{f,j}$, is the probability of a loss of containment of the protected component as a result of the overpressure event.

$$P_{f,j} = min\left(\left(a \cdot D_f \cdot F_{MS}\right) \cdot e^{\left(b \cdot \frac{P_{o,j}}{MAWP}\right)}, 1.0\right) \tag{5.99}$$

Where a and b are the constants from Table ~~5~~6.11 for $P_{f,j}$ are based on the Design Margin, DM, from Table ~~5~~6.10 for the protected component material of construction determined in STEP 1.2.

During PRD overpressure events, the probability of loss of containment in the protected component increases. An undamaged component $\left(D_f = 1\right)$ has an upper limit probability of loss of containment of 1.0 when the overpressure is equal to the burst pressure (the expected failure pressure of the component). The burst pressure of the component is estimated using the design margin times the MAWP (with design margins for components constructed in accordance with various codes are shown in Table 6.10. Alternatively, the burst pressure can be more accurately calculated using a more advanced analysis such as Svensson's method [11]. For damaged components $\left(D_f >> 1\right)$, the probability of loss of containment of 1.0 may occur at pressures much lower than the damaged component burst pressure, (see Figure 6.6).

b) Selection of DF Class when PRD RBI Is Performed Without Fixed Equipment

The D_f for the protected component may be specified using a DF class defined in Table 6.12 if a fixed equipment RBI study is not available. This D_f assignment is more qualitative than when an RBI analysis conducted to determine component D_f.

6.3.6 Calculation Procedure

The following calculation procedure may be used to determine the probability of a PRD failing to open.

a) STEP 2.1—Grade the PRD inspection and test histories for each inspection using Part 2, Annex 2.C, Table 2.C.3.1 for guidance. Grade each inspection as Pass/Fail and No Leak/Leak, assign the confidence factors, CF_i, and calculate the time duration, $t_{dur,i}$.

   1) STEP 2.1.1—Grade each inspection and test Part 2, Annex 2.C, Table 2.C.3.1.

   2) STEP 2.1.2—Record the inspection and test result as Pass/Fail and assign the appropriate CF_i.

   3) STEP 2.1.3—Calculate the service duration, $t_{dur,i}$, for each inspection.

   4) STEP 2.1.4—Determine if the PRD was overhauled. If the PRD was overhauled, the date of the most recent overhaul is the date to be used in STEP 2.7 (Figure 6.7). The owner-operator may consider using the return to service date instead of the overhaul date to calculate the next inspection date. The owner-operator should define the timeframe (delay between overhaul and in-service date) for when the return to service date should be used. Refer to Section 5.1.4 and Section 5.1.5 for more information for acceptable time delays before a retest should be considered.

b) STEP 2.2—Select the most recent inspection and test history and service duration, $t_{dur,i}$.

c) STEP 2.3—Determine the default values for the Weibull parameters, β and η_{def}, based on category of service severity (Section 6.3.4a), selection of the default POFOD curve (Section 6.3.4c), type of PRD (Sections 6.3.4c through 6.3.4e) and using Table 6.5 and Table 6.6.

d) STEP 2.4—Determine the adjustment factor, F_c, for conventional PRDs discharging to a closed system or flare (Section 6.3.4f).

e) STEP 2.5—Determine the environmental adjustment factor for conventional PRDs, F_{env}, using Table 6.6.

f) STEP 2.6—Calculate the modified characteristic life, η_{mod}, using Equation (5.100), η_{def} from STEP 2.3, and F_c STEP 2.4.

$$\eta_{mod} = F_c \cdot F_{env} \cdot \eta_{def} \tag{5.100}$$

g) STEP 2.7—Calculate the updated characteristic life, η_{upd}, using η_{mod} from STEP 2.6 and PRD inspection and test history from STEP 2.1.

1) STEP 2.7.1—Calculate the prior probability of failure, $P_{f,prior}^{prd}$, using Equation (5.94) and the time period, $t_{dur,i}$, from STEP 2.6.

2) NOTE: for the first inspection record, η_{mod}, from STEP 2.1 is used with subsequent inspection records using η_{upd} from STEP 2.7.6.

3) STEP 2.7.2—Calculate the prior probability of passing, $P_{p,prior}^{prd}$, using Equation (5.95).

4) STEP 2.7.3—Determine the conditional probability of pass test result, $P_{p,cond}^{prd}$, using Equation (5.96).

5) STEP 2.7.4—Determine the conditional probability of failed test result, $P_{f,cond}^{prd}$, using Equation (5.97).

6) STEP 2.7.5—Calculate the weighted POF, $P_{f,wgt}^{prd}$, using the equations in Table 5.9.

7) STEP 2.7.6—Calculate the η_{upd} using Equation (5.101) using Weibull parameters β from STEP 2.3, and the weighted POF, $P_{f,wgt}^{prd}$, established in STEP 2.7.5.

$$\eta_{upd} = \left(\frac{t_{insp}}{\left(-\ln\left(1 - P_{f,wgt}^{prd}\right)\right)^{\frac{1}{\beta}}} \right) \tag{5.101}$$

8) STEP 2.7.7—Repeat these steps for each of the inspection records available for the PRD to calculate the final η_{upd}.

h) STEP 2.7.8—Calculate the POFOD as a service duration, $t_{dur,i}$, for the PRD using Equation (5.102) and η_{upd} from STEP 2.7.7.

$$P_{fod} = 1 - exp\left[-\left(\frac{t_{dur,i}}{\eta_{upd}} \right)^{\beta} \right] \tag{5.102}$$

i) STEP 2.8—For each overpressure scenario, determine the overpressure adjustment factor, $F_{op,j}$, using Equation (5.98).

j) STEP 2.9—Calculate the adjusted POFOD using Equation (5.103) and $F_{op,j}$ from STEP 2.8.

$$P_{fod,j} = P_{fod} \cdot F_{op,j} \tag{5.103}$$

k) STEP 2.10—For each overpressure demand case, determine the initiating event frequency, EF_j, using Table 6.3 or based on owner–~~user~~operator experience for the overpressure demand case.

l) STEP 2.11—Determine the demand rate reduction factor, $DRRF_j$, accounting for layers of protection that may reduce the probability of an overpressure of the protected component, see Section 6.3.3b and Table 6.3 for guidance.

m) STEP 2.12—For each overpressure demand case, determine the demand rate, DR_j, placed on the PRD, using Equation (5.92).

n) STEP 2.13—Determine the MAWP of the protected equipment.

o) STEP 2.14—Calculate the protected component damage adjusted DF, D_f. The DF should be determined at the PRD service duration, $t_{dur,i}$, from STEP 2.2 for a DF as a function of time. If a fixed equipment RBI analysis has not been completed, the DF may be estimated using Table 6.12.

p) STEP 2.15—Calculate the protected component POF at the overpressure, $P_{f,j}$, using Equation (5.99) and the overpressure is determined in STEP 1.3 of Section 6.2.2.

q) STEP 2.16—Calculate the PRD POF, $P_{f,j}^{prd}$, using Equation (5.91) using $P_{fod,j}$ from STEP 2.9, DR_j from STEP 2.12, and $P_{f,j}$ from STEP 2.15.

r) STEP 2.17—Repeat STEP 2.1 through STEP 2.16 for each component protected by the PRD.

6.4 Probability of Leakage (POL)

6.4.1 Overview

The POL case is a function of failure during continuous operation. Industry data associated with POL, P_l , is presented in failures/year with not impacted by demand rate.

a) Categories of Service Severity

Guidance on selecting the proper service severity for the POL case is provided in Table 6.13. The owner–~~user~~operator's experience with a PRD in a particular service provides guidance for selecting the severity.

b) Default POL Rates vs. Time in Service

A set of Weibull curves tracks provided for the POL case are based on from data of PRDs in continuous service (i.e. a continuous demand). The data are collected in units of failures/year and are not modified by a demand rate. Table 6.14 provides the default PRD POL vs. time information using a Weibull function to describe the three types of service (mild, moderate, and severe). These data are currently based on a limited amount of industry data and should be supplemented by owner–~~user~~operator data where available.

The default cumulative POL distribution curves for spring-loaded conventional PRVs using the Weibull function to describe the three categories of service severity are provided in Figure 6.3 as an example.

c) Default Weibull Parameters for Balanced Bellows PRVs

The Weibull parameters for the POL curve for balanced bellows PRVs provided in Table 6.14 match the industry failure rate data. These data reflect a minor increase in the POL compared to conventional PRVs.

d) Default Weibull Parameters for Pilot-operated PRVs

The design of pilot-operated PRVs provide a better seal as the operating pressure approaches the set pressure. Owner–~~user~~operator Weibull parameters for conventional or pilot-operated PRVs should be used, if available, until improved failure rate data are developed for η characteristic life for leakage provided in Table 6.14.

e) Default Weibull Parameters for Rupture Disks

Since no industry data are available for rupture disk leakage, Weibull parameters are based on the mild severity curve for conventional PRVs (see Section 6.3.4e for additional information).

f) Adjusted Default POL Curve for PRVs Containing Soft Seats

Soft seats (O-rings) are often added to spring-loaded PRVs to reduce the potential for leakage across the seat. When a conventional or balanced bellows PRV contains a soft seat design, the η parameter for the default POL Weibull curve is increased by a factor of 1.25 in accordance with the following factors:

$F_s = 1.25$ for soft seated designs

$F_s = 1.0$ for all other cases

g) Environmental Modifiers to the Default POFOD and POL Data

Table 6.6 provides environmental adjustment factors, F_{env} , for the POL Weibull curves (Section 6.3.4g)

h) Set Pressure Adjustment

The POL decreases as the ratio of operating pressure to set pressure, $\frac{P_s}{P_{set}}$, decreases, as shown in Table 6.8.

i) Presence of an Upstream Rupture Disk

The POL is negligible (i.e. $P_l^{prd} = 0.0$) and the COF=0 with a rupture disk installed upstream of the PRV.

j) Modification of Leakage Rates Based on PRD Specific Inspection and Test Data

The characteristic life updating based on inspection and test history is the same as the approach described in Section 6.3.4h for the POFOD case.

6.4.2 POL Calculation Procedure

The PRD POL is calculated using the following steps.

a) STEP 3.1—Determine default Weibull parameters, β and η_{def}, based on category of service severity and PRD type (Section 6.3.4a through Section 6.3.4i).

b) STEP 3.2—Apply an adjustment factor, F_s, for the presence of soft seats (Section 6.3.4j).

c) STEP 3.3—Apply an adjustment factor, F_{env}, for environmental factors (Section 6.3.4j).

d) STEP 3.4—Calculate the modified characteristic life, η_{mod}, using Equation (6.104).

$$\eta_{mod} = F_s \cdot F_{env} \cdot \eta_{def} \tag{5.104}$$

e) STEP 3.5—Calculate the updated characteristic life, η_{upd}, using η_{mod} from STEP 3.4 and PRD inspection and test history from STEP 2.6.

1) STEP 3.5.1—Calculate the prior probability of leak, $P_{f,prior}^{prd}$, using Equation (5.94) and the time period, $t_{dur,i}$, from STEP 2.6.

NOTE: for the first inspection record, η_{mod}, from STEP 2.1 is used with subsequent inspection records using η_{upd} from STEP 3.5.6.

2) STEP 3.5.2—Calculate the prior probability of no leak, $P_{p,prior}^{prd}$, using Equation (5.95).

3) STEP 3.5.3—Determine the conditional probability of no leak test result, $P_{p,cond}^{prd}$, using Equation (5.96).

4) STEP 3.5.4—Determine the conditional probability of leak test result, $P_{f,cond}^{prd}$, using Equation (5.97).

5) STEP 3.5.5—Calculate the weighted POF, $P_{f,wgt}^{prd}$, using the equations in Table 6.9.

6) STEP 3.5.6—Calculate η_{upd} using Equation (5.101) using Weibull parameters β from STEP 2.3, and the weighted POF, $P_{f,wgt}^{prd}$, established in STEP 3.5.5.

7) STEP 3.5.7—Repeat these steps for each of the inspection records available for the PRD to calculate the final η_{upd}.

f) STEP 3.6—Calculate the set pressure factor, F_{set}, based on the PRD type, operating pressure, P_s, and set pressure, P_{set} (see Table 6.8).

g) STEP 3.7—Calculate the updated characteristic life, η_{upd}, from STEP 3.5.7 using Equation (5.105).

$$P_l^{prd} = 1 - exp\left[-\left(\frac{t_{dur,i}}{\eta_{upd}}\right)^{\beta}\right] \cdot F_{set} \tag{5.105}$$

6.5 PRD Consequence of Failure to Open on Demand (COFOD)

6.5.1 General

The Consequence of Failure to Open On Demand (COFOD) calculates for event outcomes such as fires, explosions, and toxic exposure are described in Part 3. A PRD failure to open on demand will result in the protected component being exposed to significantly higher pressures than during normal operation. The PRD COFOD calculates the impact of each demand case scenario failure at the overpressure.

Table 6.15 shows the expected potential consequences on a pressure vessel as a percent of overpressure above the MAWP. Table 6.15 is provided as a qualitative discussion of the potential risks to pressure vessels due to an overpressure and is not intended to indicate any specific event outcome. The methodology accounts for the effects of overpressure on protected equipment by increasing the probability of loss of containment. At an overpressure equal to the burst pressure (estimated to be the design margin times the MAWP), the probability of loss of containment is conservatively assumed to be equal to 1.0 (Section 6.3.4i).

The COFOD, $C_{f,j}^{prd}$, is calculated for each overpressure demand case scenario as follows:

a) STEP 4.1—For each overpressure demand case, calculate the financial COFOD, $C_{f,j}^{prd}$, for the protected component using the overpressure from STEP 1.6 and methodology in Part 3.

6.6 Consequence of Leakage (COL)

6.6.1 General

The PRD consequence of leak is typically less significant than a component loss of containment resulting from a PRD COFOD. While the frequency of leakage is less significant, a leak may result in a high risk ranking of the PRD.

The COL, C_l^{prd}, from PRDs is calculated by summing the following costs and using Equation (5.106):

$$C_l^{prd} = Cost_{inv} + Cost_{env} + Cost_{sd} + Cost_{prod} \quad (5.106)$$

- Lost inventory cost based on the product of the cost of fluid, the leakage rate (Section 5.4), and the estimated number of days to discover the leak (Table 6.15).
- Regulatory and environmental costs associated with leakage.
- Downtime cost to repair or replace the PRD if a leaking or stuck open PRD cannot be tolerated.
- Production cost while conducting the repair or replacement of the leaking PRD.

For a multiple PRD installation, the POL for any one specific PRD does not increase. However, since the number of PRDs increases, the POL and the associated consequences increases in proportion to the number of PRDs protecting the system.

6.6.2 Estimation of PRD Leakage Rate

Analysis of industry bench test data indicates approximately 8.4% of PRVs tested leaked during a bench test between 70% and 90% of the set pressure, 6.6 % of PRVs leaked at pressures below 70% of the set pressure and an additional 2.4 % of PRVs leaked significantly below 70% of their set pressure. A summary of the leakage rates used for the consequence calculation is provided in Table 6.17.

A leakage rate of 1% of the PRD rated capacity, W_c^{prd}, (calculated at normal operating conditions) was used for minor or moderate leaks and calculated using Equation (5.107). A minor or moderate leakage, C_l^{mild}, represents 90% of all of potential leakage cases, as shown in Table 6.17.

$$lrate_{mild} = 0.01 \cdot W_c^{prd} \quad (5.107)$$

The leakage rate for a stuck open or spurious leaks is assumed to be 25% of the PRD rated capacity, W_c^{prd} and calculated using Equation (5.108). A leak from a stuck open PRD, C_l^{so}, represents 10% of all potential leakage cases.

$$lrate_{so} = 0.25 \cdot W_c^{prd} \quad (5.108)$$

The rated capacity of the PRD, W_c^{prd}, can usually be found on the PRD datasheet. It can also be calculated using the methods presented in API 520, Part 1 [8].

6.6.3 Estimated Leakage Duration

The leakage duration, D_{mild}, is calculated mild or moderate leakage, as shown in Table 6.14, assuming that mild leakage from larger PRDs will be discovered sooner than leakage from smaller PRDs. The leakage duration for the stuck open case is calculated using Equation (5.109), assuming that an immediate PRD repair is required with an isolation time of 30 minutes.

$$D_{so} = \frac{30 \text{ min}}{60 \text{ min/hr} \cdot 24 \text{ hr/day}} = 0.021 \text{ days} \quad (5.109)$$

6.6.4 Credit for Recovery of Leaking Fluid

The cost of lost inventory is not considered to be as severe when the unit has a flare recovery system installed or the discharge from the PRD is to a closed system. A recovery factor, F_r, based on the discharge location of the PRD as follows:

$F_r = 0.5$ if the PRD discharges to flare and a flare recovery system is installed

$F_r = 0.0$ if the PRD discharges to a closed system

$F_r = 1.0$ for all other cases

6.6.5 Lost Inventory Cost

The cost of lost fluid inventory, $Cost_{inv}$, is calculated using Equation (5.110) or Equation (5.111) from mild or stuck open leaks. When calculating the consequence of leakage, the fluid cost, $Cost_{flu}$, is based on the process fluid at the PRD physical location.

$$Cost_{inv}^{mild} = 24 \cdot F_r \cdot Cost_{flu} \cdot D_{mild} \cdot lrate_{mild} \tag{5.110}$$

$$Cost_{inv}^{so} = 24 \cdot F_r \cdot Cost_{flu} \cdot D_{so} \cdot lrate_{so} \tag{5.111}$$

6.6.6 Environmental Cost

The environmental cost, $Cost_{env}$, is calculated when PRD leakage is released to the atmosphere or a flare system and may require cleanup costs or results in regulatory fines.

6.6.7 Shutdown for Repair PRD Cost

The cost associated with repair and maintenance, $Cost_{sd}$, is calculated if a leaking PRD cannot be tolerated, by using the following costs:

$Cost_{sd} = \$1000$ for PRDs < NPS 6 inlet size

$Cost_{sd} = \$2000$ for PRDs $\geq$ NPS 6 inlet size

It is recommended that actual owner–~~user~~operator work order costs be used that are associated with maintenance, inspection and test, and repair of the PRD.

