THE IMPLEMENTATION OF RESIDUAL RISK ANALYSIS FOR EXPLOSION PROTECTION SYSTEMS

by

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ABSTRACT

For industrial explosion protection, residual risk analysis determines the likelihood that a given protection scheme will fail to mitigate an explosion occurrence, where one or more points of a system are subject to failure. Current design practice for providing explosion protection measures for industrial hazards follows a process where, although the designer satisfies accepted industry codes and standards, the result is a system where the risk of failure remains unknown. This thesis proposes and demonstrates the use of a methodology to assist design engineers in constructing an explosion protection system that meets a specified quantifiable level of risk. This new methodology can assist building owners and decision makers in selecting a design that best meets their risk-based goals and objectives.

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This work, the capstone to my academic career, is dedicated to my family.

EXECUTIVE SUMMARY

The problem of unmitigated explosions from inadequate explosion protection poses serious threats to the operation of industrial processes, to the personnel who work with and around the processes, and to the entire surrounding community. The hypothesis of this thesis is that residual risk analysis can be applied at the design stage to minimize unwanted explosions and their associated consequences, The amount of residual risk in a specific design of a protection scenario can then be both quantified and discussed by all the stakeholders. This know level of risk thus serves as a tool useful for making investment and design decisions based on improvements in an explosive hazard or process' risk position.

In exploring the solution to the industry problem, this thesis:

- Documents the current procedure for explosion protection system design which satisfies the minimum governing requirements;
- Introduces the residual risk analysis work of Date et, al. (2009), as a quantitative calculation tool for the mitigation of an explosion occurrence;
- Considers the relationship between the calculated residual risk value and Safety Integrity Levels for analyzing explosion protection system designs against appropriately benchmarked levels of risk ;
- Proposes an updated design methodology for explosion protection which utilizes residual risk analysis, safety integrity levels, and system optimization;
- Demonstrates the use of the proposed methodology to present the owner with a quantified residual risk associated with a particular design; and
- Upgrades the discussions from "meets code" to "probability of failure on demand" and involves all the stakeholders in an interactive discussion of risk.

This work bridges much of the gap between the theory of residual risk analysis and the practical implementation of this theory for real world applications. Future work is needed in the areas of information gathering; public policy, advancements to the theoretical model and quidance for the practicing engineer before widespread use of the methodology becomes practical for the design engineer and before the methodology can be incorporated into the relevant codes and standards.

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1 Introduction

Problem Statement

Unmitigated or ineffectually mitigated explosions pose serious threats to the operation of industrial processes, to the personnel who work with and around the process, and to the surrounding community. Design engineers rely on the applicable codes and standards as well as governing explosion protection principles to successfully mitigate an explosion. However, designing a system that satisfies code does not necessarily mean that the risk associated with the system is appropriate for its application. The standards used to design explosion protection systems and the governing principles which every system must maintain, do not quantify risk. This thesis proposes a new process for providing an explosion protection system which delivers key stakeholders the necessary information to make decisions based on a reduction in the probability of failure.

Governing Principles for Successful Explosion Protection

The aim of explosion protection systems is to mitigate explosions. For explosion protection to be successful, the following information and provided inequalities must hold true for all systems. If any one of these parameters is violated, then the system is no longer adequate. The inequalities shown are derived from the overall layout of the process, which includes the materials moving through the process, as well as the probability of an explosion in each vessel (plant item).

Explosion Pressure Reduction

Explosion pressure reduction is delivered by using passive explosion venting or by designing an active explosion suppression system. Every vessel has a given maximum pressure shock resistance, P_S, at which a maximum pressure may be exerted onto the walls and its structural integrity will remain intact. The maximum explosion pressure, P_{MAX}, that can be exerted on a vessel is mostly independent of the volume of the vessel and is primarily determined by the type of fuel and its state while in the vessel (Date, et. al., 2009). In simplistic terms, if the maximum pressure, P_{MAX}, is greater than the vessel strength P_S, there will be a failure of the vessel. When providing explosion protection on vessels, the component's function is to relieve or inhibit the pressure increase below the threshold of P_S.

When using explosion venting for explosion protection mitigation, it is crucial that the vents are sized such that the reduced pressure, P_{RED}, is less than P_S. This will ensure that the explosion does not compromise the vessel's structural integrity. When using an explosion suppression to provide a reduced pressure, the system will discharge a chemical suppressant during the incipient stages of the explosion to mitigate and inhibit an explosion. Similar to the venting, explosion suppression systems deliver a reduced pressure, P_{RED}, that is less than the vessels ultimate pressure shock resistance, P_S. Thus the governing equation for explosion pressure reduction is as follows (Date, et. al., 2009):

$$P_{\mathrm{Re}d} \leq P_S$$

Equation 1: Explosion Pressure Reduction Governing Inequality

Barrier Establishment

Where two or more vessels are connected with ducting, an explosion event will produce a flame front that will propagate through the connection and ignite the second vessel. An active or passive barrier is established within the interconnecting ducting, which protects the downstream vessel from an explosion. Passive barriers, such as air-lock rotary valves, serve as an explosion isolation barrier without a control panel or detection circuit. Active systems, which include a chemical suppression discharge or mechanical valve, require detectors and control panels to establish an interconnection barrier. The time it takes for the establishment of these barriers can be described as T_B (Moore and Lade, 2009). One of the most crucial features in explosion isolation is the determination of the amount of time it takes for the flame front to reach the isolation barrier. The time of arrival, T_D, is a complex parameter which incorporates explosion intensity, ignition location, process flow velocity and direction, explosion detection, and explosion duration (Moore and Lade, 2009). For active systems, the barrier must be established before the time it takes for the flame front to propagate to the barrier location. Thus governing equation for barrier is establishment is as follows (Date, et. al., 2009):

$$T_B \leq T_D$$

Equation 2: Barrier Establishment Governing Inequality

Equipment Failure

The most fundamental assumption when designing a system is that the components will all function properly and operate when called upon. However, no piece of equipment is completely reliable. If a major component of fire protection equipment (i.e. suppression agent release, control panel, detectors) fails, the system will not operate as desired (Date, et. al., 2009). The reliability of a particular explosion protection component is a function of hardware specific Mean Time Between Failures (MTBF), which are used to determine its probability of failure at any given event. The more reliable the piece of protection equipment is, the less likely the component will fail to operate when it is called upon.

Current Practice - Designing to Code

The National Fire Protection Association (NFPA) codes and standards prescribe the appropriate methods to address known explosion hazards in industrial processes. For each type of vessel and for each type of industrial process, there are multiple designs which satisfy the acceptable codes and standards. NFPA 68, *Standard on Explosion Protection by Deflagration Venting,* and NFPA 69, *Standard on Explosion Prevention Systems,* provide the minimum protection of vessels that are considered explosive hazards. NFPA 68's purpose is to provide the user with "the criteria for venting deflagrations in vessels", which can minimize the destructive effects of a deflagration (NFPA 68, 2007). It is used in conjunction with NFPA 69, which purpose is to provide the "minimum requirements for installing systems for the prevention of explosions in enclosures that contain flammable concentrations of flammable gases, vapors, mists, dusts, or hybrid mixtures."(NFPA 69, 2008) NFPA 654, *Standard for the Prevention of Fire and Dust Explosions from the Manufacturing Processing, and Handling of Combustible Particulate Solids*, determines the design objectives for explosion protection systems, and sets the conditions on when to use either NFPA 68 or NFPA 69 when calculating for a particular process arrangement (NFPA 654, 2006).

Together, the NFPA standards provide the minimum level of explosion protection and mitigation. In addition to the minimum code requirements, manufacturers have listed and approved software to calculate certain characteristics for successful explosion mitigation and protection which are either used to meet the approved requirements specific to the manufacturer (Fenwal Protection Systems, 2009). The fundamentals of the software are based on the information provided from NFPA 68, 69 and other published technical papers. This software is developed through theory and approved by actual testing by agencies such as FM Global and any other ATEX notified body (Fenwal Protection Systems, 2009).

When utilizing the applicable NFPA standards, it is possible to offer multiple acceptable designs which provide satisfactory explosion protection. However, a methodology to compare and quantitatively determine the system which will provide the highest level of protection or which system carries the highest inherent risk in its design does not currently exist. A cost benefit analysis is beneficial, but the process owner can not determine the associate risk of one code-complying system over another.

Inherent in the current design methodology is the seldom discussed fact that even for a code compliant design, the end user will assume a certain level of (unknown) risk to the system.. There is a need to quantify this remaining (residual) risk of failure to the system and to understand what further risk reduction is necessary or desirable. In utilizing residual risk analysis for explosion protection systems, one is able to quantify the risk of failure of all proposed systems, which allows the process owner to select a system based on a measurable benefit rather than by solely qualitative means.

Residual Risk Analysis

For all industrial process involving explosive hazards, there will always be some non-zero level of assumed risk, as it is impossible to safeguard any process 100%. By providing safeguards (preventative and protective) for explosive hazards, much of the risk is mitigated, and the remaining risk that the system inherently carries is considered the residual risk. As used in this thesis, residual risk is the probability or likelihood that a given protection scheme will fail to mitigate an explosion occurrence, where one or more points of a system are subject to this failure (Date, et. al., 2009).

For explosion protection systems, the residual risk is a function of five major factors:

- the layout of the process being protected;
- the Mean Time Between Failures (MTBF) of protection hardware;
- the reduced pressure in relation to the plant strength;
- the time to barrier establishment in relation to the propagating flame jet; and
- the probability of an explosion

As the residual risk moves lower, the probability of a failure becomes smaller. The challenge is to design and balance a code-satisfying protection system that carries residual risk levels acceptable to the process owner.

Hypothesis

Residual risk analysis is a tool that can be used by process owners and design engineers to make investment and design decisions based on improvements in an explosive hazard or process' risk position. Understanding the current process used to provide explosion protection systems including all the stakeholders in place, the milestones from project launch to completion, and the tools used to specify designs to meet current codes and standards will lead to a new methodology for providing and explosion protection system.

2 Existing Process for Providing Explosion Protection

Provided in this section is a graphical representation of the current process for providing an explosion protection system for any given plant. The process has been determined through synthesizing the information given in all the code language. It utilizes a three phase approach, which consists of an assessment phase, design phase, and acceptance phase. Each phase is delineated by the milestones performed in each section. The assessment phase consists of gathering information and providing a risk assessment. The design consists of calculating and providing a code satisfying explosion protection system. Finally, the acceptance phase is made up of stakeholders determining the appropriateness of the design proposals.

The stakeholders involved in this current process include the following:

- Third party consultants, (which can include hired insurance consultants;
- Design engineers or manufacturers;
- Process owners; and
- Authority Having Jurisdiction (AHJ)

These stakeholders are responsible for utilizing different tools to achieve each milestone outlined in each phase of the current process. The tools include NFPA Standards, Center for Chemical Process Safety guidelines, insurance regulations, manufacturer-specific design standards, and other tools explained below. To understand the process in its entirety, all portions of the process map will be explained on a phase by phase basis.

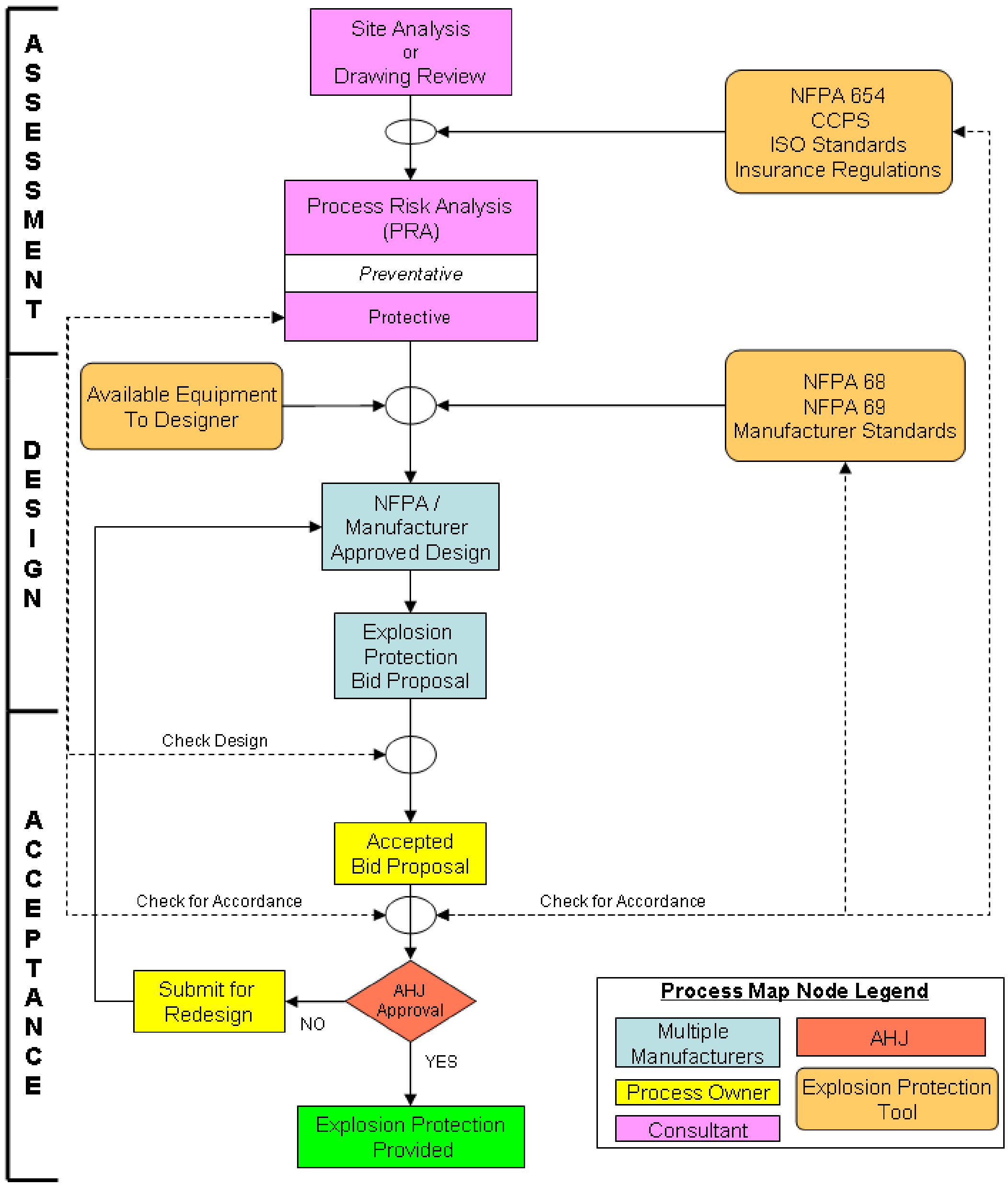

Figure 1: Current Process for Providing an Explosion Protection System

Assessment Phase

The Assessment Phase consists of two major milestones, which are to be explained below:

1) Site Analysis or Drawing Review
2) Process Risk Analysis

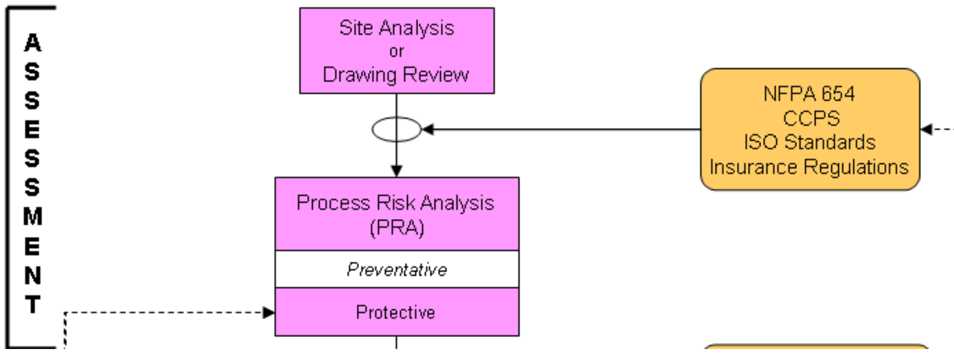

Figure 2: Current Process Assessment Phase

The first step in providing an explosion protection system is for a consultant, insurance provider, or equivalent third-party reviewer to perform an engineering review of process to understand the geometry and physical characteristics of the plant. Determining items such as types of vessels, geometry and physical characteristics of the vessels, interconnections between vessels, products of explosivity, and other layout features are crucial to developing the next milestone. After conducting the scope of the process and the equipment that actually presents the hazard, it is necessary to provide a risk assessment or analysis that would directly determine the proper means for protecting the process equipment, which is known as the Process Risk Analysis.

In order to complete Process Risk Analysis (PRA), the consultant must utilize several explosion protection tools, which are inserted into the process map along the process flow. To develop a PRA in the United States, NFPA 654: *Standard for the Prevention of Fire and Dust Explosions from the Manufacturing, Processing, and Handling of Combustible Particulate Solids* will typically be the governing document. NFPA 654 also suggests in Annex A.7.1.1 that the AIChE *Center for Chemical Process Safety (CCPS)* can be utilized to determine the hazard protection measures. If insurance providers are establishing the PRA, they may have their own standards to develop the appropriate protection measures for the explosive hazards. Additionally, there are ISO standards and other means to determine the protection measures to reduce the risk that involve Hazard Operability Studies (HAZOP), historical data, and fault tree analyses. The PRA typically considers two major sections: preventative measures and protective. The focus of this thesis is to provide an analysis of the protection measures needed to reduce the overall risk to a level acceptable to the process owner. Therefore, any preventative measures to provide explosion protection will not be considered in this study. The PRA provides the acceptable explosion protective measures, which the owner will then bid to manufacturers to provide a system design.