6.6.8 Lost Production Cost

The cost of lost production, $Cost_{prod}$, to repair a leaking PRD is calculated using Equation (5.112) or Equation (5.113). Production losses are not considered when spare PRDs are installed in parallel or in cases where isolation valves underneath the PRD offer flexibility to repair without shutting down. For the stuck open case, it is assumed that prolonged operation cannot be tolerated and the production cost is calculated using Equation (5.114).

$$Cost_{prod}^{mild} = 0.0 \quad \text{if a leaking PRD can be tolerated or if the PRD can be isolated and repaired without requiring a shutdown} \tag{5.112}$$

$$Cost_{prod}^{mild} = Unit_{prod} \cdot D_{sd} \quad \text{if a leaking PRD cannot be tolerated} \tag{5.113}$$

$$Cost_{prod}^{so} = Unit_{prod} \cdot D_{sd} \quad \text{for a stuck open PRD} \tag{5.114}$$

6.6.9 Calculation of Final Leakage Consequence

The final leakage consequence is calculated for the two leaks cases discussed above.

a) Minor or Moderate Leakage

The final consequence of the minor or moderate leakage, $Cost_l^{mild}$, is calculated using Equation (5.115).

$$Cost_l^{mild} = Cost_{inv}^{mild} + Cost_{env} + Cost_{sd} + Cost_{prod}^{mild} \tag{5.115}$$

b) Stuck Open Leakage

The final consequence of the stuck open leak case, $Cost_l^{so}$, is calculated using Equation (5.116).

$$Cost_l^{so} = Cost_{inv}^{so} + Cost_{env} + Cost_{sd} + Cost_{prod}^{so} \tag{5.116}$$

c) Final Leakage Consequence

The final total leakage weighted consequence is calculated using Equation (5.117).

$$C_l^{prd} = 0.9 \cdot Cost_l^{mild} + 0.1 \cdot Cost_l^{so} \tag{5.117}$$

6.6.10 COL Calculation Procedure

The following procedure may be used to determine the PRD COL.

a) STEP 5.1—Determine the flow capacity of the PRD, W_c^{prd} from the PRD datasheet or calculated using the methods presented in API 520, Part 1 [8].

b) STEP 5.2—Calculate the minor or moderate leakage rate, $lrate_{mild}$, using Equation (5.107) and the rated capacity of the PRD obtained in STEP 6.1.

c) STEP 5.3—Calculate the stuck open leakage rate, $lrate_{so}$, using Equation (5.108) and the rated capacity of the PRD obtained in STEP 6.1.

d) STEP 5.4—Determine the leakage duration, D_{mild}, using Table 6.16.

e) STEP 5.5—Determine the stuck open duration, D_{so}, using Equation (5.109).

f) STEP 5.6—Calculate the cost of lost inventory for leakage, $Cost_{inv}^{mild}$ using Equation (5.110), recovery factor, F_r, from Section 5.6.4, and based on the PRD discharge location and discharge location.

g) STEP 5.7—Calculate the cost of lost inventory for stuck open, $Cost_{inv}^{so}$, using Equation (5.111), recovery factor, F_r, from Section 5.6.4, and based on the PRD discharge location.

h) STEP 5.8—Determine the environmental consequence associated with PRD leakage, $Cost_{env}$.

i) STEP 5.9—Determine the consequence associated with repair and maintenance of the PRD, $Cost_{sd}$. Default values based on PRD size are given in Section 6.6.7 or actual owner–~~user~~operator costs may be used.

j) STEP 5.10—Calculate the cost of lost production for mild leaks, $Cost_{prod}^{mild}$, using Equation (5.112) or Equation (5.113) based on whether or not PRD leakage can be tolerated and the ability to isolate and repair a leaking PRD without a unit shutdown.

k) STEP 5.11—Calculate the costs of lost production for the stuck open case, $Cost_{prod}^{so}$, using Equation (5.114).

l) STEP 5.12—Calculate the final consequence associated with mild leakage, $Cost_l^{mild}$, using Equation (5.115).

m) STEP 5.13—Calculate the final consequence associated with a stuck open PRDs, $Cost_l^{so}$, using Equation (5.116).

n) STEP 5.14—Calculate the total final leakage consequence, C_l^{prd}, using Equation (5.117).

6.7 Risk Analysis

6.7.1 Failure to Open on Demand Risk

The calculation of risk for a PRD failing to open at a specified service duration, $t_{dur,i}$ is calculated for each applicable overpressure demand case scenario using Equation (5.118).

$$Risk_{f,j}^{prd} = P_{f,j}^{prd} \cdot C_{f,j}^{prd} \tag{5.118}$$

The overall risk for the fail to open case is calculated by the sum of the risks for each overpressure demand case scenario using Equation (5.119).

$$Risk_f^{prd} = \sum_{j=1}^{ndc} P_{f,j}^{prd} \cdot C_{f,j}^{prd} \tag{5.119}$$

Where, j represents each of the number of overpressure demand case scenarios, ndc.

If the PRD protects multiple components, the risk calculations are repeated for each protected component. The final risk is the maximum risk calculated for each protected component.

6.7.2 Leakage Risk

Risk associated with PRD leakage is calculated using Equation (5.120):

$$Risk_l^{prd} = P_l^{prd} \cdot C_l^{prd} \tag{5.120}$$

6.7.3 Total Risk

The total PRD risk is calculated using Equation (5.121).

$$Risk^{prd} = Risk_f^{prd} + Risk_l^{prd} \tag{5.121}$$

6.7.4 Risk Calculation Procedure

The following summarizes the calculation procedure for the failure to open case.

a) STEP 6.1—Calculate the failure to open on demand risk associated with each applicable overpressure demand case scenario, $Risk_{f,j}^{prd}$, using Equation (5.118).

b) STEP 6.2—Calculate the total risk for the failure to open case, $Risk_f^{prd}$, with the sum risk associated with each applicable overpressure demand case scenarios using Equation (5.119).

c) STEP 6.3—Calculate the risk for the PRD leakage case, $Risk_l^{prd}$ using Equation (5.120).

d) STEP 6.4—Calculate the total risk using Equation (5.121).

6.8 Inspection Planning Based on Risk Analysis

6.8.1 RBI Intervals

Risk increases as a function of time as both PRD POF and the probability of PRD leakage increases with time. The recommended PRD interval is calculated based on the date at which the PRD risk reaches the risk target (Part 1, Section 4.4.2).

6.8.2 Effect of PRD Inspection, Testing, and Overhaul on Risk Curve

Figure 6.7 shows the effect of inspection and test, and repair of the PRDs and illustrates the effect of the setting a risk target. The example presented in Figure 6.7 uses a risk target of $25,000/year and resulted in inspection intervals of 5 years. Alternatively, if the risk target were $10,000/year, the resulting inspection interval would have been every 3 years.

Since PRDs are normally overhauled or replaced at the time of testing, the risk of failure goes to zero after inspection and test since the PRD is returned to an as-new condition after overhaul.

6.8.3 Effect of PRD Testing Without Overhaul on Risk Curve

A PRD is us typically overhauled after inspection and test to return the PRD to the as-new condition. Occasionally a PRD is not overhauled after inspection and test. For example, a pop test performed in the shop in the as-received condition may be returned to service without overhaul. Or for example an in situ pop test may be performed without a shop inspection and test. In the case where an overhaul has not been performed, confidence is increased that the PRD was in working condition is gained but the PRD was not restored to an as-new condition. The POF and leakage curves are adjusted (η characteristic life). If the test passed, the test interval will be increased but the risk does not go to zero as if the PRD was overhauled.

6.9 Nomenclature

A^{prd} is the orifice area of the PRD, in^2 (mm^2)

A^{prd}_{total} is the total installed orifice area of a multiple PRD installation, in^2 (mm^2)

C_{env} is the environmental consequence from PRD leakage, $

$C^{prd}_{f,j}$ is the PRD COF to open associated with the j^{th} overpressure demand case, $

C^{mild}_{l} is the consequence of a mild or moderate leak through a PRD, $

C^{prd}_{l} is the PRD consequence of leakage, $

C^{so}_{l} is the consequence of a stuck open PRD, $

C^{mild}_{prod} is the consequence of lost production of mild or moderate leaks, $

C_{sd} is the consequence associated with the repair and maintenance of the PRD, $

CF is the confidence factor placed on the inspection effectiveness

CF_{fail} is the confidence factor that a failed test represents the true condition of the PRD at the time of the test

CF_{i} is the confidence factor placed on the inspection effectiveness associated with the i^{th} historical inspection record

CF_{pass} is the confidence factor that a passed test represents the true condition of the PRD at the time of the test

$Cost_{env}$ is the environmental costs due to a PRD leak, $

$Cost_{flu}$ is the cost of the lost fluid, $/lb ($/kg)

$Cost_{inv}$ is the lost inventory or fluid costs due to a PRD leak, $

$Cost^{mild}_{inv}$ is the cost of lost inventory due to a minor or moderate PRD leak, $

$Cost^{so}_{inv}$ is the cost of lost inventory due to a stuck open PRD, $

$Cost_{prod}$ is the production losses as a result of shutting down to repair a PRD, $

$Cost^{mild}_{prod}$ is the production losses as a result of shutting down to repair a mild or moderate leaking PRD, $

$Cost^{so}_{prod}$ is the production losses as a result of shutting down to repair a stuck open PRD, $

$Cost_{sd}$ is the maintenance and repair costs associated with a PRD, $

D_f is the damage factor as a function of time for equipment components protected by the PRD

D_{mild} is the duration that a minor or moderate PRD leak will go undiscovered, days

D_{sd} is the number of days required to shut a unit down to repair a leaking or stuck open PRD, days

D_{so} is the duration of a stuck open PRD, days

DR_j is the demand rate associated with the j^{th} overpressure demand case, demands/year

DR_{total} is the total demand rate on a PRD, demands/year

$DRRF_j$ is the demand rate reduction factor associated with the j^{th} overpressure demand case

EF_j is the initiating event frequency associated with the j^{th} overpressure demand case, demands/year

F_{MS} is the management systems factor

F_a is the multiple PRD installation adjustment factor

F_c is the adjustment factor for conventional PRVs

F_{env} is the adjustment factor for environmental factors

F_{op} is the adjustment factor for overpressure

$F_{op,j}$ is the adjustment factor for the overpressure for the j^{th} overpressure demand case

F_r is the recovery factor applied to lost inventory

F_s is the adjustment factor for the presence of soft seats

F_{set} is the adjustment factor for the ratio of operating pressure to set pressure

$F(t)$ is the cumulative failure density function or unreliability

gff_n is the GFF for the protected equipment associated with the n^{th} hole size, failures/year

gff_{total} is the total GFF for the protected equipment, years

$lrate_{mild}$ is the leakage rate of a mild or moderate leaking PRD, lb/hr (kg/hr)

$lrate_{so}$ is the leakage rate for a stuck open PRD, lb/hr (kg/hr)

$MAWP$ is the maximum allowable working pressure of the protected equipment, psig (kPa)

$MTTF$ is the mean time to failure

ndc is the number of demand cases

$P_{f,j}$ is the POF (loss of containment) of the protected equipment associated with the j^{th} overpressure demand case, failures/year

$P_f\,(t)$ is the POF (loss of containment) of the protected equipment, failures/year

$P_{f,cond}^{prd}$ is the conditional POFOD, failures/demand

$P_{f,j}^{n}$ is the POF (loss of containment) of the protected equipment for the n^{th} hole size associated with the j^{th} overpressure demand case, failures/year

$P_{f,j}^{prd}$ is the POF of a PRD associated with the j^{th} overpressure demand case, failures/year

$P_{f,prior}^{prd}$ is the prior POFOD, failures/demand

$P_{f,wgt}^{prd}$ is the weighted POFOD, failures/demand

P_{fod} is the PRD POFOD, failures/demand

$P_{fod,j}$ is the PRD POFOD associated with the j^{th} overpressure demand case, failures/demand

P_l^{prd} is the PRD POF, failures/year

$P_{l,wgt}^{prd}$ is the weighted POL, failures/demand

P_o is the overpressure likely to occur as a result of a PRD failing to open upon demand, psig (kPa)

$P_{o,j}$ is the overpressure likely to occur as a result of a PRD failing to open upon demand, associated with the j^{th} overpressure demand case, psig (kPa)

$P_{p,prior}^{prd}$ is the prior probability of passing on demand, failures/demand

P_s is the storage or operating pressure of the protected equipment, psig (kPa)

P_{set} is the set pressure of the PRD, psig (kPa)

$Risk^{prd}$ is the total risk for a PRD, \$/year

$Risk_f^{prd}$ is the risk of a PRD failure to open, \$/year

$Risk_{f,j}^{prd}$ is the risk of a PRD failure to open associated with the j^{th} overpressure demand case, \$/year

$Risk_l^{prd}$ is the risk of PRD leakage, \$/year

$R(t)$ is the risk as a function of time, ft^2/year (m^2/year) or \$/year

t is time, years

$t_{dur,i}$ is the actual duration between inspections associated with the i^{th} historical inspection record, years

t_{insp} is the service duration, years

$Unit_{prod}$ is the daily production margin on the unit, \$/day

W_c^{prd} is the rated capacity of a PRD, lb/hr (kg/hr)

β is the Weibull shape parameter

η is the Weibull characteristic life parameter, years

η_{def} is the Weibull characteristic life parameter based on the default service severity chosen for a specific PRD, years

η_{mod} is the Weibull characteristic life parameter modified to account for installation factors, design features, overpressure and environmental factors, years

η_{upd} is the Weibull characteristic life parameter updated to account for inspection history, years

6.10 Tables

Table 6.1—Basic Data Needed for the PRD Module

Data	Description	Data Source
PRD type	Type of PRD — 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.	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 POFOD and POL curves. FAIL TO OPEN — Mild — Moderate (default) — Severe LEAKAGE — 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 5.2. 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 (see Section 5.1.6) — Inlet and outlet piping condition [see Section 5.2.4 i),1)]	User specified
Protected equipment details	Operating conditions, design conditions, dimensions, damage mechanisms, GFF, and DFs	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.	Fixed equipment
Injury costs	Cost of serious injury, $	Fixed equipment
Environmental costs	Environmental fines and costs associated with PRD leakage or loss of equipment containment, $/event	Fixed equipment
Production costs	Cost of lost production, $	Fixed equipment
Unit costs	Cost to replace unit, $/ft^2	Fixed equipment

Table 6.2—Overpressure Scenario Logic

Initiating Event Frequency	Equipment Type	PRD DRRF	Qualifier	Overpressure Potential	Background and Comments
Overpressure Scenario—Fire					
1 per 250 years See Lees [13] page A7-7, states major fire at plant 1 every 10 years	All equipment types	0.1 Industry experience justifies this value	N/A	Calculated burst pressure or estimated as design margin × MAWP	— Modified by industry data that indicate 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
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	Calculated burst pressure or estimated as design margin × MAWP	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	
Overpressure Scenario—Electrical Power Failure					
0.08 per year (1 per 12.5 years) power supply failure per table on page 9/30 of [13]	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	Calculated burst pressure or estimated as design margin × MAWP	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	

Initiating Event Frequency	Equipment Type	PRD DRRF	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 [13] 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	Calculated burst pressure or estimated as design margin × MAWP	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 × 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	Calculated burst pressure or estimated as design margin × MAWP	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		Calculated burst pressure or estimated as design margin × MAWP	Added increase in potential overpressure with fired/radiant heat transfer. Assumption is made that rupture occurs.
Overpressure Scenario—Control Valve Fail Close at Outlet					

Initiating Event Frequency	Equipment Type	PRD DRRF	Qualifier	Overpressure Potential	Background and Comments
1 per 10 years [14] 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	Calculated burst pressure or estimated as design margin × MAWP	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 × 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	Calculated burst pressure or estimated as design margin × MAWP	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		Calculated burst pressure or estimated as design margin × MAWP	Added increase in potential overpressure with fired/radiant heat transfer. Assumption is made that rupture occurs.

Overpressure Scenario—Control Valve Fail Open at Inlet, Including the HP/LP Gas Breakthrough Case

Initiating Event Frequency	Equipment Type	PRD DRRF	Qualifier	Overpressure Potential	Background and Comments
1 per 10 years [14] 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—Runaway Chemical Reaction					
1 per year	All equipment	1.0		Calculated burst pressure or estimated as design margin × MAWP	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 [15])	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
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 × 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	

Initiating Event Frequency	Equipment Type	PRD DRRF	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 $^{1}/4$ in. 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 POF is 1.0 (flange leaks), modeled as a $^{1}/4$ in. 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—Liquid Overfilling					

Initiating Event Frequency	Equipment Type	PRD DRRF	Qualifier	Overpressure Potential	Background and Comments
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	Calculated burst pressure or estimated as design margin × MAWP	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 × MAWP of upstream pressure source vessel	

Table 6.3—Default Initiating Event Frequencies

Overpressure Demand Case	Event Frequency	EF_j (events/year)	$DRRF_j$ (See Notes 2 and 3)	Reference
1. Fire	1 per 250 years	0.0040	0.10	[12]
2. Loss of cooling water utility	1 per 10 years	0.10	1.0	[12]
3. Electrical power supply failure	1 per 12.5 years	0.080	1.0	[12]
4a. Blocked discharge with administrative controls in place (see Note 1)	1 per 100 years	0.010	1.0	[13]
4b. Blocked discharge without administrative controls (see Note 1)	1 per 10 years	0.10	1.0	[13]
5. Control valve failure, initiating event is same direction as CV normal fail position (i.e. fail safe)	1 per 10 years	0.10	1.0	[14]
6. Control valve failure, initiating event is opposite direction as CV normal fail position (i.e. fail opposite)	1 per 50 years	0.020	1.0	[14]
7. Runaway chemical reaction	1 per year	1.0	1.0	
8. Heat exchanger tube rupture	1 per 1000 years	0.0010	1.0	[15]
9. Tower P/A or reflux pump failures	1 per 5 years	0.2	1.0	
10a. Thermal relief with administrative controls in place (see Note 1)	1 per 100 years	0.010	1.0	Assumed same as blocked discharge
10b. Thermal relief without administrative controls (see Note 1)	1 per 10 years	0.10	1.0	Assumed same as blocked discharge
11a. Liquid overfilling with administrative controls in place (see Note 1)	1 per 100 years	0.010	0.10	[12]
11b. Liquid overfilling without administrative controls (see Note 1)	1 per 10 years	0. 10	0.10	[12]

NOTE 1 Administrative controls for isolation valves are procedures intended to ensure that personnel actions do not compromise the overpressure protection of the equipment.

NOTE 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 firefighting efforts are usually taken to minimize overpressure.

NOTE 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.

NOTE 4 Where the Item Number has a subpart (such as “a” or “b”), this clarifies that the overpressure demand case will be on same subpart of Table 5.3.

Table 6.4—Categories of PRD Service Severity (Fail Case Only)

PRD Service Severity	Characteristic MTTF	Characteristic of Failure	Expected Stream Characterization	Typical Temperature	Examples of Service
Mild	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	— Clean hydrocarbon products at moderate temperature — No aqueous phase present — Low in sulfur and chlorides	Low temperature, always << 500 °F	Examples include: product hydrocarbon streams (including lubricating oils), liquefied petroleum gas (LPG, BFW, low-pressure steam, and clean gasses such as nitrogen and air
Moderate	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 over a long period of time	— Hydrocarbons that may contain some particulate matter — A separate aqueous phase may be present, but is a minor component — Clean, filtered, and treated water may be included in this category — Some sulfur or chlorides may be present	Up to 500 °F (may exist)	Examples include: intermediate hydrocarbon streams, in-service lube and seal oils, process water (NOT cooling water or boiler feed water), and medium- to high-pressure steam
Severe	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)	— High-temperature hydrocarbon streams 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	> 500 °F	Examples include: heavy hydrocarbon streams such as crude, amine services, cooling water, corrosive liquids and vapors, and streams containing H_2S

NOTE 1 MTTF does not reflect replacement history, where the history indicates a renewal of the asset without a failure noted.

NOTE 2 Refer to Table 5.13 for the categories for the LEAK case.

Table 6.5—Default Weibull Parameters for POFOD

Fluid Severity	Conventional and Balanced Bellows PRVs [1]		Pilot-operated PRVs [2]		Rupture Disks [3]	
	β	η_{def}	β	η_{def}	β	η_{def}
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

NOTE 1 The η_{def} parameter values for conventional PRVs are reduced by 25 % if the discharge is to a closed system or to flare; see Section 5.2.4 g).

NOTE 2 The η_{def} parameter values for pilot-operated PRVs 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; see Section 5.2.4 e).

NOTE 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; see Section 5.2.4 f).

Table 6.6—Environmental Adjustment Factors to Weibull η Parameter

Environment Modifier	Adjustment to POFOD η Parameter	Adjustment to POL η Parameter
Operating temperature 200 °F < 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

NOTE 1 Some pilot-operated PRVs operate extremely well with operating ratios approaching 98 %. In these cases, the environmental factor should not be applied (reference API 520, Part 1).

NOTE 2 This factor should not be applied if the environmental factor for operating ratio is already applied.

Table 6.7—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}	0.4	0.5	0.70	0.9
Fail, CF_{fail}	0.4	0.70	0.95	0.95
No leak, CF_{noleak}	0.4	0.5	0.70	0.9
Leak, CF_{leak}	0.4	0.70	0.95	0.95

Table 6.8—Set Pressure Factor

PRV Type	Set Pressure Factor
Pilot-operated PRVs	$F_{set} = 1 - \left[\dfrac{0.95 - min\left[0.95, \dfrac{P_S}{P_{set}}\right]}{0.95}\right]$
Rupture disks	$F_{set} = 1$
Conventional PRVs and balanced bellows PRVs	$F_{set} = 1 - \left[\dfrac{0.90 - min\left[0.90, \dfrac{P_S}{P_{set}}\right]}{0.90}\right]$
NOTE 1 P_S denotes the operating pressure and P_{set} denotes the set pressure.	