Design Phase

The Design Phase consists of two major milestones, which are to be explained below:

1) NFPA/Manufacturer Approved Design
2) Explosion Protection Bid Proposal

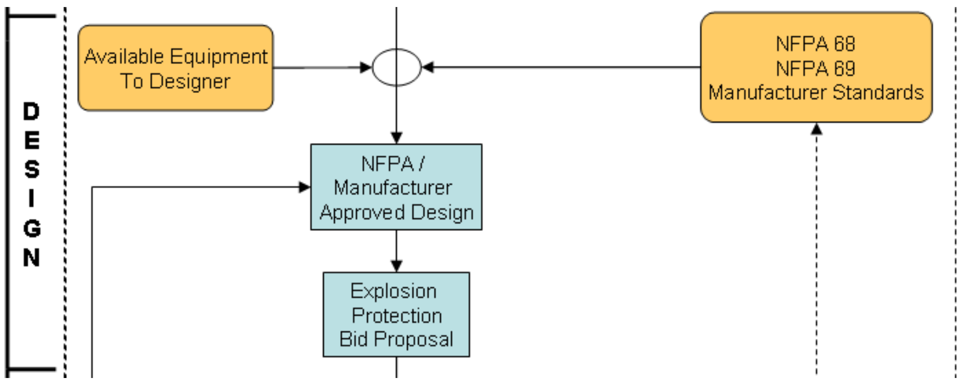

Figure 3: Current Process Design Phase

Manufacturers and design engineers will use the PRA to provide an NFPA or manufacturer accepted explosion protection design. NFPA 68, NFPA 69, as well as manufacturer design standards are the governing documents to build a satisfying protection scheme. Design engineers and manufactures calculate the proper venting, suppression, isolation, and any other methods of protection that are recommended in the PRA, with the purpose of satisfying the minimum requirements of the governing documents. Since the designer has a limited set of equipment to select from, it his/her responsibility to determine which of the available equipment fulfills the parameters of the minimum required design. Once these basic calculation parameters are met (i.e. the total venting area and suppression agent required as governed by the minimum requirements of the standards) an overall design scheme that satisfies the minimum requirements of both NFPA standards and the manufacturers design standards is established.

There is an associated cost and reliability with each protection component selected. Under the current process, it is the responsibility of the manufacturer or design engineer to use cost optimization and engineering judgment to provide a competitive bid that satisfies the minimum requirements of the governing standards. Once there is a level of comfort to satisfying the code at a reasonable cost (as determined by the designer), a bid proposal will be submitted to the process owner for selection. Once the design work is completed and the bid submitted, the Acceptance Phase begins.

Acceptance Phase

The acceptance phase consists of two major milestones, which are to be explained below:

1) Accepted Bid Proposal
2) Authority Having Jurisdiction (AHJ) Approval

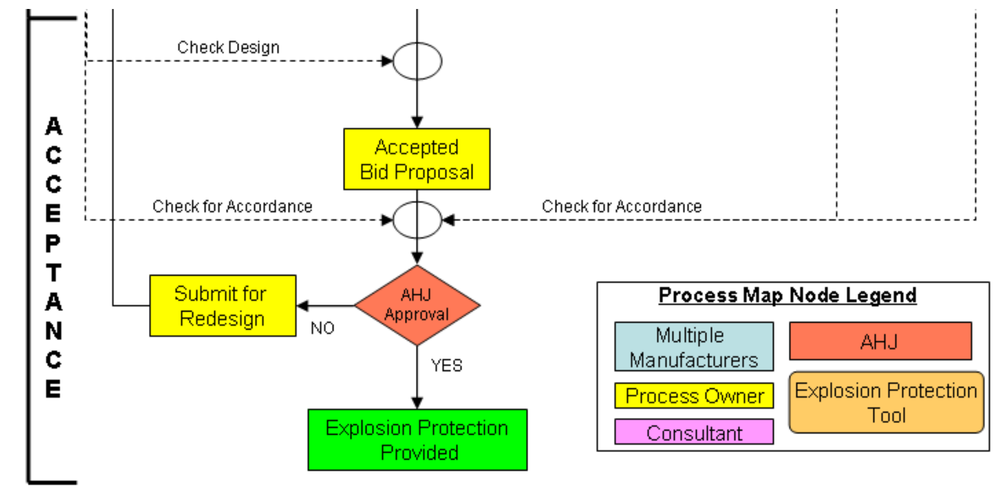

Figure 4: Current Process Acceptance Phase

During the acceptance phase, it is the process owner's responsibility to select one of the bid designs that satisfies all the requirements set forth by the assessment phase at a cost that is agreeable. If there are multiple bids being offered as design solutions from multiple designers or manufacturers, it is possible that all designs are determined appropriate by both the PRA guidance and the minimum requirements of the accepted standards. Therefore it is the process owner's responsibility to use a selecting process (i.e. cost-benefit) to determine which of the explosion protection bids are best suited for the needs of the hazard and project.

Once the bid is selected by the process owner, an Authority Having Jurisdiction (AHJ) will then check the design for its accordance with three major explosion protection tools: the PRA, the tools used to derive the PRA, and the standards and design guides deemed acceptable for governing explosion protection in that jurisdiction. If the AHJ deems the design meets the minimum required design parameters, then the protection system will be provided to the end user. However, if the AHJ deems that the system does not meet the minimum required design, or requires additional levels of protection, the process owner will then resubmit the design to the manufacturer for corrections.

Existing Process Analysis

The current process is based on information pieced together from all the governing documents and from interviews with experts in the field. It provides accepted explosion protection systems for various kinds of explosive hazards; however, no process is without its weakness. The current process can not and does not provide quantifiable benefit of increased safety. It is clear that the owner has the ability to decipher the costs between multiple manufacturers' bids; however, it is difficult to assess whether or not the process owner can assess the benefit on a system by system basis. The current analysis of an explosion protection design's acceptability relies on the determination that it satisfies the PRA and governing standards. Moreover, the process allows for a system that satisfies the minimum requirements to carry an unacceptable level of residual risk to the process owner or for its environment.

The true benefit of a design gets lost because the design is checked against the accepted benchmark standards. While society accepts this procedure as good measure, it fails to provide a pathway for optimizing a design to ensure that design carries risk levels aligned with the end user's acceptability level. Because of this, the design engineer is able to design an explosion protection scheme that satisfies all components of the approved standards, while not being entirely risk appropriate for the process owner or the environment in which the process operates.

Utilizing residual risk analysis can serve as a quantitative assessment tool for the process owner to determine which system both meets the needs of the risk analysis and the budget. Residual risk analysis benchmarked against Safety Integrity Levels (SIL) may provide design engineers the guidance for developing a system that meets an acceptable risk threshold as per the determination of a process owner or insurance regulation. Additionally, matching SIL and residual risk can serve as a quantitative assessment tool for the design engineers to optimize a design to meet the needs of the risk assumed and the budget of the process owner.

3 Residual Risk Analysis for Explosion Protection Systems

Date et al.(2009), have developed a model that quantifies the total residual risk for explosion protection systems. The model is a function of the five major factors that drive residual risk, and is based on a set of equations which work in conjunction with directed graph representation. A synopsis of the model and mathematics, as well as a worked example, is presented below.

Directed Graph Representation

Industrial explosion hazards often involve multiple vessels and multiple connection paths which will become means of flame propagation. Date et al (2009), approach the concept of a connected system using directed graph representation, which is outlined below. The directed graphical representation considers the layout of the system and how each individual vessel affects the entire system in an synergistic approach. In the representational design, each vessel in the system is defined as a vertex. In the event of an ignition, the edges between the vertices represent possible paths of flame propagation such as ducting. Between any pair of adjacent vertices, there are two directed edges in opposite directions which show the likelihood of flow at any event. Each edge is associated with a certain weighting which represents the directional probability of flame propagating along the connection.

The best way to understand directed graph representation is through an example. Date et al (2009) provide a synopsis of a spray drying process and how it would be represented in their model, which is restated below. Figure 5, reflects the process prior to being represented mathematically. In this process, a product is spray dried, and then passes through two fluid bed driers that further reduce the moisture content of the final product. Dust content in the drying air is separated by a ganged pair of cyclones, and returned via a fine particulate return line to the spray drier.

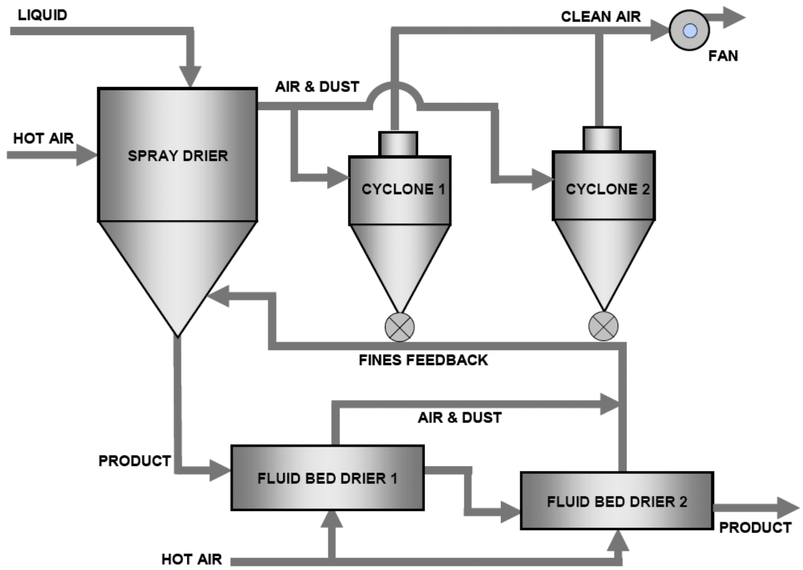

Figure 5: Example Process for Explosion Protection

Figure 6 shows the corresponding directed graph representation for the above process (Date, et. al., 2009). Vertices are just abstract representations of the plant vessels, and each vertex has multiple edges and the result is a much simpler representation of the system.

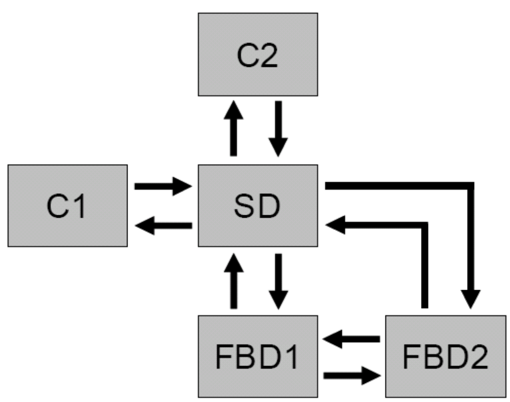

Figure 6: Directed Graphical Representation

Spray Dryer	SD
Cyclone 1	C1
Cyclone 2	C2
Fluid Bed Dryer 1	FBD1
Fluid Bed Dryer 2	FBD2

Table 1: Directed Graph Representation Nomenclature

Mathematical Theory

Using directed graph theory, Date et al.(2009), developed a series of equations to quantify risk based on the above direct graph representation, the governing equations for protection, and the assumed MTBF for all equipments. Below is a synopsis of the directed graph equations and what they represent. A calculated example and analysis of each step is provided in the next section.

Each vessel or plant item and associated explosion protection system, *i* (vertex *i*), within the process plant is characterized by a set of parameters. The ignition probability of vertex *i* is characterized by $Q_E(i)$, and that for a given process plant and over a given time that $\sum Q_E(i) = 1$ (Date, et. al., 2009). This means that there will be one ignition occurrence somewhere in the process plant. The risk of failure of any vertex *i* from ignition in vertex *j*, is denoted by $R(i, j)$ and can be calculated as the summation of the risk of hardware failure, $Q_h(i)$, and the risk of inadequate explosion protection. The equation is as follows (Date, et. al., 2009):

$$R(i, j) = Q_h(i) + (1 - Q_h(i))(Q_{vessel}(i, j))$$

Equation 3: Risk of Failure in Vertex i due to Ignition in j

$Q_{Vessel}(i, j)$ represents the how close the reduced pressure, P_{RED}, is to the vessel strength P_S, which includes and accounts for any design safety factors for the computation of the residual risk (Date, et. al., 2009). $Q_h(i)$ can be calculated from the following equation (Date, et. al., 2009):

$$Q_h(i) = \alpha(i) + (1 - \alpha(i))\beta(i) + (1 - \alpha(i))(1 - \beta(i))(\gamma(i))$$

Equation 4: Probability of Explosion Protection Hardware Failure on Vertex i

The items in Equation 4 for $Q_h(i)$ represent the hardware failure of venting panels, α, explosion detectors, β, and explosion suppressors, γ that may be installed as part of an explosion protection system. $Q_h(i)$ essentially represents the probability that an unmitigated explosion occurs in vessel *i* due to hardware failure (Date, et. al., 2009).

The risk of failure of any vertex in the system due to an ignition in vertex *i* will be denoted as $\delta(i)$ and determined by the following equation (Date, et. al., 2009):

$$\delta(i) = Q_E(i)\left(R(i,j) + (1 - R(i,j))\sum_{j \in \phi_i} Q^S(i,j)R(i,j) \right)$$

Equation 5: Per Ignition Residual Risk

Each $R(i, j)$ is calculated using Equation 3, which can be substituted into Equation 5. ϕ_i represents the set of vertices adjacent to the initial vertex in question, vertex *i*, and $Q^s(i, j)$ represents the total flame propagation probability from vessel *i* to *j* (Date, et. al., 2009). The flame propagation is dependent on the geometric configuration and the explosion hazard itself, as well as the reliability of any explosion isolation hardware such as a mechanical or chemical barrier (Date, et. al., 2009). The initial term in Equation 5 represents the event where an ignition in vertex *i* causes an unmitigated explosion in that vessel. The second term with the sum over *j* represents an event where there is no unmitigated explosion in vertex *i*, but flame propagation to vertex *j* which causes an unmitigated explosion in vertex *j* (Date, et. al., 2009). Essentially, Equation 5 gives a calculation for the residual risk on a per-ignition basis, due to the ignition in vessel *i*.

A different way to calculate the risk is to approach it on a per-vertex approach, where the total risk of failure in each vertex is due to ignition in the same vertex or in any of the connecting vertices. This risk is denoted as ζ_i, and is given in the following equation (Date, et. al., 2009):

$$\zeta(i) = Q_E(i)R(i,j) + \sum_{i \in \phi_j} Q_E(j)(1 - R(j,j))Q^S(j,i)R(i,j)$$

Equation 6: Per Vertex Residual Risk

The total or overall residual risk, *R*, can be calculated by Equation 7, where there is a summation over all ζ_i (Date, et. al., 2009). One may achieve the same results of Residual risk by using a summation over $\delta(i)$ with an inclusion of the probability of failure of the suppression system control panel (Date, et. al., 2009).

$$R = \sum \left\{ \pi(j) + (1 - \pi(j)) \sum_{i \in \Psi_j} \zeta_i \right\}$$

Equation 7: Total Residual Risk

The probability of failure of the suppression control panel is denoted by π. The summation is over all control panel zones and $\Psi_1, \Psi_2 \ldots$ represents the MTBF for each of the given control zones (Date, et. al., 2009).

Calculated Example

To get an understanding of how the residual risk calculations works, a simple hand calculated example is provided. The example represents a two-vessel system connected with ducting, which is protected only by passive venting. It is assumed that only one vent is required for each vessel, and the given dimensions are provided on the drawing below. Additionally, there are given parameters that are needed to successfully calculate the residual risk for this system. Below is the nomenclature table, which outlines each variable.

Nomenclature

$Q_E(i)$	Ignition probability in vessel i. For this example we are assuming 100% probability
$P_{red}(i,j)$	Reduced explosion pressure in vertex i following an ignition in vertex j.
$P_s(i)$	Pressure shock resistance of vertex i.
$Q_{vessel}(i,j)$	Probability that the explosion protection hardware does not fail, but the reduced explosion pressure is still higher than the pressure shock resistance of the vessel:
$\alpha(i)$	Probability of any vent panel failure
$\beta(i)$	Probability of any detector failure
$\gamma(i)$	Probability of any suppressor failure
$R(i,j)$	Risk of failure of any vertex i due to ignition in vertex j
$t_b(i,j)$	Time from ignition for the isolation barrier to be established between vessels i and j
$t_f(i,j)$	Time taken for the flame front to arrive at the barrier location between vessels i and j
$Q_{barrier}(i,j)$	Probability that the isolation hardware is actuated and the barrier established, but the barrier is deployed too late and flame passes from vessel i to j
$Q_f^s(i,j)$	Probability of flame propagation between connected vessels i and j which then leads to an enhanced explosion in j. For this example we are assuming 100% probability in each direction.
$Q^s(i,j)$	Total flame propagation probability from vessel i to j which then leads to an enhanced explosion in j, taking into account any explosion isolation provision
$Q^h(i,j)$	Probability of explosion isolation hardware failure between vessels i and j
$Q_h(i)$	Probability of explosion protection hardware failure on vessel i
δ_i	Residual risk of failure of any vertex due to an ignition in any vertex i
ζ_i	Residual risk of failure of vessel i due to an ignition in the same vessel or any vessel directly connected
Φ_j	The set of vertices adjacent to vertex i.
V_1	Source vessel where ignition occurs
V_2	Vessel connected to V_1
π_3	Reciprocal Mean Time Between Failures (MTBF) value for the vent panels chosen in this example

Table 2: Residual Risk Nomenclature

Givens:

$P_{S1} = 0.50bar$

$P_{red1} = 0.49bar$

$P_{S2} = 0.25bar$

$P_{red2} = 0.24bar$

$Q_{E1} = 1$

$Q_{E2} = 1$

$\pi_3 = 2\times10^{-5}$

$Q_f^s(1,2) = 1.0$

$Q_f^s(2,1) = 1.0$

Vent Panel Quantity per Vessel = 1
No Detection
No Suppression
No Isolation

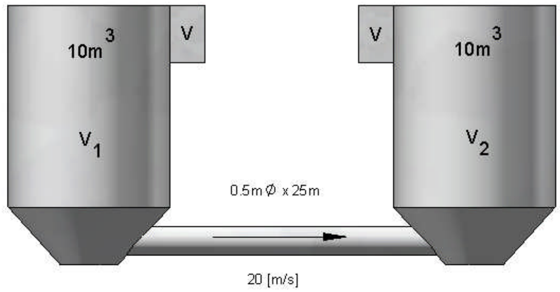

Figure 7: Calculated Example Process Layout

Determine Q_h: *Probability of explosion hardware failure on vessel i*

In this set of calculations, we want to determine the overall probability that any piece of hardware that is used to mitigate an explosion (i.e. venting in this example) will fail to do so. In scenarios where there is more than one piece of equipment such as venting on one vessel, suppression on another with some isolation barrier in between, the overall probability is a function of the respective MTBF's using a statistical approach.

$$Q_h(i) = \alpha(i) + (1-\alpha(i))\beta(i) + (1-\alpha(i))(1-\beta(i))(\gamma(i))$$

$$\alpha(i) = \alpha_1(i) + (1-\alpha_1(i))(\alpha_2(i))$$

$$\alpha_n(i) = \sum_{j=0}^{K_n-1} \pi_n(i)(1-\pi_n(i))^j$$

$$\gamma(i) = \gamma_1(i) + (1-\gamma_1(i))(\gamma_2(i))$$

$$\gamma_n(i) = \sum_{j=0}^{K_n-1} \pi_n(i)(1-\pi_n(i))^j$$

$$\beta(i) = \sum_{j=0}^{K_n-1} \pi_n(i)(1-\pi_n(i))^j$$

Solved Example

$$\alpha_1(i) = \sum_{j=0}^{1-1} 2\times10^{-5}\left(1-2\times10^{-5}\right)^0 = 2\times10^{-5}$$

$\alpha_2(i) = 0$ [Only 1 vent type]

$$\alpha(i) = 2\times10^{-5} + \left(1-2\times10^{-5}\right)(0) = 2\times10^{-5}$$

$$\gamma_n(i) = \sum_{j=0}^{K_n-1} \pi_n(i)(1-\pi_n(i))^j = 0 \text{ [No Suppression]}$$

$$\beta(i) = \sum_{j=0}^{K_n-1} \pi_n(i)(1-\pi_n(i))^j = 0 \text{ [No Detection]}$$

$$Q_h(1) = Q_h(2) = 2\times10^{-5} + \left(1-2\times10^{-5}\right)0 + \left(1-2\times10^{-5}\right)(1-0)(0) = 2\times10^{-5}$$

For this calculated example, since there is only one type of hardware on the entire system, we can see that the overall probability is purely a function of its reciprocal MTBF value.