Table 6.9—Inspection Updating Equations

Inspection Effectiveness and Result	Equation for Weighted POFOD
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	
Fairly effective pass	
Highly effective fail	$P_{f,wgt}^{prd} = P_{f,cond}^{prd}$
Usually effective fail	$P_{f,wgt}^{prd} = 0.5 \cdot P_{f,prior}^{prd} + 0.5 \cdot P_{f,cond}^{prd}$
Fairly effective fail	
Ineffective/No Inspection	$P_{l,cond}^{prd} = CF_l \cdot P_{l,prior}^{prd} + (1 - CF_{nl}) \cdot P_{nl,prior}^{prd}$

Table 6.10—Design Margins for Various Codes of Construction

Construction Code	Design Margin
ASME Section VIII, Div. 1, pre-1950	5.0
ASME Section VIII, Div. 1, 1950–1998	4.0
ASME Section VIII, Div. 1, 1999 and later	3.5
ASME Section VIII, Div. 2, pre-2007	3.0
ASME Section VIII, Div. 2, 2007 and later	2.4
ASME B31.3	3.0
AS 1210	3.5

NOTE For any construction code not listed in this table or when design by analysis was utilized to design the equipment, it is the responsibility of the owner–~~user~~operator to determine the design margin.

Table 6.11—Constants for Design Margin

Design Margin	Constant a	Constant b
5	2.28E-06	2.598628
4	9.57E-07	3.464837
3.5	4.79E-07	4.157804
3	1.69E-07	5.197255
2.4	1.82E-08	7.42465

NOTE 1: Constants a and b are used in Equation 5.99

NOTE 2: A g_{ff} of 3.06E-05 is used to calculate constant a.

Table 6.12—DF Classes for Protected Equipment

DF Class	DF	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. Damage mechanisms have been identified and inspection data are available.
Minor	200	One or more damage mechanisms have been identified, limited inspection data available, and fairly minor evidence of damage. Single damage mechanism identified, recent inspection indicates minor evidence of damage.
Moderate	750	Moderate damage found during recent inspection. Low susceptibility 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 6.13—Categories of PRD Service Severity (LEAK Case Only)

PRD Service Severity	Typical Temperature	Expected Stream Characterization	Examples of Service
Mild	Low temperature, always << 500 °F	Many heavy liquid streams such as crude oil tend not to leak through a PRD and are considered mild service severity	— Cooling water and amine services are examples of corrosive/fouling fluids that do not leak — Clean fluids such as LPG, air, and nitrogen are MILD leakage services
Moderate	Up to 500 °F (may exist)	Most of the intermediate and product hydrocarbon streams and most hydrocarbon vapors	— Lube, seal and cycle oils, and process water (**NOT** cooling water, condensate, or BFW)
Severe	>500 °F	High-temperature services	BFW/condensate, steam, and corrosive liquids such as caustic and acids
NOTE Refer to Table 5.4 for the categories for the FAIL case.			

Table 6.14—Default Weibull Parameters for POL

Fluid Severity	Conventional PRVs [1]		Balanced Bellows PRVs [1]		Pilot-operated PRVs [2]		Rupture Disks [3]	
	β	η_{def}	β	η_{def}	β	η_{def}	β	η_{def}
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

NOTE 1 The η_{def} parameter values are increased by 25 % for conventional and balanced PRVs that have soft seats.

NOTE 2 The η_{def} parameter values for pilot-operated PRVs are currently based on the conventional PRV data, since there are currently no failure rate data to support otherwise.

NOTE 3 Without any failure rate data for rupture disks, the conventional PRV values for Mild service were used.

Table 6.15—Potential Consequences of Pressure Vessel Overpressure

Accumulation (% over MAWP)	Significance [11]	Potential Consequence
10 %	ASME code allowable accumulation for process upset cases (non-fire) protected by a single PRD	No expected consequence at this accumulation level
16 %	ASME code allowable accumulation for process upset cases protected by multiple PRDs	No expected consequence at this accumulation level
21 %	ASME code allowable accumulation for external fire relief cases regardless of the number of PRDs	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 6.16—Estimated Leakage Duration from PRDs

PRD Inlet Size (in.)	**Leak Duration Discharge to Flare or Closed System, D_{mild}** (days)	**Leak Duration Discharge to Atmosphere, D_{mild}** (days)
≤ 3/4 in.	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 6.17—Estimated Leakage Rate from PRVs

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	Severe	2.4	10	25

6.11 Figures

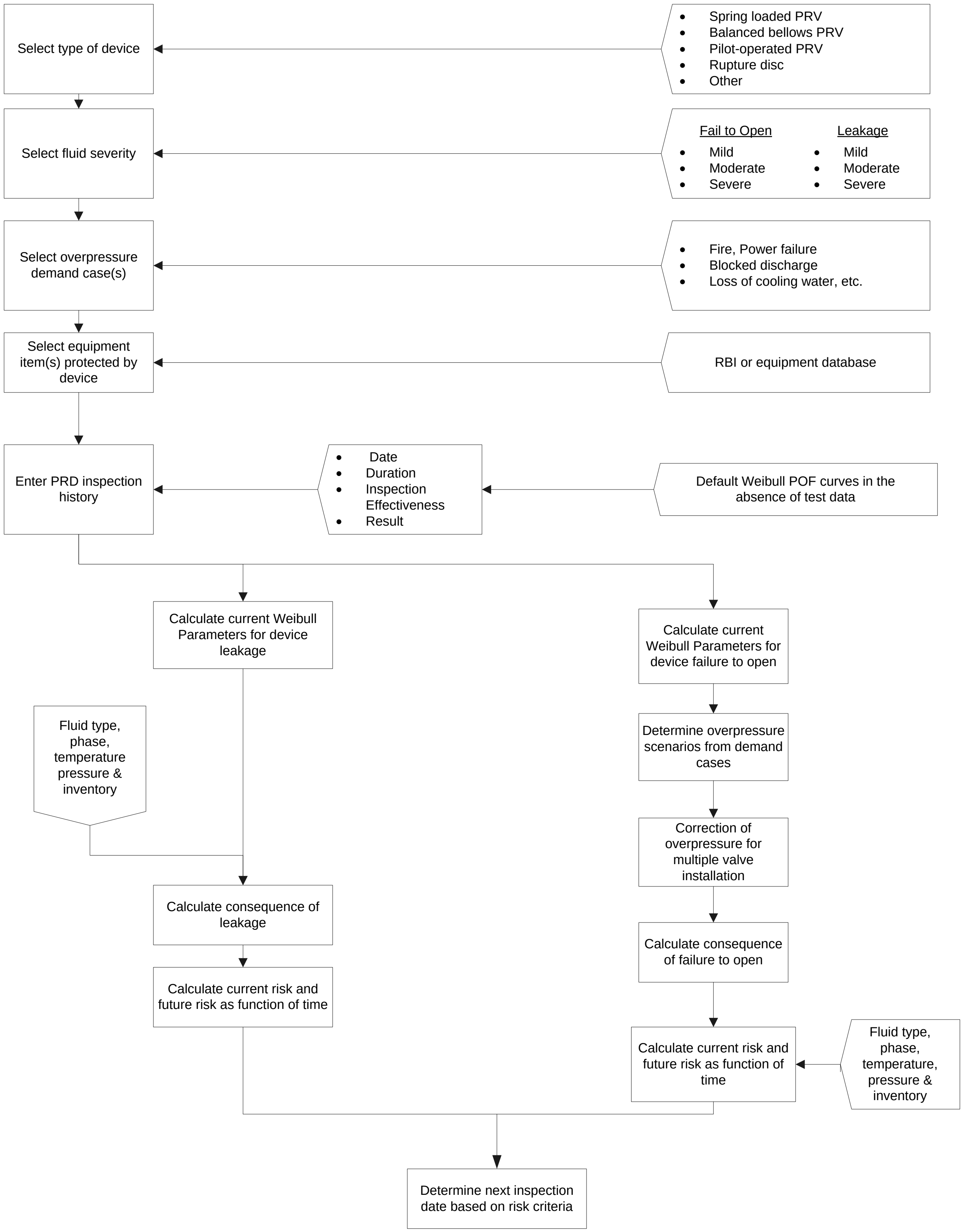


Figure 6.1—PRD RBI Methodology

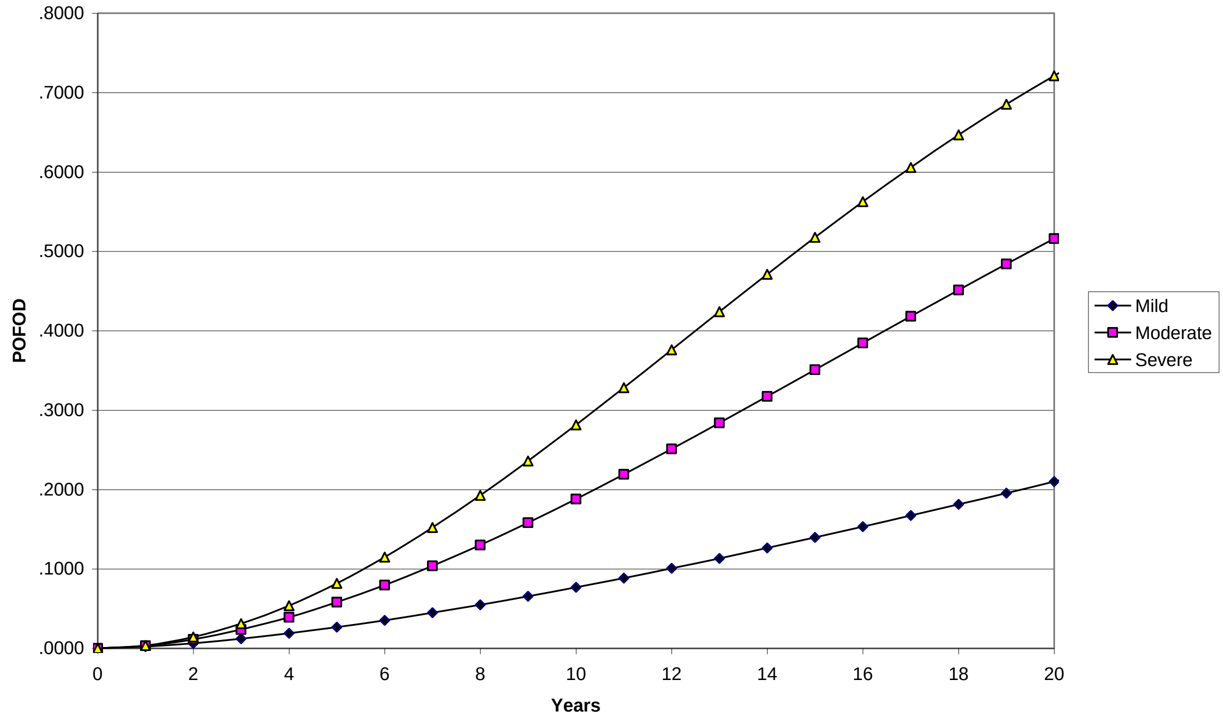


Figure 6.2—Default Conventional PRV Fail to Open on Demand Weibull Curves

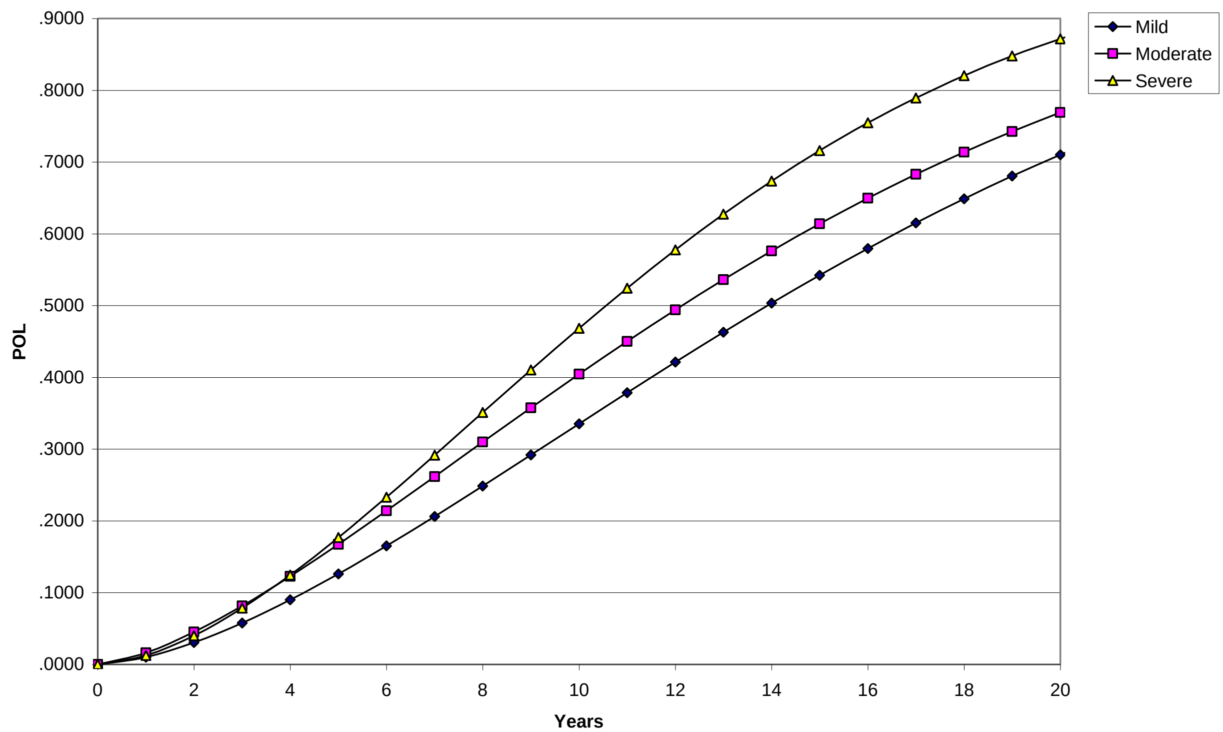


Figure 6.3—Default Leakage Failure Rate for Conventional PRVs

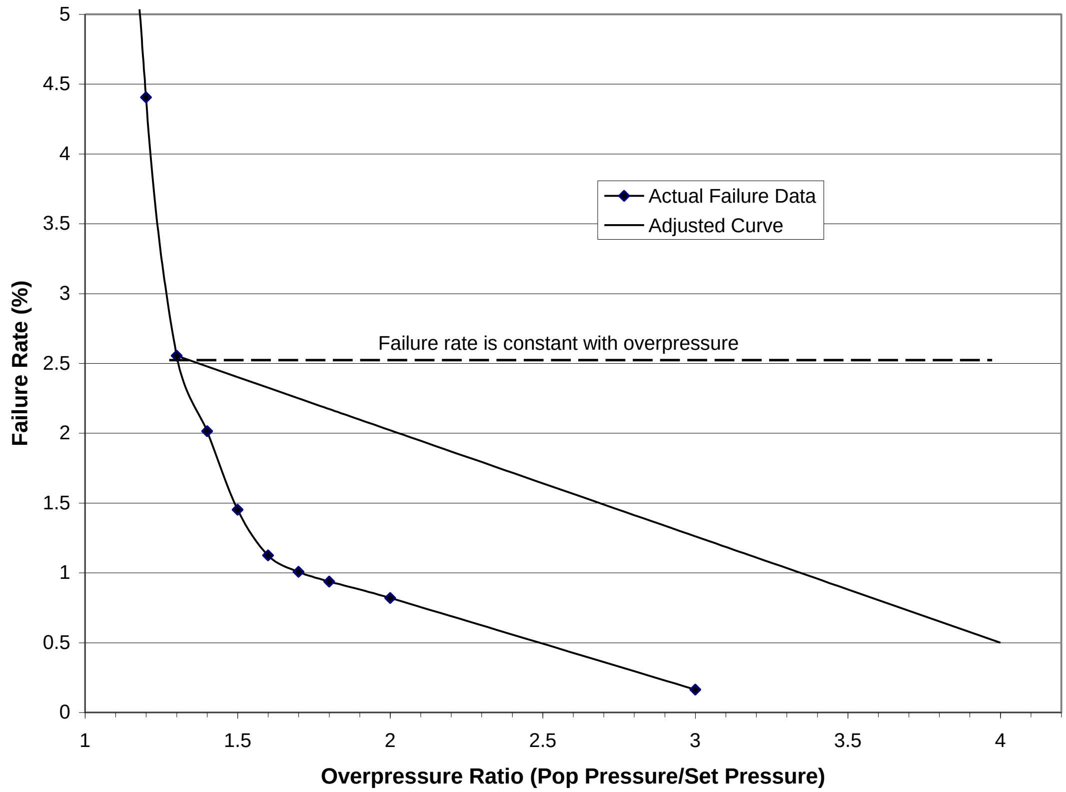


Figure 6.4—PRD Failure Rate As a Function of Overpressure

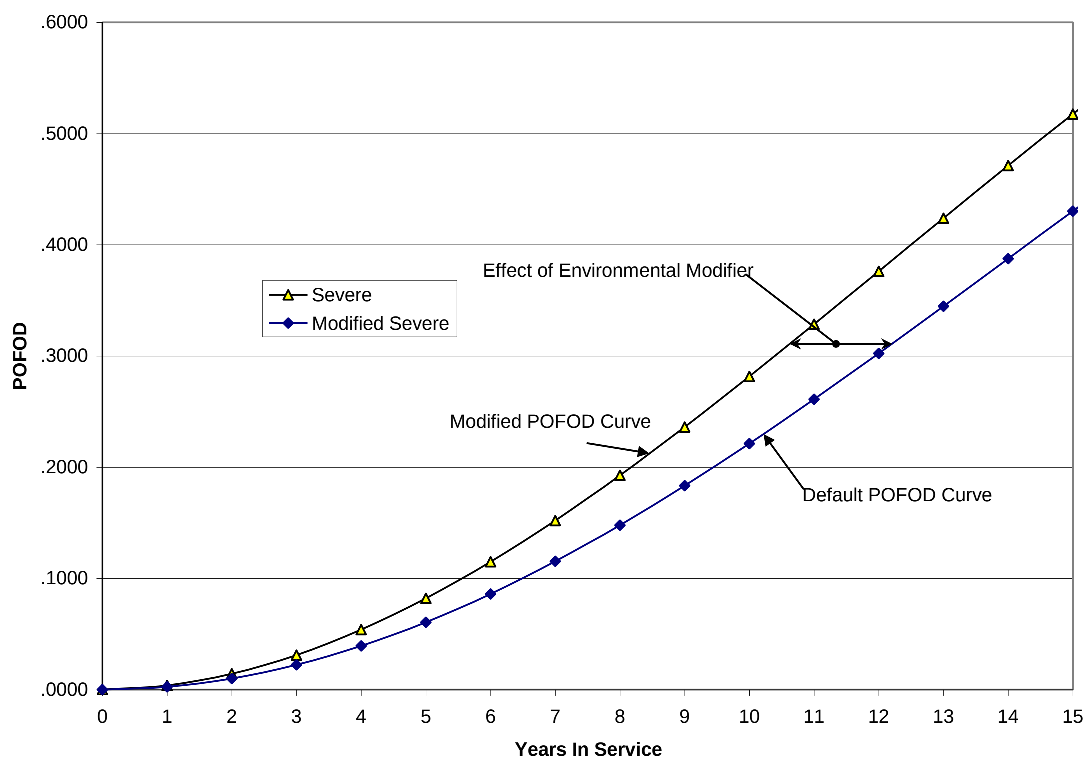
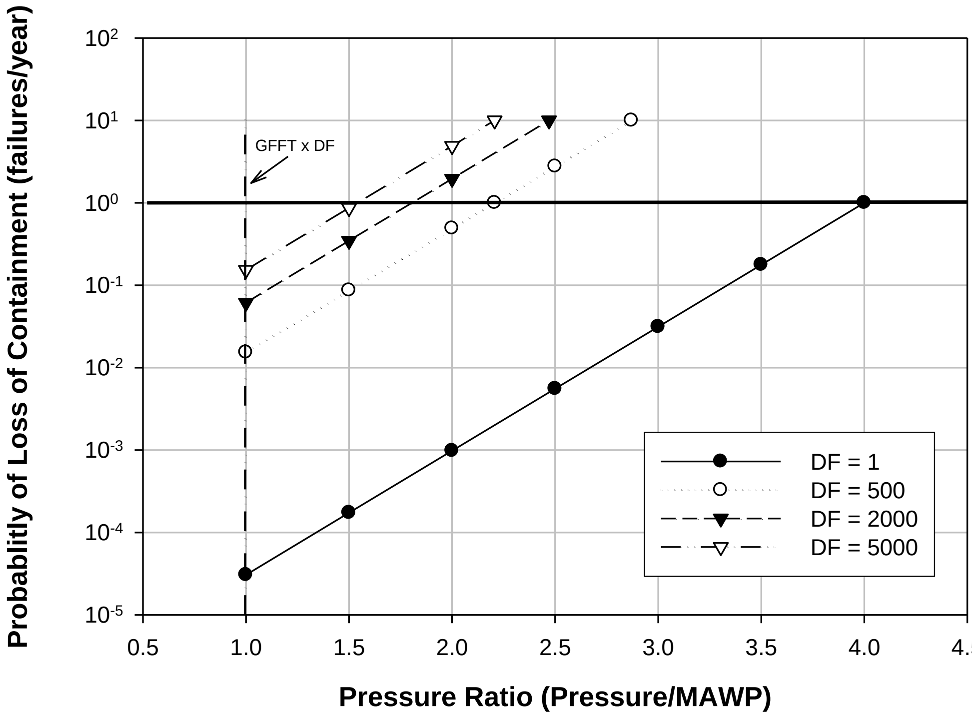