Determine Q_{vessel}(i,j): *Probability that the explosion protection hardware does not fail but where $P_{red}>P_s$*

In this step the calculation being performed here is a statistical analysis of the likely distribution that the hardware works, but does not have enough margin of safety to properly maintain the reduced pressure below the ultimate plant strength. For this example, we are calculating to see what the likelihood that the vent panels rupture, but do not relieve the pressure enough to maintain the plant integrity. The calculation is a function of its surroundings, where the downstream vessels have an impact on the upstream vessels, and vice versa.

$Q_{vessel}(i,j) = P(P_{red}(i,j) - P_s(i) > 0)$ which is equal to $Q_{vessel}(i,j) = 1 - Z$

$$StdevP_{red}(i) = 10\%(P_{red}(i))$$
$$NormP_{red}(i,j) = 5\%(P_{red}(i)) + (P_{red}(i))$$
$$MP_{red}(i) = NormP_{red}(i) - 2(StdevP_{red}(i))$$

$$StdevP_s(i) = 10\%(P_s(i))$$
$$NormP_s = 5\%(P_s(i)) + (P_s(i))$$
$$MP_s(i) = NormP_s(i) - 2(StdevP_s(i))$$

$$M(\mathrm{P_{red}}(i) - \mathrm{P_s}(i)) = \mathrm{P_{red}}(i) - \mathrm{P_s}(i)$$
$$Stdev(\mathrm{P_{red}}(i) - \mathrm{P_s}(i)) = \sqrt{(\mathrm{P_{red}}(i))^2 - (\mathrm{P_s}(i))^2}$$

Normal Distribution Z: $M(\mathrm{P_{red}}(i) - \mathrm{P_s}(i))$, $Stdev(\mathrm{P_{red}}(i) - \mathrm{P_s}(i))$

Solved Example

$$NormP_{red}(i) = 0.49$$

$$StdevP_{red}(i) = 10\%(0.31) = 0.049$$
$$MP_{red}(i) = 0.31 - 2(0.326) = 0.392$$

$$NormP_s = 0.50$$
$$StdevP_s(i) = 10\%(0.5) = .05$$
$$MP_s(i) = 0.50 - 2(0.05) = 0.60$$

$$M\left(\mathrm{P_{red}}(i) - \mathrm{P_s}(i)\right) = 0.392 - 0.60 = -0.208$$
$$Stdev\left(\mathrm{P_{red}}(i) - \mathrm{P_s}(i)\right) = \sqrt{(0.049)^2 - (0.05)^2} = 0.0700$$

Normal Distribution Z: -0.208, 0.0700 = .999

$$Q_{vessel}(1{,}1) = 1 - Z = 1.48 \times 10^{-3}$$
$$Q_{vessel}(1{,}2) = 1 - Z = 4.32 \times 10^{-3}$$
$$Q_{vessel}(2{,}1) = 1 - Z = 9.15 \times 10^{-4}$$
$$Q_{vessel}(2{,}2) = 1 - Z = 2.79 \times 10^{-3}$$

From the above calculations one may see that there is some variability between each vertex, as the calculation takes into consideration the components adjacent to the vessel as well as the vessel itself. For example, $Q_{vessel}(1,2)$ looks at vessel 1 with respect to vessel 2 and it shows that when considering both vessels, the overall likelihood that the equipment selected will not mitigate the explosion is $4.32\text{x}10^{-3}$. The biggest factor in reducing the likelihood of the reduced pressure exceeding the overall plant strength is the initial design delta between the values. If the designed reduced pressure was even smaller, then the likelihood or probability that the hardware would not achieve a pressure less than the plant shock resistance is lower. The closer we move the expected reduced pressure towards the plant shock resistance, the greater the probability that the hardware will not properly mitigate the explosion.

Determine R*(i,j): The risk of failure of any vertex i due to ignition in vertex j*

This calculation step is used to determine what the overall risk of failure is with respect to ignition in that vessel, or ignition in an adjacent vessel. The overall risk is calculated with the above likelihood of failures of both complete hardware failure, and failure to operate within the defined range. The equation is as follows:

$$R(i, j) = Q_h(i) + (1 - Q_h(i))(Q_{vessel}(i, j))$$

Solved Example

$R(1,1) = 2\times10^{-5} + (1-2\times10^{-5})(1.48\times10^{-3}) = 1.50\times10^{-3}$
$R(1,2) = 2\times10^{-5} + (1-2\times10^{-5})(4.32\times10^{-3}) = 4.34\times10^{-3}$
$R(2,1) = 2\times10^{-5} + (1-2\times10^{-5})(9.15\times10^{-4}) = 9.35\times10^{-4}$
$R(2,2) = 2\times10^{-5} + (1-2\times10^{-5})(2.79\times10^{-3}) = 2.81\times10^{-3}$

As one may see the values align mostly with the Q_{vessel} numbers because of the static hardware failure number. As one may see, although incredibly small, there is still a chance that even with an appropriately designed system, there is a chance for failure.

Determine Q^s: *Total flame propagation probability from vessel i to j*

This calculation looks at the probability of flame propagation between two vessels. It looks at the flame probability (which for this example is given as 1.0), the probability of failure of isolation hardware, and the probability of the barrier (chemical or mechanical) being established before the flame jet reaches the point of deployment.

$$Q^s = Q_f^s(i,j)\left(Q^h(i,j) + (1-Q^h(i,j))(Q_{Barrier}(i,j))\right)$$

$Q_f^s(i,j)$ = Flame probability; given

$Q^h(i,j) = \pi_3(i) + (1-\pi_3(i))(\pi_4(i,j))$ or $Q^h(i,j) = (\pi_4(i,j))$

$Q_{Barrier}(i,j) = P(t_b(i,j) - t_f(i,j) > 0)$

$\left(Q^h(i,j) + (1-Q^h(i,j))(Q_{Barrier}(i,j))\right)$ = Hardware Failure

Solved Example

$Q_f^s(1,2) = 1.0$

$Q_f^s(2,1) = 1.0$

$\left(Q^h(i,j) + (1-Q^h(i,j))(Q_{Barrier}(i,j))\right)$= 1; No Hardware means a theoretical total failure of the isolation equip.

$Q^s(1,2) = 1.0(1) = 1.0$
$Q^s(2,1) = 1.0(1) = 1.0$

Since there is no hardware for this particular example and because I have considered that the overall likelihood of flame propagation to be 100%, we can assume that the total flame propagation from vessel 1 to vessel 2 is 1.0 or 100%.

Determine δ(i): ***Residual risk of failure of any vertex due to an ignition in any vertex i***

This calculation takes all the information already derived from the above calculations and determines the residual risk of failure at any vessel due to ignition in any vessel in the entire system. We are assuming Q_E to be 1.0 or 100%, because the check is to see what will happen hen ignition occurs. If the likelihood of ignition is much less than 1.0 then the overall probability will move down accordingly. However, for determining system safety, it is necessary to look at the system when it is being used and to do this a value of 1.0 is necessary.

$$\delta(i) = Q_E(i)\left(R(i,j) + (1 - R(i,j))\sum_{j\in\phi_i} Q^S(i,j)R(i,j)\right)$$
$$\delta(1) = Q_E(1)\big(R(1,1) + (1 - R(1,1))Q^S(1,2)R(2,1)\big)$$
$$\delta(1) = (1)\left(1.50\times10^{-3} + \left(1 - 1.50\times10^{-3}\right)(1.0)\left(9.35\times10^{-4}\right)\right) = 2.44\times10^{-3}$$

$$\delta(2) = Q_E(2)\big(R(2,2) + (1 - R(2,2))Q^S(2,1)R(1,2)\big)$$
$$\delta(2) = (1)\left(2.81\times10^{-3} + \left(1 - 2.81\times10^{-3}\right)(1.0)\left(4.34\times10^{-3}\right)\right) = 7.13\times10^{-3}$$

Residual risk analysis looks at the system as a complete entity with separate parts; it will show that downstream vessels and non-attached vessels will have an impact have an impact on each and every vessel or vertex in that system. This example only has two vessels and they are attached, so the calculation is simple.

Determine ξ (i): ***Residual risk of failure of vessel i due to an ignition in the same or any connected vessel***

This calculation step determines the overall likelihood of failure when ignition occurs in that vessel or in any vessel that is attached to vessel being calculated.

$$\zeta(i) = Q_E(i)R(i,j) + \sum_{i\in\phi_j} Q_E(j)(1 - R(j,j))Q^S(j,i)R(i,j)$$
$$\zeta(1) = Q_E(1)R(1,1) + Q_E(2)(1 - R(2,2))Q^S(2,1)R(1,2)$$
$$\zeta(1) = (1)\left(1.5\times10^{-3}\right) + (1)\left(1 - 2.81\times10^{-3}\right)(1.0)\left(4.34\times10^{-3}\right) = 5.83\times10^{-3}$$

$$\zeta(2) = Q_E(2)R(2,2) + Q_E(1)(1 - R(1,1))Q^S(1,2)R(2,1)$$
$$\zeta(2) = (1)\left(2.81\times10^{-3}\right) + (1)\left(1 - 1.5\times10^{-3}\right)(1.0)\left(9.35\times10^{-4}\right) = 3.74\times10^{-3}$$

This example only has two interconnected vessels, so the probability calculations are straightforward and only one calculation is needed per vessel.

Determine: *Total Residual Risk*

The total risk is just a summation of the individual components.

$\sum \delta(i)$ or $\sum \zeta(i)$

Solved Total Residual Risk

$$\sum \delta(i) = 2.44 \times 10^{-3} + 7.13 \times 10^{-3} = 9.57 \times 10^{-3}$$
$$\sum \zeta(i) = 5.83 \times 10^{-3} + 3.74 \times 10^{-3} = 9.57 \times 10^{-3}$$

Total Residual risk = $9.57 \text{x} 10^{-3}$, or an expected system failure of 1-in-104 events

In Summary

This is an example used to display how the calculations are supposed to work. Since the calculations look at each component with respect to the entire system, fixing one trouble area is not as simple as just adding some components. Everything has an effect on downstream and upstream vertexes. Understanding how the calculations work is crucial to understanding how risk is actually changed; where changing components and protection strategies can greatly alter the overall residual risk to the system. In instances where the per-ignition risk is different than the per-vertex risk, the highest residual risk will govern. When a problem area arises, one must remember that both that individual vertex as well as all the attached vertexes has an important role in shaping the overall residual risk of the system.

(i,j)	Q_E(i,j)	Q_h(i,j)	Q_{vessel}(i,j)	R(i,j)	Q^s_f(i,j)	Hardware Failure	Q^s(i,j)
(1,1)	1	2.00E-05	1.48E-03	1.50E-03	-	-	-
(1,2)	-	-	4.32E-03	4.34E-03	1.0	1.0	0.8
(2,1)	-	-	9.15E-04	9.35E-04	1.0	1.0	0.6
(2,2)	1	2.00E-05	2.79E-03	2.81E-03	-	-	-

Table 3: Example Interim Calculation Summary

Ignition	$\delta(1)$	2.44E-03	
	$\delta(2)$	7.13E-03	*One-in*
	Residual Risk	9.57E-03	104

Table 4: Example Per Ignition Residual risk

Vertex	$\zeta(1)$	5.83E-03	
	$\zeta(2)$	3.74E-03	*One-in*
	Residual Risk	9.57E-03	104

Table 5: Example Per Vertex Residual risk

Residual Risk Overview and Analysis

Currently there is no widely available tool for the calculation of residual risk. A user could perform the hand calculations, which may probe cumbersome on particularly complicated designs. Conversely, the equations can be programmed or developed a computer calculation program to quickly calculate more complicated system designs. From the calculated example above, it is determined that the residual risk of failure for this particular explosion protection design is 9.57 x 10^{-3}, which translates to 99.04% system availability. With the use of residual risk analysis presented by Date et al. (2009), design engineers and process owners now have the ability to quantify the level of residual risk with any particular explosion protection system design.

With these calculations, one must consider that residual risk analysis only calculates the risk associated with a design; it does not determine if this risk is acceptable or appropriate for its environment. For the example above, there is no way to determine if a residual risk of failure is 9.57 x 10^{-3} is appropriate for the system. Environmental considerations such as the location of the process and the consequence of a system failure are not considered when calculating residual risk. To understand the complete picture of residual risk it is important to utilize benchmarked standards to assess the appropriateness of a design. Coordinating residual risk with Safety Integrity Levels may provide the process owners and design engineers the necessary information for making investment decisions based on risk with respect to its environment or the process owners risk reduction goals.

4 Safety Integrity Levels

Assigning a Safety Integrity Level

The Safety Integrity Level (SIL) presents the Probability of Failure on Demand (PFD) and the amount of risk reduction that is necessary to mitigate risk associated with a process to an acceptable level in qualitative and quantitative terms (Summers, 1998). The origin of SIL comes from the International Electrotechnical Commission Standards IEC-61508 and IEC-61511 and is defined as the likelihood of a safety-related system satisfactorily performing the required safety functions under all the stated conditions, within a stated period of time (IEC 61508, 1998). A discussion of these IEC standards and the background to SIL can be seen in Appendix 1.

To assign a benchmark or SIL for a particular hazard, the consequence of failure needs to be understood. Quantitative and qualitative inputs are used to determine the impact of a failure on property, employee safety and the surrounding community. The design of the Safety Instrumented Systems (SIS) will be developed to reduce the risk to the required SIL (Summers, 1998). Understanding the consequence of failure as well as the frequency of occurrence will determine the required availability and associated PFD of the system. In Table 6, provided below, a qualitative assessment of the consequence of failure is used to determine the appropriate SIL for a particular hazard. Suppose a risk assessment determines that an unmitigated explosion for a particular explosive hazard results in some employee and community impact. The appropriate SIL classification is SIL-3 and the entire explosion protection scheme must conform to this benchmarked level. Typically a target SIL will be determined in conjunction with the process owner as a target level of tolerable risk.

SIL	Consequence of Failure
4	Catastrophic Community Impact
3	Employee and Community Impact
2	Major Property and Protection. Possibly Injury to employee
1	Minor Property and Production Protection

Table 6: Qualitative Determination of SIL

Table 7 provides the Probability of Failure on Demand with a benchmarked SIL as determined through a risk assessment detailed above. Corresponding to consequence of failure, there are associated probabilities of failure on demand for component parts.

SIL	Availability Required	Probability of Failure on Demand	Occurrence of Failure
4	>99.99%	10^{-5} to 10^{-4}	1-in-100,000 to 1-in-10,000
3	99.90 – 99.99%	10^{-4} to 10^{-3}	1-in-10,000 to 1-in-1,000
2	99.00 – 99.90%	10^{-3} to 10^{-2}	1-in-1,000 to 1-in-100
1	90.00 – 99.00%	10^{-2} to 10^{-1}	1-in-100 to 1-in-10

Table 7: Probability of Failure on Demand and SIL

For explosion protection systems, one is not considering an individual protection component's MTBF as sole qualification for achieving a benchmarked SIL. Rather, the entire system's risk reduction, as a function of the explosion protection system in its entirety, is used to evaluate system availability against the consequence of failure. In using Table 7, achieving SIL-3 for the example explosive hazard would require system availability between 99.90% and 99.99% and the associated probability of failure on demand between 10^{-3} and 10^{-4}. In more simplistic terms, the occurrence of failure can be expected between 1-in-1,000 and 1-in-10,000 events for which the system may be called upon to perform as designed.

Connecting Residual Risk to SIL

Under the current process for providing explosion protection, the process owner has the task of determining which of the bids for explosion protection are best suited for their respective process or hazard. Even with the introduction of residual risk analysis for explosion protection system design, it is difficult to benchmark what a process owner would consider an acceptable rating. To benchmark a residual risk result, it is important to understand that explosion protection systems are considered as a low-demand system, where the system is not continuously called upon to perform as designed (IEC 61511, 2004). With this, the low-demand probability of failure on demand is on the same order of magnitude as the calculated residual risk analysis. This allows direct correlation between the calculated residual risk and the system PFD for a particular SIL. This relationship gives a benchmarked qualitative and quantitative risk level associated with every system design when calculating the residual risk.

The current process to provide explosion protection does not allow for a seamless integration of residual risk analysis and SIL to be applied as calculation tools for quantitatively assessing an explosion protection system. From this, it is important to develop a new process in which these concepts can be incorporated to provide risk appropriate protection systems while maintaining the satisfying requirements of the existing code and standard structure.

5 New Process for Providing Explosion Protection

As outlined above, it is proposed that residual risk analysis is a feasible tool for process owners and design engineers to make investment and design decisions based on improvements in an explosive hazard or process' risk position. In utilizing residual risk analysis in combination with Safety Integrity Levels, it is possible to provide safer processes by determining the overall quantifiable risk associated with an explosion protection scheme. Provided in this section is a new process which integrates residual risk analysis, SIL, and decision nodes to optimize explosion protection schemes to appropriate risk levels while maintaining the minimum design requirements of prescriptive codes and standards. Establishing this framework is a key delivery of this thesis.

New Process Overview

The new process utilizes a three phase approach which includes an Assessment, Design, and Acceptance Phase. During the Assessment Phase, the primary function is to understand the process at hand and develop a protection strategy with risk reduction goals in mind. The Design Phase focuses on providing an acceptable explosion protection design by using prescriptive codes and the process analysis determined in the Assessment Phase. The Acceptance Phase checks the design against several factors to determine whether it meets the goals of the process owner and of the minimum governing standards.