Figure 6.5—Effect of Environmental Factors on PRD Weibull Curves



For an example pressure vessel with:

1. gff_{total} = 3.06 × 10^{-5}
2. Design margin = 4
3. Estimated burst pressure of 4 × MAWP

Figure 6.6—Probability of Loss of Containment As a Function of Overpressure

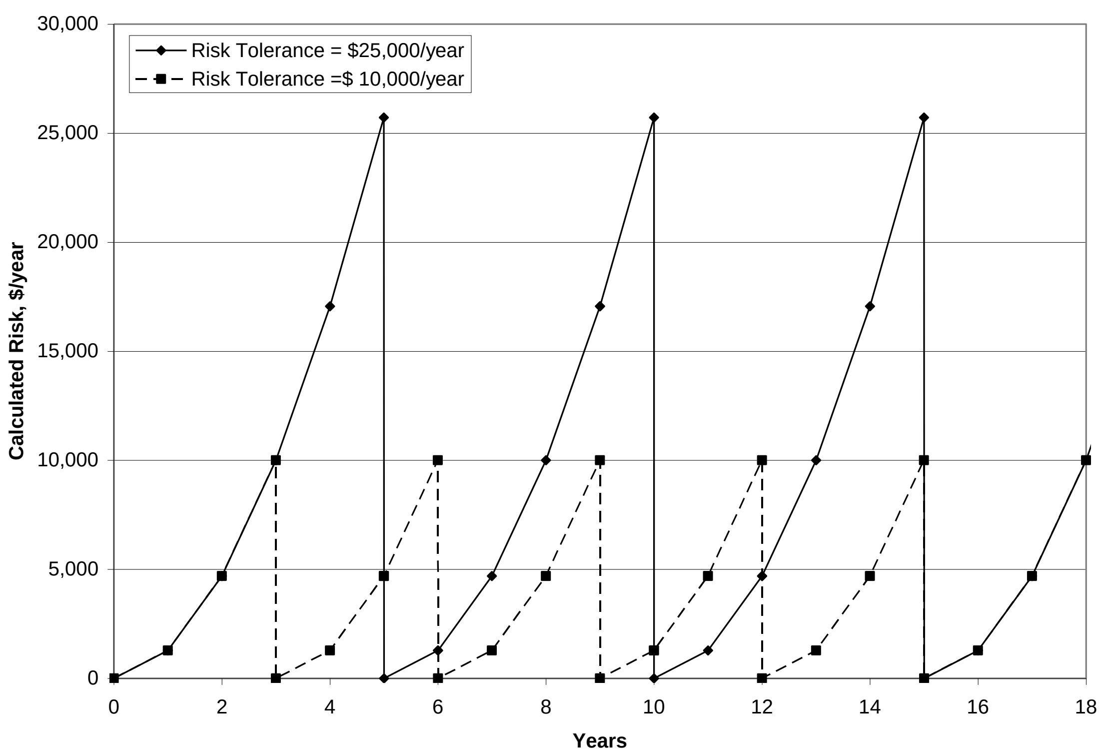


Figure 6.7—Inspection Test Updating of PRDs

7 Steam System

7.1 Overview

7.1.1 General Background

Steam systems account for approximately 30% of the total energy used in a typical petroleum refinery [17] based on US Department of Energy figures,. Steam systems are utilized throughout the plant for motive, heating and process purposes, such as in the steam turbine driver for the recycle gas compressor, the re-boiler for the depropanizer column, and for stripping steam for crude distillation.

Steam system specialists work with plants to identify opportunities to reduce the amount of energy consumed by steam systems to stay competitive. Steam system maintenance costs should also be optimized and to protect health and safety issues as well as avoid unplanned downtime. The integrity and efficiency of steam-using equipment is critical to the operation and productivity of petrochemical industry. In addition, steam distribution systems and steam tracing systems which provide the heat necessary to maintain flow rates in product distribution lines, vessels and reactors [18].

Routine inspection and testing of steam-using systems consisting of steam traps, associated lines, and equipment is required to avoid failures of the traps, associated lines or equipment , leading to failure of the system. Such failures have resulted in a significant loss of steam and have led to personal injury.

A risk-based approach to evaluate the criticality of equipment in steam-using systems is covered here to set inspection and testing interval or possible mitigation actions. The scope of this section includes steam traps, associated steam distribution lines, and equipment using steam. The methodology involves the use of reliability data for steam trap types in the form of Weibull parameters.

It is assumed that devices have been designed in accordance with specific design standards and sized, selected, and installed appropriately. It is also assumed that the devices are included in inspection plans. The fundamental approach is to determine the POF from plant-specific data if available, or to be determined from industry default data. These inputs are used to generate a POF as a function of time via a Weibull statistical approach. The consequence of device failure is calculated using methods outlined in Part 3, modified to include different failure scenarios. The combination of consequence and time-based POF provides a risk value which increases with time between inspections/tests. This allows inspection and test intervals to be determined based on risk targets. Figure 7.1 illustrates the basic methodology required for the determination of POF and is the basis for setting up inspection and test intervals or any mitigation actions.

7.1.2 Steam Application Types

Steam is essential for heating, mechanical drives and several other applications in process plants and steam traps are commonly used to ensure that steam is not wasted. ~~In process plants, steam is essential for heating, mechanical drives and several other applications. In each case, steam traps are commonly used to ensure that steam is not wasted.~~A steam trap is a type of automatic valve which filters out condensate (for example condensed steam) and non-condensable gases such as air without letting steam escape. As described in ANSI/FCI 69-1-1989, a steam trap is a self-contained valve which automatically drains the condensate from a steam-containing enclosure while remaining tight to live steam, or if necessary, allows steam to flow at a controlled or adjusted rate [19]. Most steam traps will also pass non-condensable gases while remaining tight to live steam. Various types of steam trap mechanisms (operating principles) have been developed to automatically discharge condensate and non-condensable gases. The most widely used mechanisms are those reliant on differences in temperature, specific gravities, and pressure. Each of these types of steam traps has its own advantages and applications.

Steam traps are usually required to drain condensate from steam piping, steam-using process and comfort heating equipment, tracer lines, and drive-power equipment such as turbines. Each of these applications may require the steam trap to perform a slightly different role.

In general, there are five major steam application groups that use steam traps: steam distribution lines; steam-heated equipment; steam-driven equipment; steam tracing; and direct steam applications. These systems can be indispensable in delivering the energy needed for operating an industrial plant, including process heating (e.g., heat exchangers) and steam tracing systems, as well as mechanical drives (e.g., steam turbines).

Examples of equipment used in steam systems, illustrating the importance of their application to the refining process, are listed in Table 7.1.

7.2 The Definition of Steam System

7.2.1 Overview

The role of the steam system is to reliably supply steam of the highest quality to the steam-using equipment. In order for this to be achieved, condensate is quickly and efficiently removed through steam traps to the correct condensate discharge location (CDL). Therefore, steam systems are an integral part of the process plant. A steam system consists of a combination of a steam-using equipment and its associated lines with steam traps. Figure 7.2 shows multiple steam using systems with the following components:

a) Steam traps

b) Associated steam lines (distributing and condensate)

c) Equipment (steam-using equipment)

Depending on the system design, mechanical pumps or control valves may be installed in place of steam traps (as shown in Figure 7.2~~6.1~~).

COF is a key driver for a Risk-Based Inspection (RBI) approach in steam-using/distribution systems, for assessment of steam traps, associated steam lines, and steam-using equipment (as described in Section 7.~~6.~~2).

7.2.2 Steam Trap

Steam traps are a type of automatic valve which filters out condensate (i.e., condensed steam) and non-condensable gases, such as air, without letting steam escape. In industry, steam is used regularly for heating or as a driving force for mechanical power. Steam traps are used in such applications to ensure steam is not wasted. Based on the operating principles of steam traps, they can be classified as mechanical, thermostatic or thermodynamic. Table 7.2 describes different types of steam traps for each of the above categories.

7.2.3 Steam Lines

Steam lines supply steam to the steam-using equipment. As described, condensate is removed through steam traps installed at CDLs. The steam flow rates are typically higher in steam distribution lines than in other equipment, reaching velocities of > 100 ft/s (30 m/s). At these speeds, when the cross-sectional area of a line section is liquid full, slugs of condensate can be carried through the piping at high velocities, causing water hammer. Potentially, this may cause failures of piping, valves, and equipment as well as personal injuries. ~~At these speeds, when the cross-sectional area of a pipe section is completely filled by water, slugs of condensate can be carried through the piping at high velocity causing water hammer, which may cause failures through damage to piping, valves and equipment and may result in personal injuries.~~ The higher velocities in steam lines should be considered during design when the location of trap installations is being decided. ~~The higher flow velocities in steam lines shall therefore be taken into account during decisions regarding location and design of trap installations.~~

7.2.4 Steam-Using Equipment

As described in Section 7.1.2, there are many applications for steam and, depending on the application, various types of steam-using equipment are used. Table 7.1 provides examples of five steam application groups.

7.2.5 Steam System Damage Mechanism Equipment and Failure Modes

~~7.2.4.1~~7.2.5.1 Background

The role of steam distribution lines is to reliably supply high quality steam to steam-using equipment. Condensate Is quickly and efficiently removed through steam traps installed in proper CDL installations. CDLs are susceptible to failures due to blockage (cold) or leakage (described in Section 7.2.5.3.1 and Section 7.2.5.3.2). This methodology currently does not cover freeze protection of CDLs.

~~A sudden release of steam or scalding water can occur due to failure modes such as water hammer. Water hammer has been cited by Paffel [204] as the primary problem in steam systems. Water hammer is a known vulnerability in steam systems and is sometimes referred to as Condensate Induced Water Hammer. This most commonlyWater Hammer occurs when steam is introduced into cold pipework which has not been drained sufficiently. As the steam cools, it turns into condensate, taking up a smaller volume in the pipework than steam. This produces a vacuum or pocket into which the water flows rapidly, creating an impact against the pipework.~~

The failures described in this section will also result in equipment failure consequences such as industrial steam turbine erosion failures, flooding of heat exchangers, failures in steam tracing systems, failures in flare systems (loss of steam will prevent atomizing of gases prior to burning), distillation towers and strippers.

7.2.5.2 Damage Mechanism

7.2.5.2.1 Water hammer

A sudden release of steam or scalding water can occur due to failure modes such as water hammer. Water hammer has been cited by Paffel [20] as the primary problem in steam systems and is sometimes referred to as Condensate Induced Water Hammer. Water Hammer occurs when steam is introduced into cold pipework which has not been drained sufficiently. As the steam cools, it turns into condensate, taking up a smaller volume in the pipework than steam. This produces a vacuum or pocket into which the water flows rapidly, creating an impact against the pipework.

Water hammer generated in steam and condensate recovery systems is ordinarily classified via two main causes:

a) High-speed condensate slamming into, for example piping

b) Sudden condensation of steam, which produces walls of condensate that crash into each other.

When water hammer occurs, a momentary abrupt pressure change of over 1450 psi (10 MPa) may occur inside the piping. The change in pressure may result in an impact and can cause pipe rupture, severely jarring piping, equipment or machinery housings, possibly resulting in damage to gaskets and valve flanges or the valves themselves. Water hammer in steam distribution piping interrupts service and can cause failures leading to personal injury and property damage. According to historical failures, 82% of steam systems experience some type of water hammer. In a typical steam-using system, water hammer causes 67% of premature steam system component failures [17].

Water hammer events are commonly caused by the following systemic failures:

a) Failure to ensure water (condensate) has been removed using steam traps and drains prior to admitting steam into the piping system.

b) Failure to correctly maintain steam traps, drain, and to blowdown valves (in order to preserve operable condition).

c) Failure to ensure an adequate number of steam traps and drains have been installed at locations conducive to condensate removal.

d) Failure to operate system valves correctly as well as failure to use bypass valves to safely warm system piping downstream of isolation valves.

7.2.5.3 Failure Modes

~~7.2.4.1.1~~7.2.5.3.1 Steam Trap Blockage Leading to Water Hammer

~~When a steam trap is blocked, the~~ Condensate cannot be discharged when the steam trap is blocked, often resulting in water hammer contributing to potential equipment damage.

~~7.2.4.1.2~~7.2.5.3.2 Steam Trap Leakage

Leakage is another mode of steam trap failure resulting in energy waste and poor environmental compliance. The failure consequence of leakage is described in Section 7.4.2

7.3 Probability of Failure Methodology

7.3.1 Use of Weibull Curves

The POF for steam systems is calculated using a two parameter Weibull distribution as expressed in Equation (5.122) as shown in Part 1, Section 4.1.3. Use of Weibull curves for establishing POF is further described in Part 1, Section 4.1.3.

$$P_f = 1 - exp\left(-\left(\frac{t}{\eta}\right)^{\beta}\right) \tag{5.122}$$

Where β is the Weibull shape parameter, η is the Weibull characteristic life parameter, in years, and t is the independent variable time in years.

The POF of the specific trap is related to identifiable process and installation conditions. Such conditions may be related to design, operational and maintenance/inspection history conditions. Also associated with failure are conditions such as poor manufacturing and installation and excessive piping vibration. Improper installations or poor operational and maintenance condition may also increase the POF.

~~7.3.1~~7.3.2 Required data

The basic data required for the evaluation of POF for steam systems are listed in Table 7.3.

~~7.3.2~~7.3.3 Overview

This section presents a procedure to calculate the POF for a steam system. Figure 7.2 provides an overview of the POF calculation framework for steam using systems. POF is a function of time for a range of steam trap types and properties, using Weibull fitting of steam trap failure data. The POF of the associated lines is then derived and combined with the steam-using equipment generic failure frequencies to calculate a system POF. Final POF values are obtained by tailoring the POF for steam traps and equipment to local conditions by customized probability factors.

As described in Section 7.2, a steam system consists of a combination of steam-using equipment and its associated lines with steam traps ~~equipment and its associated lines~~. The POF of each system will be considered as the combined effect of individual equipment with its associated traps for both leakage and blockage, i.e.:

$$P(t)_{f,final,leak\,(steam\ using\ system)} = P(t)_{f\,(equ)} \cdot P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} \tag{5.123}$$

$$P(t)_{f,final,cold\,(steam\ using\ system)} = P(t)_{f\,(equ)} \cdot P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} \tag{5.124}$$

$$P(t)_{f,final}(steam\ using\ systems) = P(t)_{f(equ)} \cdot P(t)_{f,final_{(ST,MP\ or\ CV)}}$$

~~*Where :*~~ ~~(5.123)~~

$P(t)_{f,final_{(ST,MP\ or\ CV)}}$ *is the combined POF calculated for muliple steam traps, mechanical pumps and control valves in the associated lines.*

The procedure for calculation of $P(t)_{f,final,leak_{(ST,MP\ or\ CV)}}$ and $P(t)_{f,final,cold_{(ST,MP\ or\ CV)}}$ is provided in Section 7.3.4 and Section 7.3.5 . $P(t)_{f(equ)}$ is the POF calculated for the steam using equipment as explained in Section 7.3.6.

~~7.3.3~~7.3.4 Probability of Failure (Steam Line)

~~7.3.3.1~~7.3.4.1 POF for Steam Traps, Mechanical Pumps and Control Valves

Analysis has been carried out on the historical time to failure data (for various failure types) and a Weibull distribution has been fitted. As described in Section 7.3.1, Weibull functions are suitable for such analysis with the added advantage of having the ability to evaluate large populations of data to seek trends. In the absence of large sets of failure data, the functions are still useful as a starting point.

Equation (5.122) is the cumulative failure density function of a two parameter Weibull distribution, also referred to as the Probability of Failure (POF) for a steam trap. In this equation, t is the in-service life of the steam trap (in years), η is the characteristic life (also in years) and β is the shape parameter.

Once the scale $\boldsymbol{\eta_{def,ST}}$ (for leak and blockage) and shape β_{ST} parameters are obtained from Table 7.4. (from historical data analysis), the POF of the steam trap is calculated using Equation (5.125) for leakage and (5.126) for blockage.

$$P(t)_{f,def,leak} = 1 - exp\left[-\left(\frac{t}{\eta_{def,leak,ST}}\right)^{\beta_{ST}}\right] \tag{5.125}$$

$$P(t)_{f,def,cold} = 1 - exp\left[-\left(\frac{t}{\eta_{def,cold,ST}}\right)^{\beta_{ST}}\right] \tag{5.126}$$

The data presented in Table 7.4 are based on the best available sources and experience to date from owner-~~user~~operators. Table 7.4 introduces default Weibull parameters for the different steam trap types in both failure modes. However, it is recommended that both Weibull parameters be used by the owner-~~user~~operator where more accurate data for default shape/scale parameters are available. The default parameters in Table 7.4 are suggested for use when data is unavailable.

~~7.3.3.2~~7.3.4.2 Adjusted POF for Steam Traps, Mechanical Pumps and Control Valves

Adjustments are made to the η parameter to increase or decrease POF as a result of condition of design/installation, operation or maintenance history factors. POF is adjusted based on the adjustment multiplier for each design/installation, F_D, operational, F_O, or maintenance history, F_M, conditions. The default POF ($P(t)_{f,def,leak}$ and $P(t)_{f,def,cold}$), needs to be adjusted by the adjustment multipliers given in Table 7.5 to Table 7.13.

$$\eta_{adj,leak_{(ST,MP\ or\ CV)}} = \eta_{def,leak,ST} \cdot F_{D_{(ST,MP\ or\ CV)}} \cdot F_{O_{(ST,MP\ or\ CV)}} \cdot F_{M_{(ST,MP\ or\ CV)}} \quad (5.127)$$

$$\eta_{adj,cold_{(ST,MP\ or\ CV)}} = \eta_{def,cold,ST} \cdot F_{D_{(ST,MP\ or\ CV)}} \cdot F_{O_{(ST,MP\ or\ CV)}} \cdot F_{M_{(ST,MP\ or\ CV)}} \quad (5.128)$$

$$\cancel{\eta_{adjusted_{(ST,\ MP\ or\ CV)}} = \eta_{default,ST} \cdot F_{D_{(ST,\ MP\ or\ CV)}} \cdot F_{O_{(ST,\ MP\ or\ CV)}} \cdot F_{M_{(ST,\ MP\ or\ CV)}}} \quad \cancel{(5.124)}$$

$$P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} = 1 - exp\left[-\left(\frac{t}{\eta_{adj,leak_{(ST,MP\ or\ CV)}}}\right)^{\beta_{ST}}\right] \quad (5.129)$$

$$P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} = 1 - exp\left[-\left(\frac{t}{\eta_{adj,cold_{(ST,MP\ or\ CV)}}}\right)^{\beta_{ST}}\right] \quad (5.130)$$

$$\cancel{P(t)_{f,final_{(ST,\ MP\ or\ CV)}} = 1 - exp\left(-\left(\frac{t}{\eta_{adjusted_{(ST,\ MP\ or\ CV)}}}\right)\right)^{\beta_{ST}}} \quad \cancel{(5.125)}$$

The adjusted η parameter ($\eta_{adj,leak_{(ST,MP\ or\ CV)}}$ and $\eta_{adj,cold_{(ST,MP\ or\ CV)}}$) is used to calculate the final (tailored) POF using Equation (5.129) for leakage and Equation (5.130) for blockage for each steam trap, mechanical pump or control valve operating within a steam system. The shape factor β_{ST} used in Equation (5.129 and 5.130) is the same shape factor generated from Table 7.4. Equation (5.129) and Equation (5.130) provides the final POF for each steam trap, mechanical pump or control valve in a steam using system.

Suggested adjustment multiplier categories that need to be considered for steam traps, mechanical pumps and control valves are given in Table 7.5 to Table 7.13. It should be noted that the value of each adjustment multiplier depends on engineering judgement.

7.3.5 Multiple Steam Trap or Mechanical Pumps or Control Valves Installations

For any steam using equipment, there are several associated lines with steam traps (or mechanical pumps or control valves) installed. The lines usually have steam traps installed in parallel or series. When there are multiple steam traps (or mechanical pumps or control valves) installed, the calculated POF for any one specific steam trap in the multiple installation will remain the same. However, the overall combined POF for leakage and blockage of multiple traps (in parallel or series) should be considered for each line using Equation (5.131 and 5.132) for traps in series and Equation (5.133 and 5.134) for traps in parallel. ~~However, the overall combined POF of multiple traps (parallel or series) shall be considered for each line using Equation (5.126) or Equation (5.127).~~

$$P(t)_{f,final\ series,leak_{(ST,MP\ or\ CV)}} = 1 - \left(1 - P(t)_{f1,leak}\right) \cdot \left(1 - P(t)_{f2,leak}\right) \cdot \ldots \cdot \left(1 - P(t)_{fn,leak}\right) \quad (5.131)$$

$$P(t)_{f,final\ series,cold_{(ST,MP\ or\ CV)}} = 1 - (1 - P(t)_{f1,cold}) \cdot (1 - P(t)_{f2,cold}) \cdot \ldots \cdot (1 - P(t)_{fn,cold}) \quad (5.132)$$

$$P(t)_{f,final\ parallel,leak_{(ST,MP\ or\ CV)}} = P(t)_{f1,leak} \cdot P(t)_{f2,leak} \cdot \ldots \cdot P(t)_{fn,leak} \quad (5.133)$$

$$P(t)_{f,final\ parallel,cold_{(ST,MP\ or\ CV)}} = P(t)_{f1,cold} \cdot P(t)_{f2,cold} \cdot \ldots \cdot P(t)_{fn,cold} \quad (5.134)$$

$$\cancel{P(t)_{f,final\ series_{(ST,\,MP\ or\ CV)}} = 1 - \left(1 - P(t)_{f_1}\right) \cdot \left(1 - P(t)_{f_2}\right) \quad (5.126)}$$

$$\cancel{P(t)_{f,final\ parallel_{(ST,\,MP\ or\ CV)}} = P(t)_{f_1} \cdot P(t)_{f_2} \quad (5.127)}$$

For example, Figure 7.3 is the sample arrangement of the traps showing their capacity. Calculation of the POF for each line is given by Equation (5.133) and Equation (5.134) which allow calculation of the total POF for the lines in parallel configuration. In addition, if the capacity of Trap 1 and Trap 2 are not sufficient for the equipment requirement individually, these two traps (or mechanical pumps or control valves) are treated as series configurations (Figure 7.3b) using Equation (5.131) and Equation (5.132).

~~7.3.4~~7.3.6 POF for Equipment

As discussed in Section 7.1.2, there are different types of equipment used in steam-using systems. Examples of some of these types were given in Table 7.1. In this section, the POF calculation due to steam related failure will be covered. Equipment consists of the following:

a) Heat exchanger

b) Distillation tower/column

c) Stripper

d) Flare

e) Steam turbine

f) Piping (steam main or condensate piping)

g) Tracing (instrumentation/relief valve)

The calculation of the POF of equipment takes into account the effect of both equipment and its associated lines. It is also important to note that the calculation assumes that each individual item of equipment is independent.

For example, Figure 7.4(a) shows an arrangement of a steam turbine with traps. A block diagram for combining the POF calculation for the same system is provided in Figure 7.4(b).

The equations below are used in estimating the POF for the equipment listed above and each equipment is considered independent and assessed separately.

$$\cancel{\eta_{adjusted\_equ} = \eta_{default\_equ} \cdot \left(F_{D_{equ}} \cdot F_{O_{equ}} \cdot F_{M_{equ}}\right)} \eta_{adj,equ} = \eta_{def,equ} \cdot \left(F_{D_{equ}} \cdot F_{O_{equ}} \cdot F_{M_{equ}}\right) \quad (5.135)$$

$$\cancel{P(t)_{f,final(equ)} = 1 - exp\left(-\left(\frac{t}{\eta_{adjusted\_equ}}\right)^{\beta_{equ}}\right)} \quad P(t)_{f,final(equ)} = 1 - exp\left[-\left(\frac{t}{\eta_{adj,equ}}\right)^{\beta_{equ}}\right] \tag{5.136}$$

The default scale parameter, $\eta_{def,equ}$ and shape parameter, β_{equ} are obtained from historical data analysis. Table 7.14 shows default Weibull parameters for the different types of steam-using equipment. The data presented in Table 7.14 are based on the best available sources and experience to date from owner-operators. However, it is recommended that other Weibull parameters be used by the owner-operator where plant specific data for default shape/scale parameters are available. The default parameters in Table 7.14 are suggested when plant specific data is unavailable and are based on failure of steam systems. The POF of the steam-using equipment, $P(t)_{f,def(equ)}$ is calculated using Equation (5.122) and parameters from Table 7.14.

Similar to the approach for steam traps discussed in Section 7.3.4.2, $\boldsymbol{\eta_{adj,equ}}$ is used to calculate the final (tailored) POF (Equation (5.129)) for steam-using equipment. The shape factor β_{equ} used in Equation (5.129) is the shape factor from Table 7.14. $\boldsymbol{P(t)_{f,final(equ)}}$ is the final POF of the steam-using equipment. The adjustment multiplier categories for each design/installation, $F_{D_{equ}}$, operational, $F_{O_{equ}}$, or maintenance history, $F_{M_{equ}}$, factors are given in Table 7.15 to Table 7.17, and are used to modify the default scale parameter, $\eta_{def,equ}$. It should be noted that the value of each adjustment multiplier depends on engineering judgement.

~~7.3.5~~7.3.7 POF for Steam-Using Systems

The total POF for steam-using systems is calculated using Equation (5.123) and ~~-~~Equation (5.124) where, $\boldsymbol{P(t)_{f,final,leak_{(ST,MP\ or\ CV)}}}$ and $\boldsymbol{P(t)_{f,final,leak_{(ST,MP\ or\ CV)}}}$ is calculated from Equation (5.129) or Equation (5.130~~-~~) for individual steam traps and for multiple steam traps the procedure in Section 7.3.5 is used

7.3.8 POF after Inspection

Weibull parameters for the failure on demand curves are determined based on the analysis of a sample set of data (Section 7.3.1). However, as inspection data is collected, these parameters may be adjusted for each device based on the actual inspection results. This approach assumes that the Weibull shape parameter, β, remains constant based on the historical data and adjusts the characteristic life, η, as inspection data are collected.

The effectiveness of inspection and testing is provided in Annex 2.F, Section 2.F.11.2, Table 2.F.11.1. The probability of succeeding the inspection prior to inspection is given by Equation (5.137) and Equation (5.138).

$$P(t)_{f,prior,leak} = 1 - P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} \tag{5.137}$$

$$P(t)_{f,prior,cold} = 1 - P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} \tag{5.138}$$

After inspection, the POF is updated based on the results. Use Equation (5.139) and (5.140) if the inspection results do not show the expected failure.

$$P(t)_{f,after,leak} = \left(1 - CF_{pass}\right) \cdot P(t)_{f,prior,leak} \tag{5.139}$$

$$P(t)_{f,after,cold} = \left(1 - CF_{pass}\right) \cdot P(t)_{f,prior,cold} \tag{5.140}$$

Use Equation (5.141) and (5.142) if the inspection confirms the expected failure.

$$P(t)_{f,after,leak} = \left(1 - CF_{pass}\right) \cdot P(t)_{f,prior,leak} + \left(P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} \cdot CF_{fail}\right) \quad (5.141)$$

$$P(t)_{f,after,cold} = \left(1 - CF_{pass}\right) \cdot P(t)_{f,prior,cold} + \left(P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} \cdot CF_{fail}\right) \quad (5.142)$$

Based on the outcome of the inspection and its effectiveness the updated probability of failure after inspection is calculated using equations in Table 7.19. The characteristic life ($\eta_{adj,leak_{(ST,MP\ or\ CV)}}$ and $\eta_{adj,cold_{(ST,MP\ or\ CV)}}$) is updated based on the outcome of the inspection using Equation (5.143) and Equation (5.144).

$$\eta_{upd,leak} = \frac{t}{\left(-\ln\left(1 - P(t)_{f,wgt,leak}\right)\right)^{\frac{1}{\beta_{ST}}}} \quad (5.143)$$

$$\eta_{upd,cold} = \frac{t}{\left(-\ln\left(1 - P(t)_{f,wgt,cold}\right)\right)^{\frac{1}{\beta_{ST}}}} \quad (5.144)$$

Where, β_{ST} is shape factor established earlier and t is the inspection interval. The updated characteristic life is then used in the calculation of the POF using equation (5.145) and (5.146).

$$P(t)_{f,upd,leak} = 1 - exp\left[-\left(\frac{t}{\eta_{upd,leak}}\right)^{\beta_{ST}}\right] \quad (5.145)$$

$$P(t)_{f,upd,cold} = 1 - exp\left[-\left(\frac{t}{\eta_{upd,cold}}\right)^{\beta_{ST}}\right] \quad (5.146)$$

~~7.3.5.1~~7.3.8.1 POF after Cleaning

The steam trap POF will be reduced after each cleaning. ~~The steam trap POF will be updated if the trap is periodically cleaned.~~For example, if the periodic cleaning is done at 0.5 years and at 0.6 years, the POF will be reduced to the same POF value as at 0.1 year. At 1.1 years, the POF will be equal to the POF at 0.1 years, etc.

7.3.9 POF Calculation Procedure

The following calculation procedure is used to determine the POF due to leak and blockage for steam traps and steam using equipment. The POF of each system is calculated as the combined effect of individual equipment with its associated traps for both leak and blockage.

a) STEP 1: Identify the steam traps, mechanical pumps and control valves and associated steam using equipment in the steam system. Provide required data defined in Table 7.3.

b) STEP 2: Calculate the POF for each steam traps, mechanical pumps and control valves for both failure modes:

   1. STEP 2.1: Determine the default values of the Weibull parameters for both failure modes from Table 7.4.

   ~~1.~~2. STEP 2.2: Using Table 7.5 to Table 7.13, determine the design, operating and maintenance condition adjustment for each item (steam trap, mechanical pump and control valve).

   3. STEP 2.3: Using Equation (5.127) and Equation (5.128), adjust the Weibull parameter $\eta_{def,ST}$ based on the values in STEP 2.2 for both failure modes.

4. STEP 2.4: Calculate $P(t)_{f,final,leak_{(ST,MP\ or\ CV)}}$ and $P(t)_{f,final,cold_{(ST,MP\ or\ CV)}}$ using Equation (5.129) and Equation (5.130) based on the adjusted Weibull parameter $\eta_{adj,leak_{(ST,MP\ or\ CV)}}$ and $\eta_{adj,cold_{(ST,MP\ or\ CV)}}$ using Equation (5.127) and Equation (5.128). Repeat for each steam trap, mechanical pump and control valve.

5. STEP 2.5: For steam traps, mechanical pumps and control valves installed in parallel or series use Equations (5.131) to (5.134) for both failure modes to calculate POF.

c) STEP 3: Inspection POF updating for each steam trap, mechanical pump and control valve for both failure modes. Repeat the following steps in case of multiple steam traps, mechanical pumps and control valves

~~2.~~1. STEP 3.1: Identify the effectiveness of the inspection and testing method using Annex 2.F, Section 2.F.11.2, Table 2.F.11.1.

2. STEP 3.2: Using Equation (5.137) and (5.138), calculate the probability of not failing the inspection prior to inspection for both failure modes.

3. STEP 3.3: Identify the confidence factor (CF) associated with the inspection effectiveness and inspection result using Table 7.18.

4. STEP 3.4: Calculate $P(t)_{f,after}$ for blockage and leakage failures using Equation (5.139) and (5.140) if the inspection results do not show the expected failure and Equation (5.141) and (5.142) if the inspection confirms the expected failure.

$$P(t)_{f,after} = \left(1 - CF_{pass}\right) \cdot P(t)_{f,prior}$$ ~~(5.131)~~

$$P(t)_{f,after} = \left(\left(1 - CF_{pass}\right) \cdot P(t)_{f,prior}\right) + \left(P(t)_{f,adjusted} \cdot CF_{fail}\right)$$ ~~(5.132)~~

~~4.~~5. STEP 3.5: Calculate $P(t)_{f,wgt}$ using the appropriate equation for inspection using Table 7.19 and based on the inspection effectiveness and inspection results.

6. STEP 3.6: Calculate the updated characteristic life, using Equation (5.143) and (5.144).

$$\eta_{upd} = \frac{t}{\left(-ln\left(1 - P(t)_{f,wgt}\right)\right)^{\frac{1}{\beta_{ST}}}}$$ ~~(5.133)~~

7. STEP 3.7: Calculate the POF at year in service using Equation (5.145) and (5.146).

$$P(t)_{f,upd} = 1 - exp\left(-\left(\frac{t}{\left(\eta_{upd}\right)}\right)\right)^{\beta_{ST}}$$ ~~(5.134)~~

~~5.~~8. STEP 3.8: Calculate the POF for both failure modes, at $t_{service(ST)}$ based on the steam trap arrangement using Equation (~~.~~5.131) and (5.132) for series or Equation (5.133) and (5.134) ~~-~~for parallel configuration.

d) STEP 4: Calculate the POF for each steam using equipment:

~~6.~~1. STEP 4.1: Using the default Weibull parameters for the steam using equipment from Table 7.14.

~~7.~~2. STEP 4.2: Using Table 7.15, determine the design condition adjustment, $F_{D_{equ}}$, for the steam using equipment.

~~8.~~3. STEP 4.3: Using Table 7.16, determine the operation condition adjustment, $F_{O_{equ}}$,for the steam using equipment.

~~9.~~4. STEP 4.4: Using Table 7.17, determine the maintenance history/inspection condition adjustment, $F_{M_{equ}}$, for the steam using equipment.

5. STEP 4.5: Using Equation (5.135), adjust the Weibull parameter, $\eta_{def,equ}$, based on the values in STEPS 4.2, 4.3 and 4.4.

6. STEP 4.6: Using Equation (5.136), calculate the, $P(t)_{f,final(equ)}$, for the steam using equipment based on the adjusted Weibull parameter, $\eta_{adj,equ}$.

e) STEP 5: Calculate the final POF for the steam using system using Equation (5.123) and (5.124) for both failure modes.

7.4 Consequence of Failure Methodology

7.4.1 Background

This section presents a procedure to calculate consequence of failure (COF) for a steam system.

7.4.2 Models for Assessing COF

7.4.2.1 Overview

The calculation of the COF is performed by evaluating costs involved in different failure consequences, such as the cost of the loss of inventory, regulatory cost, cost of downtime and cost of repairs. Failure will result in a consequence, i.e. potential impact on people, as well as product loss and component damage in some cases.

COF varies with different equipment and failure modes. The following sections provide the potential costs due to failures and outlines the COF calculation steps.

7.4.2.2 Cost of Steam Loss Due to Leakage

$$\cancel{FC_{loss,D/S} = \left(\frac{leakage\ rate(kg/hr)\cdot 8760\ (hr)\cdot cost\ of\ steam\ (\$/kg)}{1000}\right)}$$

$$FC_{loss} = \left(\frac{lrate \cdot 8760 \cdot FC_{steam}}{1000}\right) \tag{5.147}$$

The leakage rate $(lrate)$ is based on historical inspection data.

7.4.2.3 Cost of Condensate Loss Due to Downstream Equipment Rupture

$$\cancel{FC_{loss,DS} = \left(\frac{condensate\ mass(kg/hr) \cdot 8760\ (hr) \cdot cost\ of\ steam\ (\$/kg)}{1000} \right)}$$

$$FC_{condensate} = \left(\frac{mass_{condensate} \cdot 8760 \cdot FC_{steam}}{1000} \right) \quad (5.148)$$

The condensate mass $(mass_{condensate})$ is calculated following the procedure recommended in Part 3, Section 4.7.2, Equation (3.14).

7.4.2.4 Cost of component damage due to rupture caused by water hammer

The temporary default component damage cost uses the recommended values from Part 3, Section 4.12.2 for heat exchangers and steam tracing main pipes, and the North American Electric Reliability Corporation (NERC) Generating Availability Data System (GADS) for steam turbines. The default values are able to be customized by the user.

7.4.2.5 Cost of production loss due to shut down or reduced service efficiency

The production loss value can be manually assigned or calculated using Equation (5.149~~37~~).

$$FC_{prod} = Unit_{prod} \cdot \left(\frac{rate_{red}}{100} \right) \cdot D_{sd} \quad (5.149\cancel{37})$$

Where, $Unit_{prod}$ is the daily profit margin on the unit ($/day). This will be input by the user. $Rate_{red}$ is the production rate reduction on a unit as a result of the equipment being out of service (%), which will also be user input. D_{sd} is the number of days required to shut down a unit in order to repair the equipment during an unplanned shutdown.

7.4.2.6 Cost of safety impact to personnel due to rupture and leakage

The steam released through leakage or rupture may result in a safety impact on personnel. The total personnel injury cost, $CA_{f,inj}$, within a certain area is calculated using Equation (5.150~~38~~).

$$FC_{inj} = CA_{f,inj} \cdot popdens \cdot injcost \quad (5.150\cancel{38})$$

Where $CA_{f,inj}$ is calculated by using the procedure in Part 3, Section 4.10.2.

The hole size used to calculate the $CA_{f,inj}$ due to rupture from blockage is the inlet/connection size using Part 3, Equation (3.70). For leakage, the medium hole size of 1 in. (25 mm) is used to calculate $CA_{f,inj}$ due to leakage using in Part 3, Equation (3.69). The $popdens$ and $injcost$ used in Equation (5.150) is defined in Part 3, Section 4.12.5. The required input parameters are listed in Table 7.20.

For multiple traps, use Equations (5.151~~39~~) and Equation (5.152~~40~~) to calculate COF.

Blockage: $\cancel{FC_{inj} = max\left(FC_{inj\_1}, FC_{inj\_2}, \ldots FC_{inj\_n}\right)}$

$$FC_{inj,cold} = max(FC_{inj,cold_1}, FC_{inj,cold_2}, \ldots FC_{inj,cold\_n}) \quad (5.151)$$

Leak: $\cancel{FC_{inj} = \left(FC_{inj\_1} + FC_{inj\_2}, \ldots + FC_{inj\_n}\right)}$

$$FC_{inj,leak} = (FC_{inj,leak_1} + FC_{inj,leak_2} + \cdots FC_{inj,leak\_n}) \qquad (5.152)$$

7.4.3 Cost Models for Different Equipment

7.4.3.1 Overview

The financial COF varies for different equipment and failure modes. A list of potential costs due to failure and calculation methods was introduced in Section 7.4.2. For freshly added applications, the various potential failure consequences are added to the 'event tree' as the starting point for financial COF model development. The financial COF is calculated differently for steam distribution system depending on the type of equipment connected. Currently, 'type of connected equipment' is one of the data requirements for steam distribution COF calculation. Section 7.4.3.2 through Section 7.4.3.10 outline the calculation methodology for estimating financial COF for different equipment.