The table below represents the primary stakeholders involved in the new process as well as the major explosion protection documents and tools used. Milestones that are highlighted magenta are performed by a consultant. Consultants can be made of insurance assessors, third party explosion consultants, explosion equipment manufacturers hired as a consultant, or a process owner if they are adept at providing design guidance. Milestones highlighted in blue are performed by design engineers or manufacturers, red is preformed by the Authority having Jurisdiction (AHJ), and yellow is preformed by the process owner. The final function component to this process map is explosion protection tools. The explosion protection tools, highlighted in orange, represent governing standards such as NFPA 654, equipment available to the design engineer, or other types of regulations. These blocks are essential explosion protection tools that are necessary to achieve the subsequent milestone in the process.

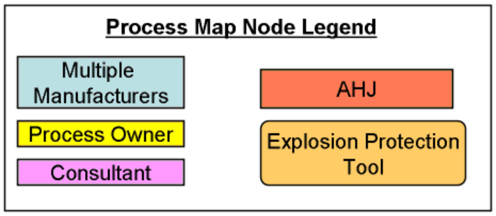

Figure 8: New Process Map Legend

There are several decision nodes in the new process for explosion protection. These decision nodes allow the stakeholder the opportunity to assess the process thus far, and determine whether or not to continue or to refine a certain aspect of the design. With the exception of the consultants, the remaining stakeholders have a decision at some point in the process, which allows for better communication among all involved parties. Additionally, there are footnotes in some of the steps which are explained in on the side of the diagram. These are used in an effort to simplify the process map to make it applicable for as many scenarios as possible.

Provided below in Figure 9 is the new process for providing explosion protection. To understand the process in its entirety, all milestones of the process map will be explained on a phase by phase basis.

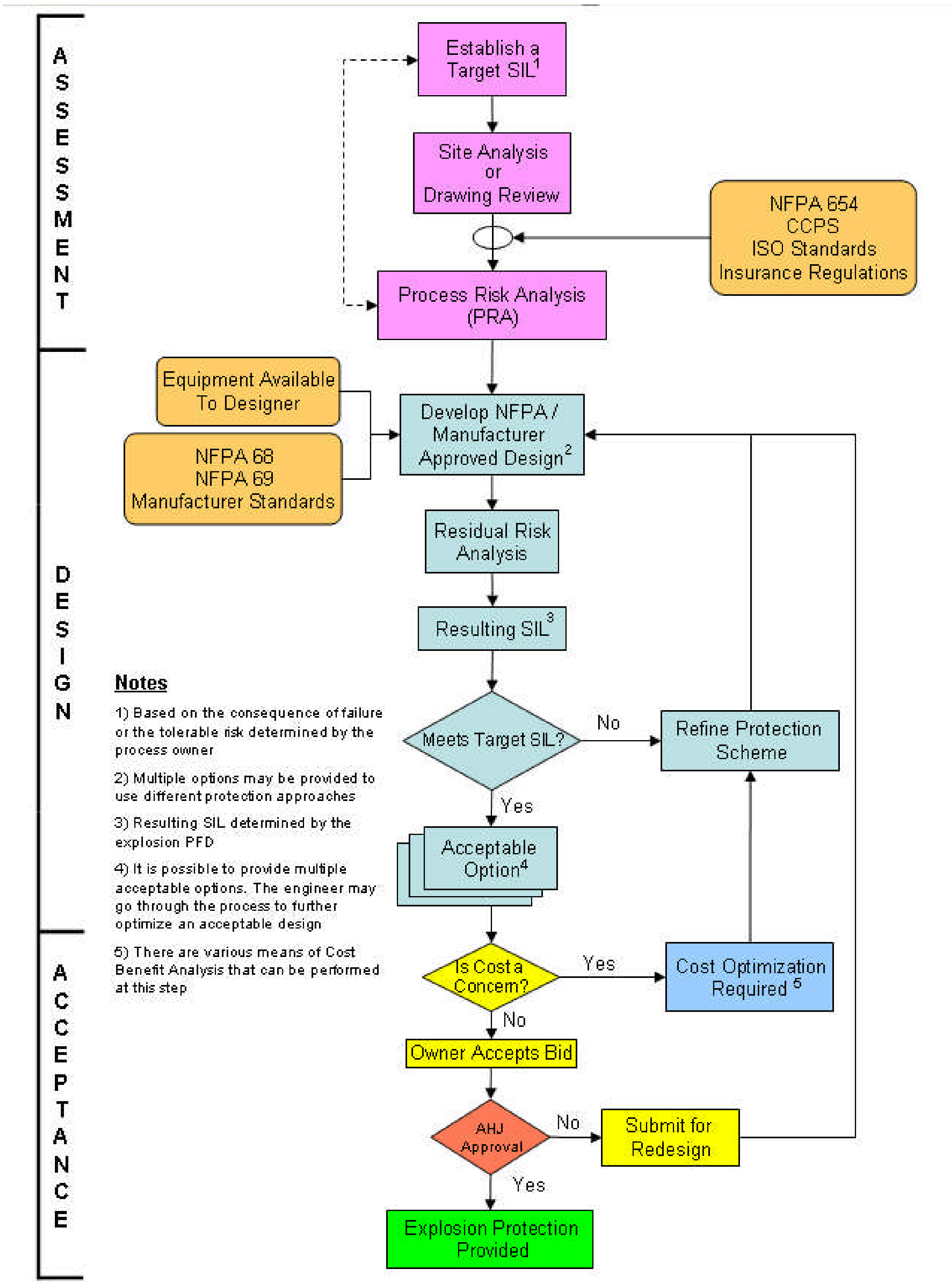

Figure 9: New Process for Providing Explosion Protection

Assessment Phase

The Assessment Phase consists of 3 major milestones, which are to be explained below:

1) Establishing an Appropriate System Safety Integrity Level (SIL)
2) Site Analysis or Drawing Review
3) Process Risk Analysis (PRA)

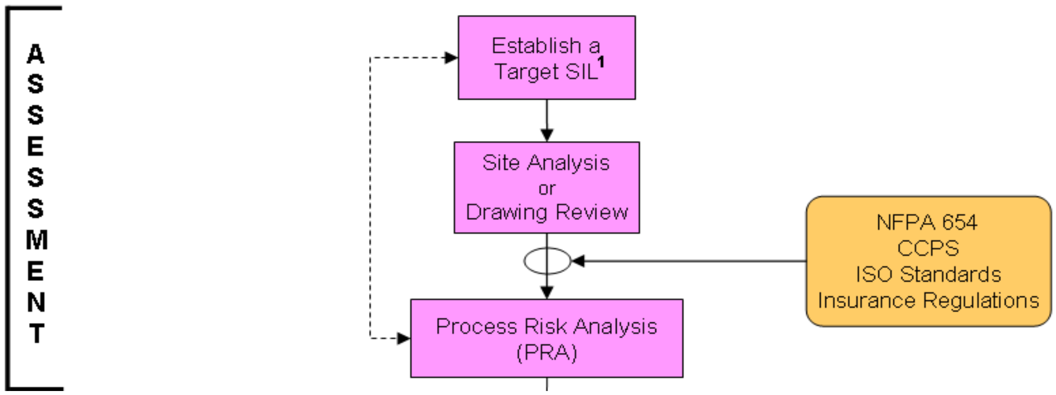

Figure 10: New Process Assessment Phase

The Assessment Phase starts with a consultant establishing an appropriate Safety Integrity Level (SIL) for the process or hazard being questioned. A target SIL is derived from the consequence of failure of the explosion protection system and the possible extent of damage to the process facility or beyond. This involves a thorough risk analysis of the process site and its environment, in conjunction with an understanding of the stakeholders' risk tolerance. The risk analysis may be performed through a Process Risk Analysis (PRA), which is explained in detail below. The established SIL for the process hazard is then implemented as part of the PRA, where the design phase will develop a protection strategy to satisfy this requirement.

Once the target SIL is established as one of the facility's overall risk goals, a site visit or engineering review of process documentation to understand the geometry and physical characteristics of the plant is performed. The information gathered in this step is crucial for setting the protection scheme as well as assessing the hazards with each process. Information gathered in this step includes but is not limited to:

- types and geometry of vessels in the process;
- location and proximity to hazards and structural members;

- size and length of interconnecting ducting;
- products of explosivity;
- fan speeds; and
- other important characteristics of the process

The information from the site analysis is used in combination with several explosion protection tools to develop the Process Risk Analysis (PRA), which provides the acceptable explosion protection measures for the process surveyed. In the United States, NFPA 654: *Standard for the Prevention of Fire and Dust Explosions from the Manufacturing, Processing, and Handling of Combustible Particulate Solids (ref)* will typically be the governing document. However, other documents such as the AIChE *Center for Chemical Process Safety (CCPS) Guidelines for Hazard Evaluations Procedures,* ISO Standards (ISO 6184), Hazard Operability Studies (HAZOP), historical data, and fault tree analyses can be used to develop the PRA. In the new process, establishing a SIL is also integrated in the PRA, to provide the minimum design criteria for mitigating residual risk to the appropriate SIL.