7.4.3.2 COF model for heat exchanger and steam turbine

The failure modes for heat exchanger and steam turbines can be either blockage or leakage and are calculated separately. The presence of an opening bypass for the steam system should be determined in the case of a blockage. If no opening bypass exists, a blockage could cause the steam system to shut down and may result in water hammer inside the equipment, causing a production loss and/or rupture. A rupture may cause a financial loss due to component damage and safety impact (personnel injury). The financial COF due to blockage without an opened bypass for heat exchanger and turbine is calculated using Equation (5.153~~41~~).

$$FC_{cold}^{HEX,Turbine} = FC_{prod} + FC_{comp} + FC_{inj} \qquad (5.153\cancel{41})$$

The consequence is calculated the same as a leakage consequence if a bypass is opened.

The total steam loss is calculated for both leakage and blockage with an open bypass. If the bypass is open, the safety impact is considered in addition to the loss of steam. Safety impact is not included for internal leakage.

If the outlet is closed while the traps are leaking, there will be a subsequent consequence of water hammer occurring to the downstream equipment/pipe in addition to steam loss from leaking traps. In the worst case, the downstream pipe will be ruptured. This will result in production loss due to downstream equipment shutdown, downstream pipe component damage, loss of condensate and associated safety impacts. The financial COF due to both leakage and blockage with an open bypass for a heat exchanger and turbine is calculated using Equation (5.154~~42~~) and Equation (5.155~~43~~):

$$\cancel{FC_{leak,open}^{HEX,Turbine} = FC_{loss} + FC_{loss,D/S}}\ FC_{leak,open}^{HEX,Turbine} = FC_{loss} \qquad (5.154)$$

$$\cancel{FC_{leak,closed}^{HEX,Turbine} = FC_{loss} + \left(FC_{prod,D/S} + FC_{comp,D/S} + FC_{inj,D/S}\right)}\ FC_{leak,closed}^{HEX,Turbine} = FC_{loss} + (FC_{prod,D/S} + FC_{comp,D/S} + FC_{inj,D/S}) \qquad (5.155)$$

7.4.3.3 COF model for general steam tracing

The failure modes for steam tracing equipment can be either blockage or leakage, which are calculated separately. Unlike a heat exchanger or turbine (as described in Section 7.4.3.2), the COF for tracing is considered for the main pipe and tracing line. When 'blockage' happens, it shall be established whether there is an opened bypass for the system or the trap is disconnected. If the bypass is closed or the trap is not disconnected, the blockage will cause the steam system to shut down or the content to cool down and possibly water hammer inside the tracing line. In one case, the steam system shut down and content sub-cooling will result in production loss in addition to the cost of main pipe cut-off (component damage). In another case, the water hammer inside the tracing line will cause the tracing line to rupture (worst case scenario), which will result in costs of the tracing line component damage in addition to associated safety impacts.

The COF due to blockage without opened bypass or trap disconnection for high temperature steam tracing is calculated using Equation (5.156~~44~~).

$$FC_{cold}^{Tracing,HT} = FC_{prod} + FC_{comp,main} + FC_{comp,line} + FC_{inj}$$

(5.156~~44~~)

If the bypass is opened or the trap disconnected, the consequence will be the same as the consequence of leakage.

For both leakage and blockage with an open bypass or trap disconnection, the calculation is the same as the consequence of leakage for a heat exchanger or turbine. The COF for both leakage and blockage with an open bypass or trap disconnection for high temperature steam tracing is calculated using Equation (5.157~~45~~) or Equation (5.158~~46~~).

$$FC_{leak,open}^{Tracing,HT} = FC_{loss} + FC_{inj}$$

(5.157~~45~~)

$$FC_{leak,closed}^{Tracing,HT} = FC_{loss} + \left(FC_{prod,D/S} + FC_{comp,D/S} + FC_{inj,D/S} \right)$$

(5.158~~46~~)

~~7.4.3.3~~7.4.3.4 COF Model for Low Temperature Steam Tracing

The failure modes can be either blockage or leakage, which will be calculated separately. The COF for tracing is considered for main pipe and tracing lines separately.

Similar to the high temperature tracing (Section 7.4.3.3), when blockage occurs, the COF is calculated using Equation (5.159~~47~~).

$$FC_{cold}^{Tracing,LT} = FC_{prod} + FC_{comp,main} + FC_{comp,line} + FC_{inj}$$

(5.159~~47~~)

For both leakage and blockage with open bypass or trap disconnection, the common failure consequence for both an open and closed system is as follows:

a) The steam leaking will result in costs from steam loss; if multiple traps are leaking, the sum of steam loss costs should be reported.

b) Leakage causes equipment shut down or overheating, which gives rise to costs from production loss.

Water hammer may occur inside the process line due to leakage may results in a rupture of the process line and costs from process line component damage and safety impact. The fluid within the process line is assigned as flammable or toxic or flammable and toxic. The semi-quantitative model to estimate safety COF is developed based on Part 3. If the fluid is both flammable and toxic, the worst case will be used.

In addition to costs listed above, for an open system (i.e. the outlet is opened), there are further safety impacts caused by leaking steam. If the outlet is closed, there is a subsequent consequence of water hammer occurring to the downstream equipment/pipe. The evaluation approach for this subsequent consequence is the same as the heat exchanger, turbine and high temperature tracing.

The COF due to both leakage and blockage with open bypass or trap disconnection for low temperature steam tracing is calculated using Equation (5.160~~48~~) and Equation (5.161~~49~~).

$$FC_{leak,open}^{Tracing,LT} = FC_{inj} + \left(FC_{loss} + FC_{comp,process} + FC_{prod,process} + FC_{inj,process} \right) \tag{5.160\cancel{48}}$$

$$\begin{aligned} FC_{leak,closed}^{Tracing,LT} &= \left(FC_{loss} + FC_{comp,process} + FC_{prod,process} + FC_{inj,process} \right) \\ &+ \left(FC_{prod,D/S} + FC_{comp,D/S} + FC_{inj,D/S} \right) \end{aligned} \tag{5.161\cancel{49}}$$

~~7.4.3.4~~7.4.3.5 COF model for steam tracing with relief valve

The relief valve is a type of valve used to control or limit the pressure in the steam tracing system. Pressure can build up as a result of a process, instrument or equipment failure. However, if the relief valve fails, there is the possibility the high pressure of the fluid within the pipe is raised further and causes leakage through the joints. In this case, the failure consequence is the sum of the cost of fluid loss and injury costs due to the leakage where the relief valve is installed (see Section 6.1.7). The financial COF calculation follows the COF equations for low temperature steam tracing.

~~7.4.3.5~~7.4.3.6 COF model for steam tracing with flow meter

A flow meter is an instrument used to measure linear, non-linear, volumetric or the mass flow rate of fluids, which can be found on both general tracing and low temperature applications. If the flow meter fails, the fluid is transported without measurement. This will not cause any safety consequence or financial loss in terms of product loss or component damage. However, without measurement, there may be a certain amount of business loss, which will be assessed by the user. In summary, the total financial COF is the same as for general tracing on a low temperature tracing system, with modified business loss which will be assessed by the user directly.

~~7.4.3.6~~7.4.3.7 COF model for distillation columns with stripping steam

The steam trap failure modes considered for distillation columns are leakage and blockage. For the failure mode of leakage when the outlet is open, financial COF is the sum of steam loss and cost of the safety impact due to condensate/steam discharge into the open air (Equation (5.154~~42~~)). If the outlet is closed, steam loss is the leakage financial COF (Equation (5.155~~43~~)). In terms of failure due to blockage when the bypass is not open, there is the possibility of condensate carry-over and/or water hammer, and the financial COF is calculated as the sum of component damage, production loss and the cost of safety impact using Equation (5.153~~41~~). If the bypass is open, the financial COF of due to blockage is the same as the COF of leakage.

~~7.4.3.7~~7.4.3.8 COF model for flare

The steam trap failure modes considered for flare are leakage and blockage. Similar to distillation columns (Section 7.4.3.7), if the steam trap of the flare leaks and its outlet is open, financial COF is the sum of steam loss and the cost of the safety impact due to condensate / steam discharge to the open air (Equation (5.154~~42~~)). Otherwise, if the outlet is closed, steam loss is the only leakage financial COF (Equation (5.155~~43~~)). In terms of failure due to blockage when the bypass is not open, there is the possibility of condensate carry-over and/or water hammer and the financial COF is calculated using Equation (5.153~~41~~) as the sum of component damage, production loss, the cost of safety impact due to pipe rupture and environmental costs due to reduced burning efficiency which will be assessed by the user directly using Equation (5.162~~50~~). If the bypass is open, the financial COF of due to blockage is the same as the COF of leakage.

$$FC_{inj} = max\left(FC_{inj,nfnt}, FC_{inj,flam}, FC_{inj,toxic}\right) \qquad (5.162\cancel{50})$$

~~7.4.3.8~~7.4.3.9 COF model for steam distribution piping

The failure modes considered for steam distribution piping are leakage and blockage. Similar to distillation columns (Section 7.4.3.7), if the steam trap of the main line leaks and its outlet is open, financial COF is the sum of steam loss and cost of the safety impact due to condensate/steam discharge to open air using Equation (5.154~~42~~). Otherwise, if the outlet is closed, steam loss is the only leakage financial COF using Equation (5.155~~43~~). In terms of failure due to blockage when the bypass is not open, there will be the possibility of water hammer; the financial COF is calculated as the sum of component damage (main line), production loss, and the cost of any safety impact (Equation (5.153~~41~~). If the bypass is open, the financial COF due to blockage is the same as the financial COF of leakage.

~~7.4.3.9~~7.4.3.10 COF model for condensate recovery line

The failure mode considered for the steam recovery line is leakage only. This is because blockage steam traps related to the recovery line are not discharging into the line, so they do not have any effect. When the recovery line fails due to a steam trap leakage, the condensate pipe may rupture due to water hammer. The financial COF is calculated as the sum of any component damage (pipe), cost of safety impact, condensate loss and downstream equipment production loss using Equation (5.158~~46~~).

7.4.4 COF calculation procedure

The following calculation procedure may be used to determine the financial consequence of failure (COF) for a steam system. The financial COF needs to be calculated for both failure modes.

a) STEP 1: Calculate the cost of steam loss due to leakage using Equation (5.147~~35~~).

b) STEP 2: Calculate the cost of condensate loss due to downstream equipment rupture using Equation (5.148~~36~~). Go to STEP 3, if no downstream equipment is connected or if the system is open i.e. the condensate is discharged to open.

c) STEP 3: Calculate the cost of production loss due to shut down or reduced service efficiency using Equation (5.149~~37~~).

d) STEP 4: Calculate the cost of safety impact to personnel due to rupture and leakage using Equation (5.150~~38~~)~~17~~. If there are multiple steam traps use Equation (5.151~~39~~) and Equation (5.152~~40~~).

e) STEP 5: Calculate the financial COF of component damage based on the type of steam using equipment as given in Section 7.4.3.2 to Section 7.4.3.10.

7.5 Risk Based Analysis

The risks due to leakage and blockage ~~to be considered are business loss and injury to people,~~ is calculated using Equation (5.163) and (5.164). Where the POF of steam system is calculated from Equations (5.123) and (5.124) for both leakage and blockage.

$$R(t)_{leak} = P(t)_{f,final,leak\ (steam\ using\ system)} \cdot FC_{leak} \tag{5.163}$$

$$R(t)_{cold} = P(t)_{f,final,cold\ (steam\ using\ system)} \cdot FC_{cold} \tag{5.164}$$

The total risk $R(t)$ is the sum of the risk due to blockage and leakage and is calculated from Equations (5.165).

$$R(t) = R(t)_{leak} + R(t)_{cold} \tag{5.165}$$

~~$R(t) = P(t)_{f,steam\ using\ system} \cdot FC$~~

~~Where $R(t)$ is the risk due to either blockage and/or leakage and the final risk is the sum of the risk arising from blockage and leakage. $P(t)_{f,steam\ using\ system}$ is obtained from Equation (5.123).~~ For the output, the risk is calculated as a function of time on a risk matrix. All of the post-assessment analysis are conducted based on this; this will be discussed in the following sections.

~~7.5~~7.6 Inspection and Risk Mitigation Planning

~~7.5.1~~7.6.1 Risk mitigation plan

~~7.5.2~~7.6.2 Overview

The mitigation plan comprises risk mitigation suggestions/actions to assist asset ~~users/owners~~owner-operator managing their steam system through the identification of the influence of each mitigation action on the system. The method for illustration of the risk target is the 'Iso-risk target'. the Iso-risk target is defined as a line of constant risk and a method of graphically showing POF and COF values in a log-log, two-dimensional plot where risk increases toward the upper right-hand corner. The value of the target risk will be determined by the user.

The possible mitigation actions listed in Section 7.6.2.1 to Section 7.6.2.3 are suggestions only and may not be applicable in all situations.

~~7.5.2.1~~7.6.2.1 Configuration of steam system

The risk can be modified by changing the configurations of the steam system, either by adding spare equipment or extra steam traps to the line or changing the type of the existing steam traps. The influence will depend on the number and location of the extra steam traps. Specifically, if extra steam traps are added, the arrangement of the steam system will be changed. The value of POF will be amended accordingly. Meanwhile, different steam traps will have a different $P(t)_{adjusted}$, which will affect the POF of the steam system (Equation (5.123)).

~~7.5.2.2~~7.6.2.2 Inspection

If an inspection is performed, or a condition monitoring device installed, the risk categories will also be shifted as the tailored characteristic life $\eta_{adjusted}$ will be updated accordingly. The procedure proposed in Section 7.3.9 will be followed. For sensors, the Confidence Factor, CF, value will be defaulted to 'usually effective'.

Cleaning of the steam trap has a significant impact on the POF; the more frequent the cleaning, the lower the POF over time.

~~7.5.2.3~~7.6.2.3 Spare equipment

If any spare equipment is included in one steam system, this may help to reduce the consequential cost of production loss. The POF can also be mitigated by intentionally releasing steam, e.g. via 'bypass open'. However, this action is not recommended due to environmental and safety viewpoints. In addition, it not only causes an increment of COF due to loss of steam, but could also lead to local corrosion damage i.e. FC_{loss} and FC_{comp} .

~~7.6~~7.7 Nomenclature

$CA_{f,inj}$	is the final personnel injury consequence area, ft^2 (m^2)
CF_{pass}	is the confidence factor for the inspection not to result in failure
CF_{fail}	is the confidence factor for the inspection results in failure
$mass_{condensate}$	is the condensate mass used in the consequence calculation associated with the nth release hole size, lb (kg)
$cost\ of\ steam$	is the cost of steam, \$/lb (\$/kg)
D_{sd}	is the time required to shut down a unit to perform a repair, days
$F_{D_{CV}}$	is the design adjustment multiplier for control valve
$F_{D_{equ}}$	is the design adjustment multiplier for steam using equipment
$F_{D_{MP}}$	is the design adjustment multiplier for mechanical pump
$F_{D_{ST}}$	is the design adjustment multiplier for steam traps
$F_{O_{CV}}$	is the operational adjustment multiplier for control valve
$F_{O_{equ}}$	is the operational adjustment multiplier for steam using equipment
$F_{O_{MP}}$	is the operational adjustment multiplier for mechanical pump
$F_{O_{ST}}$	is the operational adjustment multiplier for steam traps
$F_{M_{CV}}$	is the maintenance/inspection history adjustment multiplier for control valve
$F_{M_{equ}}$	is the maintenance/inspection history adjustment multiplier for steam using equipment
$F_{M_{MP}}$	is the Maintenance/inspection history adjustment multiplier for mechanical pump
$F_{M_{ST}}$	is the maintenance/inspection history adjustment multiplier for steam traps
FC	is the final financial consequence, \$
FC_{comp}	is the cost of component damage, \$
$FC_{comp,D/S}$	is the cost of component damage(downstream), \$
$FC_{comp,line}$	is the cost of component damage (tracing line), \$
$FC_{comp,main}$	is the cost of component damage (main pipe), \$
$FC_{comp,process}$	is the cost of component damage(process line), \$
$FC_{cold}^{HEX,Turbine}$	is the financial consequence of failure of heat exchanger and turbine due to blockage, \$
$FC_{leak,open}^{HEX,Turbine}$	is the financial consequence of failure of heat exchanger and turbine due to leakage (open system), \$
$FC_{leak,closed}^{HEX,Turbine}$	is the financial consequence of failure of heat exchanger and turbine due to leakage (closed system), \$
FC_{inj}	is the financial consequence as a result of serious injury to personnel, \$
$FC_{inj,cold}$	is the financial consequence due to blockage as a result of serious injury to personnel, \$

$FC_{inj,leak}$ is the financial consequence due to leakage as a result of serious injury to personnel, $

$FC_{inj,D/S}$ is the financial consequence as a result of serious injury to personnel (downstream), $

$FC_{inj,flam}$ is the financial consequence of as a result of serious injury to personnel due to flammable release, $

FC_{inj_n} is the financial consequence as a results of serious injury to personnel, $ for steam trap n

$FC_{inj,nfnt}$ is the financial consequence as a result of serious injury to personnel due to non-flammable, non-toxic, $

$FC_{inj,process}$ is the financial consequence as a result of serious injury to personnel (process line), $

$FC_{inj,toxic}$ is the financial consequence of as a result of serious injury to personnel due to toxic release, $

FC_{loss} is the cost of steam, $

$FC_{loss,D/S}$ is the cost of condensate loss (downstream), $

FC_{prod} is the cost of production loss, $

$FC_{prod,D/S}$ is the cost of production loss (downstream), $

$FC_{prod,process}$ is the cost of production loss (process line), $

$FC_{cold}^{Tracing,HT}$ is the financial consequence of failure of high temperature tracing due to blockage, $

$FC_{leak,open}^{Tracing,HT}$ is the financial consequence of failure of high temperature tracing due to leakage (open system), $

$FC_{leak,closed}^{Tracing,HT}$ is the financial consequence of failure of high temperature tracing due to leakage (closed system), $

$FC_{cold}^{Tracing,LT}$ is the financial consequence of failure of low temperature tracing due to blockage, $

$FC_{leak,open}^{Tracing,LT}$ is the financial consequence of failure of low temperature tracing due to leakage (open system), $

$FC_{leak,closed}^{Tracing,LT}$ is the financial consequence of failure of low temperature tracing due to leakage (closed system), $

FC_{steam} is the cost of steam, $/lb ($/kg)

$lrate$ Leakage rate is based on historical inspection data, lb/hr (kg/hr)

$injcost$ is cost of personnel injury per individual, $

$P(t)_{f,final,leak\ (steam\ using\ system)}$ is the probability of failure for steam using system due to leakage, failure/year

$P(t)_{f,final,cold\ (steam\ using\ system)}$ is the probability of failure for steam using system due to blockage, failure/year

$P(t)_{f,final,leak(ST,MP\ or\ CV)}$ is the tailored probability of failure due to leakage calculated for the associated lines (combined POF), consisting of multiple steam traps, mechanical pumps and control valves, failure/year

$P(t)_{f,final,cold(ST,MP\ or\ CV)}$ is the tailored probability of failure due to blockage calculated for the associated lines (combined POF), consisting of multiple steam traps, mechanical pumps and control valves, failure/year

$P(t)_{f,def,leak}$ is the probability of failure due to leakage of steam traps mechanical pumps and control valves based on default values for Weibull parameters, failure/year

$P(t)_{f,def,cold}$ is the probability of failure due to leakage of steam traps mechanical pumps and control valves based on default values for Weibull parameters, failure/year

$P(t)_{fn,leak}$ is the probability of failure due to leakage of steam traps mechanical pumps and control valves, n in series or parallel configurations, failure/year

$P(t)_{fn,cold}$ is the probability of failure due to blockage of steam traps mechanical pumps and control valves, n in series or parallel configurations, failure/year

$P(t)_{f,final\ series,leak_{(ST,MP\ or\ CV)}}$ is the probability of failure due to leakage for multiple steam traps, mechanical pumps and control valves in series, failure/year

$P(t)_{f,final\ series,cold_{(ST,MP\ or\ CV)}}$ is the probability of failure due to blockage for multiple steam traps, mechanical pumps and control valves in series, failure/year

$P(t)_{f,final\ parallel,leak_{(ST,MP\ or\ CV)}}$ is the probability of failure due to leakage for multiple steam traps, mechanical pumps and control valves in parallel, failure/year

$P(t)_{f,final\ parallel,cold_{(ST,MP\ or\ CV)}}$ is the probability of failure due to blockage for multiple steam traps, mechanical pumps and control valves in parallel, failure/year

$P(t)_{f,prior,leak}$ is the probability of not failing due to leakage the inspection prior to inspection, failure/year

$P(t)_{f,prior,cold}$ is the probability of not failing due to blockage the inspection prior to inspection, failure/year

$P(t)_{f,after,leak}$ is the probability of failure due to leakage after inspection depending on the results, failure/year

$P(t)_{f,after,cold}$ is the probability of failure due to blockage after inspection depending on the results, failure/year

$P(t)_{f,upd,leak}$ is the probability of failure due to leakage used for inspection updating, failure/year

$P(t)_{f,upd,cold}$ is the probability of failure due to blockage used for inspection updating, failure/year

$P(t)_{f,wgt,leak}$ is the updated probability of failure due to leakage after inspection, failure/year

$P(t)_{f,wgt,cold}$ is the updated probability of failure due to blockage after inspection, failure/year

$P(t)_{f,final(equ)}$ is the tailored probability of failure calculated for the steam using equipment, failure/year

$popdens$ is the population density of personnel or employees in the unit, personnel/ft^2 (personnel/m^2)

$Rate_{red}$ is the production rate reduction on a unit as a result of the equipment being out of service (%)

$R(t)_{leak}$ is the risk due to leakage as a function of time, $/year

$R(t)_{cold}$ is the risk due to blockage as a function of time, $/year

$R(t)$ is the risk as a function of time, $/year

t is the time at which the risk is to be calculated, years

$Unit_{prod}$ is the unit production margin ($/day)

β is the Weibull shape parameter estimated using AFT model

β_{equ} is the shape factor for equipment from Table 7.14

β_{ST} is the shape factor for steam traps, mechanical pumps and control valves from Table 7.4

η is the Weibull characteristic life parameter, years

$\eta_{def,leak,ST}$ is the scaled parameter for leakage estimated using Weibull AFT model from Table 7.4, years

$\eta_{def,cold,ST}$ is the scaled parameter for blockage estimated using Weibull AFT model from Table 7.4, years

$\eta_{adj,leak(ST,MP\ or\ CV)}$ is the tailored characteristic life (scale factor) for leakage based on condition of design/installation, operation or maintenance history factors for equipment, years

$\eta_{adj,cold(ST,MP\ or\ CV)}$ is the tailored characteristic life (scale factor) for blockage based on condition of design/installation, operation or maintenance history factors for equipment, years

$\eta_{adj,equ}$ is the tailored characteristic life (scale factor) based on condition of design/installation, operation, or maintenance history factors for equipment, years

$\eta_{def,equ}$ is the scaled parameter for equipment estimated using Weibull AFT model from Table 7.14, years

$\eta_{upd,leak}$ is the updated characteristic life for leakage after inspection results, years

$\eta_{upd,cold}$ is the updated characteristic life for blockage after inspection results, years

~~7.7~~7.8 Tables

Table 7.1 – Steam-Using Application Groups and Equipment Examples

Application Group	Equipment Example	Process Application Examples
Steam heated equipment	Process Heat Exchanger	Alkylation, distillation, gas recovery, isomerization, visbreaking, coking, storage tank heating
Direct steam application	Distillation Tower	Distillation, fractionation
	Stripper	Crude and vacuum distillation, catalytic cracking, catalytic reforming, asphalt processing, lube oil processing, hydrogen treatment
	Flare	Air-assisted flares, pressure-assisted flares, enclosed ground flares,
Steam driven equipment	Steam Turbine	Power generation, compressor mechanical drive, hydrocracking, naphtha reforming, pump mechanical drive
Steam distribution piping	Piping	Piping to distribute steam and condensate recovery
Steam tracing	Tracing	Utility stations, steam and condensate piping

Table 7.2 – Steam Trap Types for Each of Three Categories of Steam Trap

Steam trap category	Common applications	Steam trap type
Mechanical steam traps	The mainstream of traps used today on equipment that requires large discharge capacities. Temperature/pressure controlled applications with fluctuating loads	Free float
		Lever float
		Inverted bucket
Thermostatic steam traps	Where condensate back-up can be tolerated or is required in order to remove excess enthalpy, e.g. non-critical tracing	Bimetal
		Balanced pressure trap
Thermodynamic steam traps	Tracing, drip, and certain light process steam applications	Thermodynamic Disc
		Thermodynamic Piston

Table 7.3 – Basic Data Needed for POF Calculation of Steam-Using System

Data	Description	Data Source

Steam trap type	Type of steam trap:
	• Mechanical steam traps
	  ○ Free float
	  ○ Lever float
	  ○ Inverted bucket
	• Thermostatic steam traps
	  ○ Bimetal
	  ○ Balanced pressure trap
	• Thermodynamic steam traps
	  ○ Thermodynamic Disc
	  ○ Thermodynamic Piston
		User Specified

Steam trap/ mechanical pump or control valve design, operational and maintenance/ inspection history conditions	Data required on whether the following conditions apply: • Design conditions exceed maximum allowable pressure or maximum allowable temperature (PMA/TMA); • Steam trap configuration and capacity of individual steam traps; • Possibility of steam locking; • Any line bundling (i.e. inlet tracing line is heated by other bundled pipes); • No protection from weather; • Poor installation environment (i.e. higher than average failure rate at this location or area); • No strainer exists; • Trap is made of stainless steel (any grade); • Internal and/or external strainer upstream of steam trap is installed; • Operation conditions do not exceed maximum operating pressure or maximum operating temperature (PMO/TMO); • Operational stability is high, i.e. pressure/temperature/flow rate does not vary during normal operation; • Water hammer near the trap is recorded; • Disassembly preventive maintenance exists ; • Built-in integral/self-cleaning exists.	User Specified
Steam system inspection history	• Date of testing • Type of test (Effectiveness) • Results of test/inspection • Overhauled?	User Specified
Steam-Using Equipment	Steam-using equipment: • Steam Turbine • Heat Exchanger • Tracing – General • Tracing – Low Temperature (lower than 176^{o}F (80^{o}C)) • Tracing – Instrumentation • Tracing – Relief Valve • Steam Main Line • Condensate Line (Recovery) • Flare	Fixed Equipment

Data	Description	Data Source
	• Distillation Column	
Equipment Details	Operating conditions Design conditions Dimensions	User Specified

Table 7.4 – Default Weibull Parameters for Different Steam Traps, Control Valve and Mechanical Pump

Steam Trap Category	Steam Trap Type	Default β_{ST}	Default value for Leakage failure mode $\eta_{def,leak,ST}$	Default value for Blockage failure mode $\eta_{def,cold,ST}$
Mechanical steam traps	Free Float	1.8	16.1	13.8
	Inverted bucket	1.6	16.1	13.8
	Lever Float	1.7	11.7	8.5
Thermostatic steam traps	Bimetal	1.8	8	7.5
	Balanced Pressure	2	5.3	5.2
Thermodynamic steam traps	Disc	2	9.4	5
	Impulse	2	9.4	5
Control valve	-	1.8	61.5	61.5
Mechanical Pump	-	1.2	3.1	3.1

~~Steam Trap Category~~	~~Steam Trap Type~~	~~Failure Mode~~	~~Default~~ $\cancel{\beta_{ST}}$	~~Default~~ $\cancel{\eta_{default,ST}}$
~~Mechanical steam traps~~	~~Free Float~~	~~Blocked~~	~~1.8~~	~~13.8~~
		~~Leak~~		~~16.1~~
	~~Inverted bucket~~	~~Blocked~~	~~1.6~~	~~13.8~~
		~~Leak~~		~~16.1~~
	~~Lever Float~~	~~Blocked~~	~~1.7~~	~~8.5~~
		~~Leak~~		~~11.7~~
~~Thermostatic steam traps~~	~~Bimetal~~	~~Blocked~~	~~1.8~~	~~7.5~~
		~~Leak~~		~~8~~
	~~Balanced Pressure~~	~~Blocked~~	~~2~~	~~5.2~~
		~~Leak~~		~~5.3~~
~~Thermodynamic steam traps~~	~~Disc~~	~~Blocked~~	~~2~~	~~5~~
		~~Leak~~		~~9.4~~
	~~Impulse~~	~~Blocked~~	~~2~~	~~5~~
		~~Leak~~		~~9.4~~
~~Control valve~~		~~Blocked/leak~~	~~1.8~~	~~61.5~~
~~Mechanical Pump~~		~~Blocked/leak~~	~~1.2~~	~~3.1~~

Table 7.5 – Design Condition Adjustment for Steam Trap

Design Condition	Description	Adjustment Multiplier for design conditions, $F_{D_{ST}}$
Poor	If all of the below criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. If any line bundling d. No protection from weather e. Poor installation environment f. No strainer exists	0.5
Average	If any of the following criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. If any line bundling d. No protection from weather e. Poor installation environment f. No strainer exists	0.85
Good	If none of the following criteria are true AND the trap is not made of Stainless Steel (any grade) AND internal or external strainer is installed: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. If any line bundling d. No protection from weather e. Poor installation environment f. No strainer exists	1.0
Very Good	If none of the following criteria are true AND the trap is made of Stainless Steel (any grade) AND both internal and external strainer is installed: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. If any line bundling d. No protection from weather e. Poor installation environment f. No strainer exists	1.15
Steam locking: equipment configuration causing steam-condensate mixture entering the trap or piping configuration causing steam to move ahead of condensate into the trap. Line bundling: inlet tracing line is heated by other bundled pipes. Poor installation environment: higher than average failure rate at this location or area.		

Table 7.6 – Operation Condition Adjustment for Steam Trap

Operation Condition	Description	Adjustment Multiplier for design conditions, $F_{O_{ST}}$
Poor	If operation conditions exceed PMO / TMO AND operational stability is low (i.e. > 50% operation load variations expected)	0.77
Average	If operation conditions do not exceed PMO / TMO AND operational stability is medium (i.e. ≤ 50% operation load variations expected)	0.85
Good	If operation conditions does not exceed PMO / TMO AND operational stability is high (i.e. no operation load variations expected)	1

Table 7.7 – Maintenance History/Inspection Condition Adjustment for Steam Trap

Maintenance Condition	Description	Adjustment Multiplier for design conditions, $F_{M_{ST}}$
Poor	If water hammer near the trap (i.e. within 10 m) is recorded in the past AND no disassembly preventive maintenance exists.	0.65
Average	If water hammer near the trap (i.e. within 10 m) is recorded in the past AND disassembly preventive maintenance exists	0.72
Good	If water hammer near the trap (i.e. within 10 m) is not recorded AND disassembly preventive maintenance does not exist AND built-in manual cleaning exists	1.0
Very Good	If water hammer near the trap (i.e. within 10 m) is not recorded AND disassembly preventive maintenance exists AND built-in integral/self-cleaning exists	1.1

Table 7.8 – Design Condition Adjustment for Mechanical Pump

Design Condition	Description	Adjustment Multiplier for design conditions, $F_{D_{MP}}$
Poor	If all of the below criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment d. System installation is non-ideal	0.5
Average	If any of the following criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment d. System installation is non-ideal	0.8
Good	If none of the following criteria are true AND the trap is not made of Stainless Steel (any grade) AND system installation is average: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment	1.0
Very Good	If none of the following criteria are true AND the trap is made of Stainless Steel (any grade) AND system installation is ideal AND strainer installed: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment	1.25
System installation is non-ideal: functionality is affected by sizing or configuration		

Table 7.9 – Operation Condition Adjustment for Mechanical Pump

Operation Condition	Description	Adjustment Multiplier for design conditions, $F_{O_{MP}}$
Poor	If operation conditions exceed PMO / TMO AND operational stability is low (i.e. > 50% operation load variations expected) AND pump load is high (i.e. > 75% of pump capacity)	0.76
Average	If operation conditions do not exceed PMO / TMO AND operational stability is medium (i.e. ≤ 50% operation load variations expected) OR pump load is medium (i.e. 50 – 75% of pump capacity)	1.2
Good	If operation conditions do not exceed PMO / TMO AND operational stability is high (i.e. no operation load variations expected) AND pump load is low (i.e. < 50% of pump capacity)	1.6

Table 7.10 – Maintenance History/Inspection Condition Adjustment for Mechanical Pump

Maintenance Condition	Description	Adjustment Multiplier for design conditions, $F_{M_{MP}}$
Poor	If water hammer near the pump (i.e. within 10 m) is recorded in the past	0.65
Average	If water hammer near the pump (i.e. within 10 m) is not recorded AND disassembly preventive maintenance does not exist	1
Good	If water hammer near the pump (i.e. within 10 m) is not recorded AND disassembly preventive maintenance exists	2

Table 7.11 - Design Condition Adjustment for Control Valve

Design Condition	Description	Adjustment Multiplier for design conditions, $F_{D_{CV}}$
Poor	If all of the below criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment (i.e. higher than average failure rate at this location or area)	0.6
Average	If any of the following criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment (i.e. higher than average failure rate at this location or area)	0.75
Good	If none of the following criteria are true: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment (i.e. higher than average failure rate at this location or area)	1.0
Very Good	If none of the following criteria are true AND the trap is made of Stainless Steel (any grade) AND strainer installed: a. Design conditions exceed PMA / TMA b. Possibility of steam locking c. Poor installation environment (i.e. higher than average failure rate at this location or area)	1.3

Table 7.12 – Operation Condition Adjustment for Control Valve

Operation Condition	Description	Adjustment Multiplier for design conditions, $F_{O_{CV}}$
Poor	If operation conditions exceed PMO / TMO AND operational stability is low (i.e. > 50% operation load variations expected) AND load is high (i.e. > 75% of valve capacity)	0.77
Average	If operation conditions do not exceed PMO / TMO AND operational stability (i.e. ≤ 50% operation load variations expected) is medium OR load is medium (i.e. 50 – 75% of valve capacity)	0.9
Good	If operation conditions do not exceed PMO / TMO AND operational stability is high (i.e. no operation load variations expected) AND load is low (i.e. < 50% of valve capacity)	1.0

Table 7.13 – Maintenance History/Inspection Condition Adjustment for Control Valve

Maintenance Condition	Description	Adjustment Multiplier for design conditions, $F_{M_{CV}}$
Poor	If water hammer near the trap (i.e. within 10 m) is recorded in the past	0.65
Average	If water hammer near the trap (i.e. within 10 m) is not recorded AND disassembly preventive maintenance does not exist	1
Good	If water hammer near the trap (i.e. within 10 m) is not recorded AND disassembly preventive maintenance exists	1.1

Table 7.14 – Default Weibull Parameters for Steam-Using Equipment

Equipment	Default $\eta_{def,equ}$	Default β_{equ}
Steam Turbine	34.48	3
Heat Exchanger	22.73	3
Tracing – Instrumentation	52.63	3
Tracing – Relief Valve	55.56	3
Steam header	25.1	3
Condensate Line (Recovery)	21.5	3
Distillation Column	37	3
Flare	13.3	3

Table 7.15 – Design Condition Adjustment for Steam-Using Equipment

Design Condition	Description	Adjustment Multiplier for design conditions, $F_{D_{equ}}$
Poor	If all of the below criteria are true: a. No inlet steam separator b. No appropriate steam trap (type and capacity) is installed c. Major reduction in number of steam traps (as per design) d. No automatic/manual start function e. One or more locations on steam supply that require condensate drainage cannot discharge continuously	0.5
Average	If any of the following criteria are true: a. No inlet steam separator b. No appropriate steam trap (type and capacity) is installed c. Major reduction in number of steam traps (as per design) d. No automatic/manual start function e. One or more locations on steam supply that require condensate drainage cannot discharge continuously	0.7
Good	If none of the below criteria are true AND steam traps are not equipped with by-pass: a. No inlet steam separator b. No appropriate steam trap (type and capacity) is installed c. Major reduction in number of steam traps (as per design) d. No automatic/manual start function e. One or more locations on steam supply that require condensate drainage cannot discharge continuously	1.0
Very Good	If none of the below criteria are true AND all steam traps equipped with by-pass a. No inlet steam separator b. No appropriate steam trap (type and capacity) is installed c. Major reduction in number of steam traps (as per design) d. No automatic/manual start function e. One or more locations on steam supply that require condensate drainage cannot discharge continuously	1.1

Table 7.16 – Operation Condition Adjustment for Steam-Using Equipment

Operation Condition	Description	Adjustment Multiplier for design conditions, $F_{O_{equ}}$
Poor	If all of the below criteria are true: a. Superheat rate < 18°F (10°C) b. Cyclic operation c. Exceed PMO/TMO/Steam Mass d. In the case of turbine: superheat rate < 27°F (15ºC) AND (for condensing turbine only) operating vacuum > 25% weaker than design e. In the case of heat exchanger: superheat rate is ≥ 18°F (10°C) AND steam passing through outlet control valve (if existing) AND > 50% operation load variations expected AND stall condition exists (i.e. insufficient different pressure)	0.45
Average	If minimum of 4 criteria from the below are true: a. Superheat rate < 10°C (18°F) b. Cyclic operation c. Exceed PMO/TMO/Steam Mass d. In the case of turbine: superheat rate < 27°F (15ºC) AND (for condensing turbine only) operating vacuum > 25% weaker than design e. In the case of heat exchanger: superheat rate is ≥ 18°F (10°C) AND steam passing through outlet control valve (if existing) AND > 50% operation load variations expected AND stall condition exists (i.e. insufficient different pressure)	0.7
Good	If minimum of 2 criteria from the below are true: a. Superheat rate < 18°F (10°C) b. Cyclic operation c. Exceed PMO/TMO/Steam Mass d. In the case of turbine: superheat rate < 27°F (15°C) AND (for condensing turbine only) operating vacuum > 25% weaker than design e. In the case of heat exchanger: superheat rate is ≥ 18°F (10°C) AND steam passing through outlet control valve (if existing) AND > 50% operation load variations expected AND stall condition exists (i.e. insufficient different pressure)	0.85
Very Good	If none of the below criteria is true: a. Superheat rate < 18°F (10°C) b. Cyclic operation c. Exceed PMO/TMO/Steam Mass d. In the case of turbine: superheat rate < 27°F (15°C) AND (for condensing turbine only) operating vacuum > 25% weaker than design e. In the case of heat exchanger: superheat rate is ≥ 18°F (10°C) AND steam passing through outlet control valve (if existing) AND > 50% operation load variations expected AND stall condition exists (i.e. insufficient different pressure)	1.0

Table 7.17 – Maintenance History/Inspection Condition Adjustment for Steam-Using Equipment

Maintenance Condition	Description	Adjustment Multiplier for design conditions, $F_{M_{equ}}$
Poor	Ongoing likelihood of water hammer AND damage/repair AND trips reported previously AND no maintenance conducted as recommended	0.4
Average	Low likelihood of water hammer AND damage/repair AND trips reported previously AND no maintenance conducted as recommended	0.6
Good	No likelihood of water hammer AND damage/repair AND trips not reported previously in previous AND maintenance recommendations are all conducted	1.0

Table 7.18 – Level of Inspection Confidence Factor for Steam Traps, Mechanical Pumps and Control Valves

Inspection results	Confidence Factor that Inspection Result Determines the True Damage State, CF				
	Ineffective	Poorly Effective	Fairly Effective	Usually Effective	Highly Effective
Leak detected, CF_{fail}	No credit	0.3	0.6	0.85	0.95
Leak not detected, CF_{pass}	No credit	0.3	0.6	0.75	0.9
Blocked, CF_{fail}	No credit	0.3	0.6	0.85	0.95
Not Blocked, CF_{pass}	No credit	0.3	0.6	0.85	0.95

Table 7.19 – Equations for Updating POF After Inspection

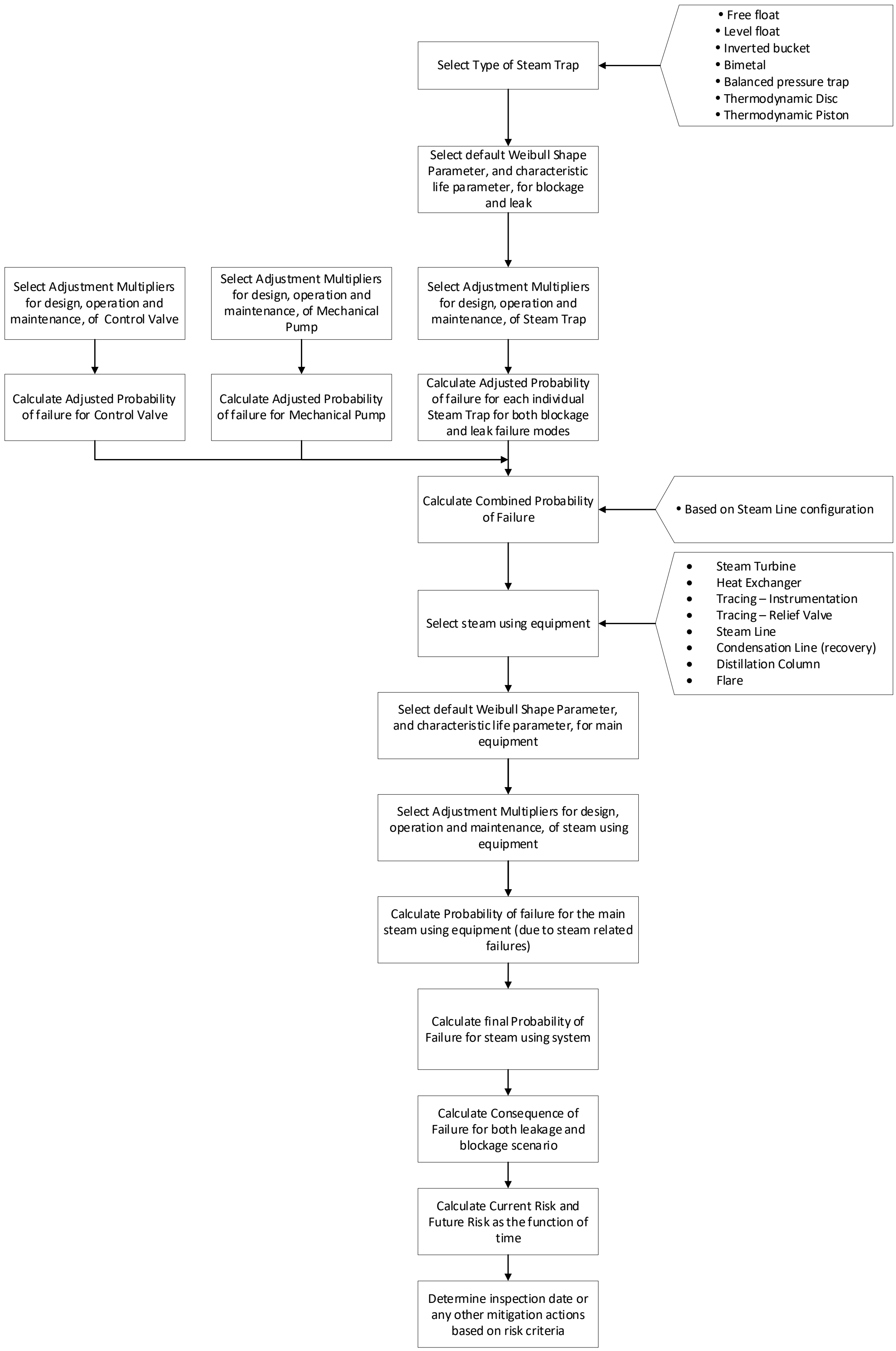
Inspection Effectiveness	Inspection results	Equation for updating the POF after inspection
Highly effective Usually effective Fairly effective Poorly Effective	No leakage or blockage detected	$P(t)_{f,wgt,leak} = P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} - 0.2 \cdot P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} \left(\frac{t}{\eta_{adj,leak_{(ST,MP\ or\ CV)}}}\right) + 0.2 \cdot P(t)_{f,final,leak_{(ST,MP\ or\ CV)}} \left(\frac{t}{\eta_{adj,leak_{(ST,MP\ or\ CV)}}}\right)$

		$P(t)_{f,wgt,cold} = P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} - 0.2 \cdot P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} \left(\frac{t}{\eta_{adj,cold_{(ST,MP\ or\ CV)}}}\right) + 0.2 \cdot P(t)_{f,final,cold_{(ST,MP\ or\ CV)}} \left(\frac{t}{\eta_{adj,cold_{(ST,MP\ or\ CV)}}}\right)$
Highly effective	Leakage or blockage detected	$P(t)_{f,wgt,leak} = P(t)_{f,after,leak}$ $P(t)_{f,wgt,cold} = P(t)_{f,after,cold}$
Usually effective		
Fairly effective		$P(t)_{f,wgt,leak} = \left(0.5 \cdot P(t)_{f,final,leak_{(ST,MP\ or\ CV)}}\right) + \left(0.5 \cdot P(t)_{f,after,leak}\right)$ $P(t)_{f,wgt,cold} = \left(0.5 \cdot P(t)_{f,final,cold_{(ST,MP\ or\ CV)}}\right) + \left(0.5 \cdot P(t)_{f,after,cold}\right)$
Poorly Effective		

Table 7.20 – Required Data for COF Assessment

Cost Description	Data Source
Cost of steam, $/kg ($FC_{steam}$)	User required
Leakage rate is based on historical inspection data, lb/hr (kg/hr) (lrate)	User required
Cost of personnel injury per individual as per Part 3, Section 4.12.5, $ ($injcost$)	User required
Population density of personnel or employees in the unit as per Part 3, Section 4.12.5, personnel/ft^2 ($popdens$)	User required
Inspection interval, 8760 hours IF not defined by user	User required
Daily production margin, $Unit_{prod}$, on the unit ($/day)	User required
Production rate reduction, $Rate_{red}$, on a unit as a result of the equipment being out of service (%)	User required
The number of days, D_{sd} , required to shut a unit down to repair the equipment during an unplanned shutdown, days	User required
The cost of production loss from downstream equipment, $ ($FC_{prod,D/S}$)	User required
The cost of production loss in process lines, $ ($FC_{prod,process}$)	User required
Component damage costs, applies to the cost of all downstream equipment as in Table 7.14, $. (FC_{comp} , $FC_{comp,line}$, $FC_{comp,main}$, $FC_{comp,process}$, $FC_{comp,D/S}$)	User required

7.9 Figures


Select Type of Steam Trap
• Free float
• Level float
• Inverted bucket
• Bimetal
• Balanced pressure trap
• Thermodynamic Disc
• Thermodynamic Piston
Select default Weibull Shape Parameter, and characteristic life parameter, for blockage and leak
Select Adjustment Multipliers for design, operation and maintenance, of Control Valve
Select Adjustment Multipliers for design, operation and maintenance, of Mechanical Pump
Select Adjustment Multipliers for design, operation and maintenance, of Steam Trap
Calculate Adjusted Probability of failure for Control Valve
Calculate Adjusted Probability of failure for Mechanical Pump
Calculate Adjusted Probability of failure for each individual Steam Trap for both blockage and leak failure modes
Calculate Combined Probability of Failure
• Based on Steam Line configuration
Select steam using equipment
• Steam Turbine
• Heat Exchanger
• Tracing – Instrumentation
• Tracing – Relief Valve
• Steam Line
• Condensation Line (recovery)
• Distillation Column
• Flare
Select default Weibull Shape Parameter, and characteristic life parameter, for main equipment
Select Adjustment Multipliers for design, operation and maintenance, of steam using equipment
Calculate Probability of failure for the main steam using equipment (due to steam related failures)
Calculate final Probability of Failure for steam using system
Calculate Consequence of Failure for both leakage and blockage scenario
Calculate Current Risk and Future Risk as the function of time
Determine inspection date or any other mitigation actions based on risk criteria

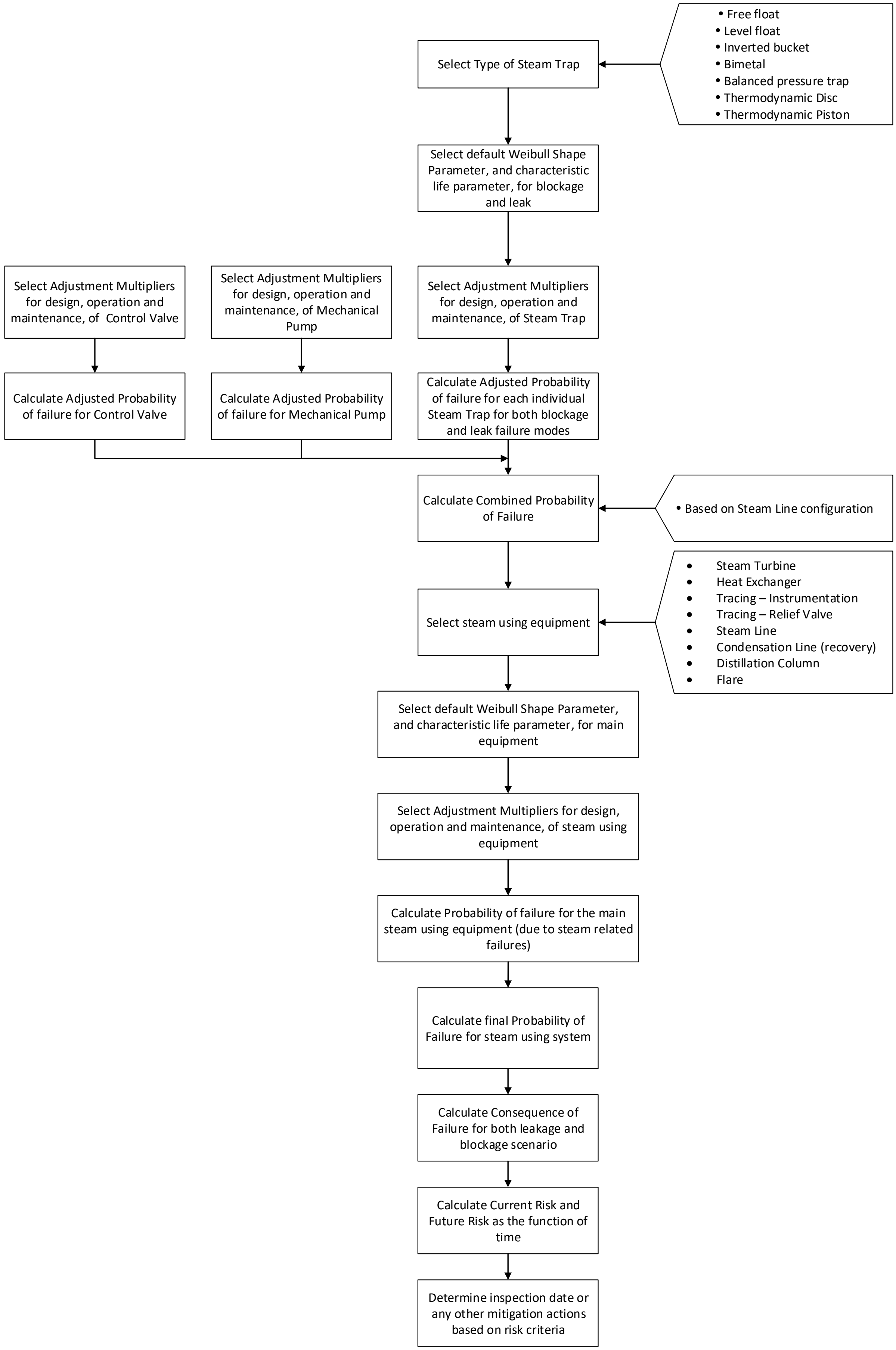

Select Type of Steam Trap
• Free float
• Level float
• Inverted bucket
• Bimetal
• Balanced pressure trap
• Thermodynamic Disc
• Thermodynamic Piston
Select default Weibull Shape Parameter, and characteristic life parameter, for blockage and leak
Select Adjustment Multipliers for design, operation and maintenance, of Control Valve
Select Adjustment Multipliers for design, operation and maintenance, of Mechanical Pump
Select Adjustment Multipliers for design, operation and maintenance, of Steam Trap
Calculate Adjusted Probability of failure for Control Valve
Calculate Adjusted Probability of failure for Mechanical Pump
Calculate Adjusted Probability of failure for each individual Steam Trap for both blockage and leak failure modes
Calculate Combined Probability of Failure
• Based on Steam Line configuration
Select steam using equipment
• Steam Turbine
• Heat Exchanger
• Tracing – Instrumentation
• Tracing – Relief Valve
• Steam Line
• Condensation Line (recovery)
• Distillation Column
• Flare
Select default Weibull Shape Parameter, and characteristic life parameter, for main equipment
Select Adjustment Multipliers for design, operation and maintenance, of steam using equipment
Calculate Probability of failure for the main steam using equipment (due to steam related failures)
Calculate final Probability of Failure for steam using system
Calculate Consequence of Failure for both leakage and blockage scenario
Calculate Current Risk and Future Risk as the function of time
Determine inspection date or any other mitigation actions based on risk criteria

Figure 7.1 – Overview of POF Calculation Framework for Steam Systems.

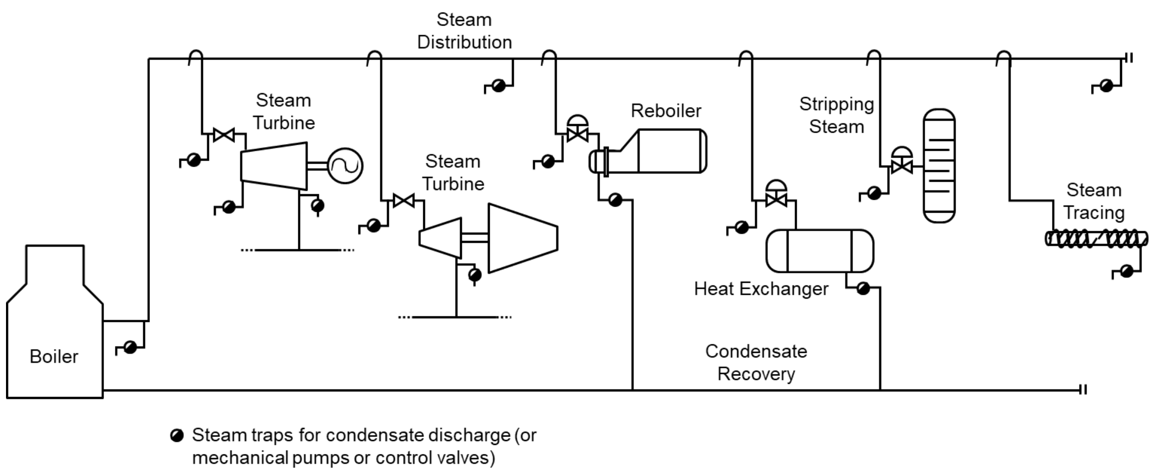


Figure 7.2 – A typical steam system containing steam traps (or mechanical pumps or control valves), steam lines and associated equipment.

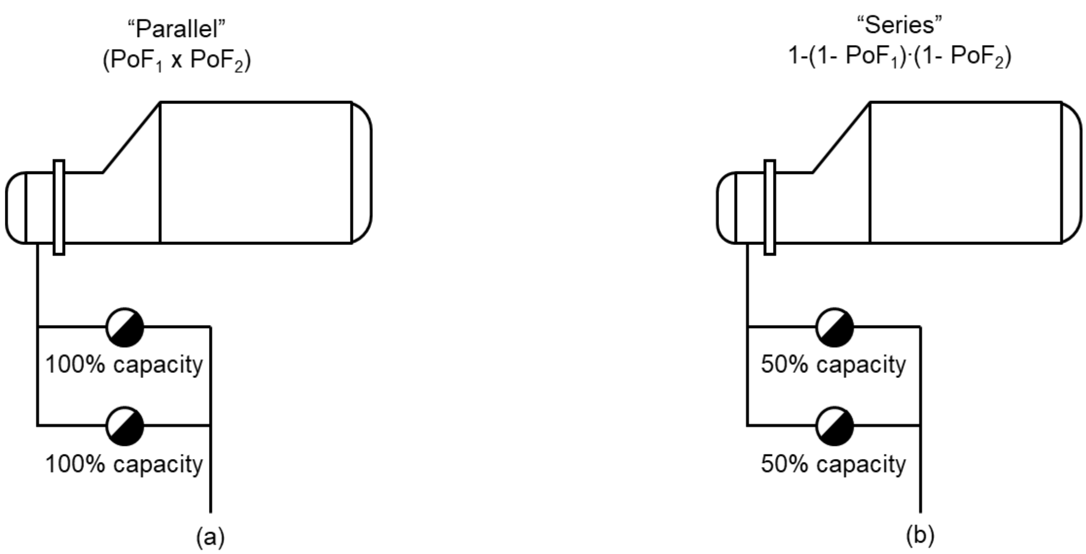
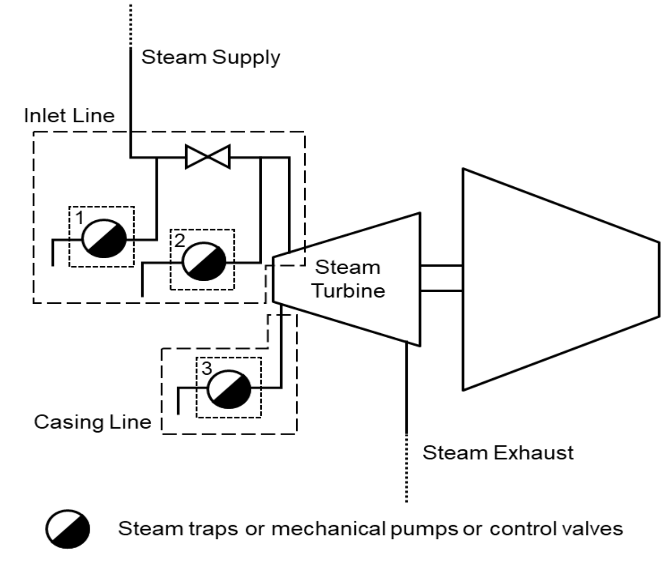
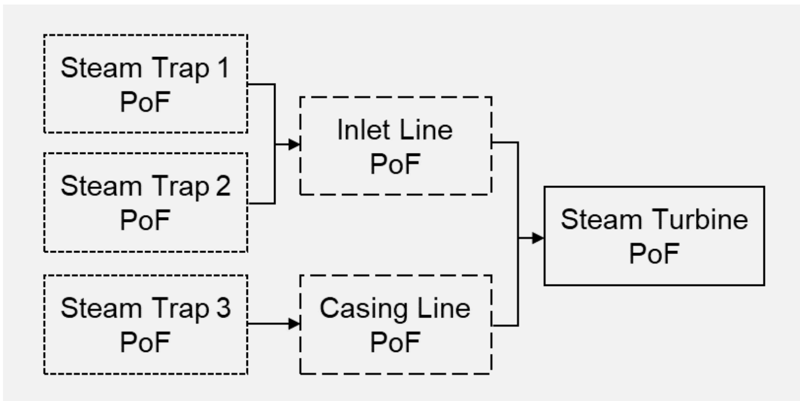


Figure 7.3 – Sample Configuration of Multiple Steam Traps (or mechanical pumps or control valves.



(a) Configuration of a steam turbine with steam traps or mechanical pumps or control valves.



(b) Block diagram for the calculation of POF for steam turbine with steam traps or mechanical pumps or control valves.

Figure 7.4 – Sample configuration of a steam turbine with steam traps or mechanical pumps or control valves.