Design Phase

The Design Phase consists of 6 major milestones, which are to be explained below:

1) Develop NFPA/Manufacturer Approved Design
2) Residual Risk Analysis
3) Resulting SIL
4) Decision Node: Meet Target SIL?
5) Refine Protection Scheme
6) Acceptable Option(s)

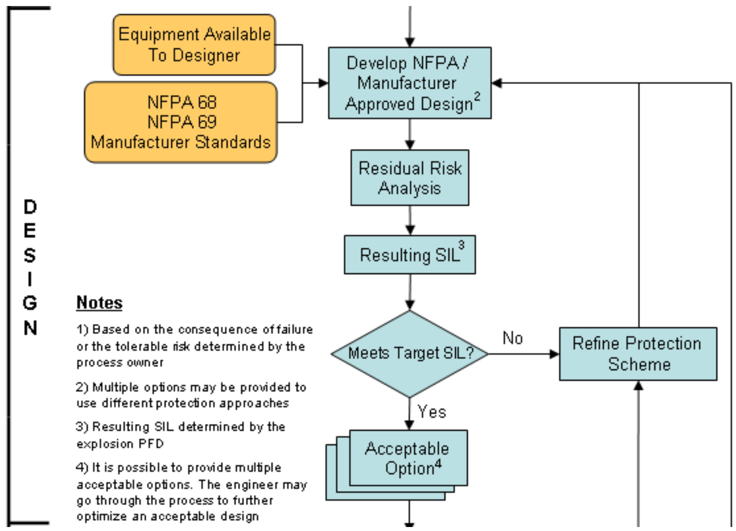

Figure 11: New Process Design Phase

Manufacturers and design engineers will use the PRA to provide an NFPA- or manufacturer-accepted explosion protection design. NFPA 68, NFPA 69, as well as manufacturer design standards are the governing documents to build a satisfying protection scheme. Design engineers and manufactures calculate the proper venting, suppression, isolation, and any other methods of protection that are recommended in the PRA, with the purpose of satisfying the minimum requirements of the governing documents. Since the designer has a range of equipment to select from, it his/her responsibility to determine which of the available equipment fulfills the parameters of the minimum required design. Once these basic calculation parameters are met, an overall design scheme that satisfies the minimum requirements of all applicable NFPA standards and the PRA is established. Although there may be multiple combinations of equipment which may result in an acceptable design, the design engineer needs to select a single baseline design for analysis. However, at this point, it is not possible to determine if the design achieves the SIL requirement established in the assessment phase, and residual risk analysis is needed for this assessment.

As noted above, Date et al. (2009) have developed a model that quantifies the total residual risk for explosion protection systems. In using the concepts and mathematics provided by this work, a calculated residual risk for each system design is achievable. The calculations will look at the per-vertex and per-ignition residual risk of failure to quantitatively determine the availability of the system design. The resulting residual risk level (availability of system) can then be correlated to the Probability of Failure on Demand (PFD) component of the SIL. At this point, the engineer now has a SIL related to PFD for the explosion protection system, which can be compared against the target SIL identified in the assessment phase (which is based on consequence of explosion system failure). This provides the design engineer with quantitative information to determine if the explosion protection system is adequately aligned with the process owner's risk tolerance target (the target SIL developed during the Assessment Phase).

The next milestone is a decision node by the design engineer to determine if the current design and the resulting SIL meet the target SIL. If the system satisfies all the benchmarked approved standards and meets the required SIL, the design is considered an acceptable option. If the process does not meet the target SIL, the design engineer will refine or optimize the design by modifying the baseline design within the constraints of the NFPA standards, the PRA, and the equipment available. This iterative process will continue until an acceptable option is produced, which satisfies the minimum requirements of the risk assessment, governing standards, and process owner's risk reduction objectives.

A key aspect of this new process is that it is possible to demonstrate that multiple acceptable options exist, which allows for system optimization on factors such as residual risk/SIL or cost. For instance, the designer may want to further reduce the overall risk to achieve a safer design than is prescribed by the target SIL. With quantitative information about each design, the manufacturer and designer engineers have the ability to optimize the design to provide multiple acceptable designs that meet or exceed the minimum requirements of the governing standards, that appropriate SIL rating and PRA, and that are cost sensitive. Once the design engineer is satisfied with the acceptable designs calculated, the process will move to the Acceptance Phase.

Acceptance Phase

The Acceptance Phase consists of four major milestones, which are to be explained below:

1) Decision Node: Is Cost a Concern
2) Accepts Bid Proposal Option
3) Cost Optimization Required

4) AHJ Approval

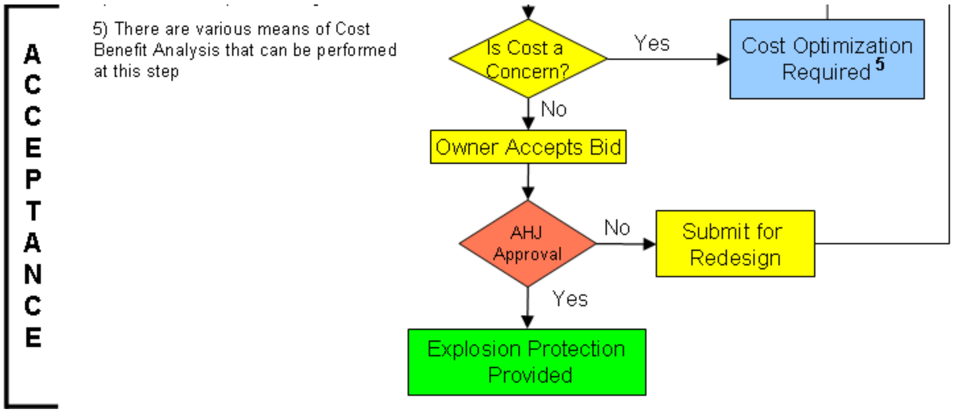

Figure 12: New Process Acceptance Phase

The Acceptance Phase starts with the process owner making a decision on cost. The acceptable option being offered from the Design Phase may cost more than the process owner is willing to accept, even though it may satisfies all the requirements of the Assessment Phase. If the cost is a concern for the process owner, the decision node allows for the process to be submitted back to the design engineer for cost optimization. There are many ways to perform cost-benefit analyses; however, the focus of this thesis is to provide a framework that gives the design engineer necessary information quantitative the benefit against the cost from an objective standpoint. If the process owner determines that cost optimization is required, the design engineer will return to the Design Phase and refine the design and calculate the residual risk to ensure it satisfies all the benchmarked requirements discussed above. If the cost of the acceptable protection scheme is not a concern, the process owner will accept the option as-is and continue to the AHJ approval milestone

Once the bid is selected by the process owner, an Authority Having Jurisdiction (AHJ) will then check the design for its accordance with three major explosion protection tools: the PRA, the tools used to derive the PRA, and the standards and design guides deemed acceptable for governing explosion protection in that jurisdiction. If the AHJ deems the design meets the minimum required design parameters, then the protection system will be provided to the end user. However, if the AHJ deems that the system does not meet the minimum required design, or requires additional levels of protection, the process owner will then resubmit the design to the manufacturer for corrections.

New Process Map Analysis

The new process for providing explosion protection presented here gives design engineers or manufacturers the ability to quantitatively analyze and optimize their explosion protection designs to give the most competitive explosion protection bid proposals. Additionally, the map yields relevant data which allows the process owner to quantitatively assess the risk and cost of each proposal. The new process allows residual risk analysis to exist as an iterative process. While the explosion protection scheme might change based on subsequent revisions of the design, the manufacturer or design engineer never abandons satisfying the minimum design requirements per the governing standards and the PRA. Modifications within the allowable limits, as determined by the prescriptive codes, will produce a safer process design or a more desirable cost as dictated by the process owner. The new process delivers explosion protection designs aligned with the risk the process owners are willing to accept at the cost they are willing to pay.

In addition to providing quantitative assessments, the new process provides stakeholder agreement in all phases of the design. While not specifically responsible for certain milestones, the process owner’s objectives for an explosion protection system are clear. By establishing a tolerable risk at the onset of the process, the design engineers can provide designs that not only meet the minimum requirements, but can do so without fear of losing safety by cutting costs. The decision loops allow for stakeholder agreement as an iterative process, rather than providing a best guess design that satisfies the minimum standards.

Current versus New Process

The hypothesis of this thesis is that residual risk analysis is a feasible tool for process owners and design engineers to make investment and design decisions based on improvements in a hazard or process’ risk position. Residual risk has the ability to determine the total system availability by alone; however, the challenge is to move the tool from academia to a real-world process flow that can be implemented across all explosion protection designs. The new process for providing explosion protection is a take-home deliverable of this work.

The goal was to position residual risk analysis into a newly established process for providing explosion protection which successfully establishes information for owners and engineers to quantitatively approach a cost-benefit analysis. Establishing the connection between residual risk analysis and benchmarked SILs provides qualitative and quantitative meaning to the availability of an explosion protection system, and goes beyond simply satisfying the minimum code requirements. Within the new process, simple feedback cycles and decision nodes make an iterative design approach based on calculated information; however, everything on paper does not necessarily work in the real world. Demonstrating the methodology with a case study confirms that the new process to provide explosion protection provides the information necessary to optimize designs while staying within the boundary conditions of minimum requirements of the governing documents, the PRA, and any assigned SIL.

Conversely, the current process for providing explosion protection systems does not require an understanding of the associated residual risk levels with a given code satisfying design. As long as the design meets the minimum requirements of the governing documents, then the process is considered appropriate and fit for purpose. With the introduction of residual risk analysis, it is proved that not all satisfying designs can be treated as equal, and that there may be risk levels that the process owner is not willing to accept.

The new process for providing explosion protection, which uses residual risk analysis benchmarked against a defined SIL, identifies the major difference between the current process for providing explosion protection to what could be used going forward. With the ability to create multiple code-compliant designs per the minimum NFPA requirements, the new process separates designs that are engineered with to provide appropriate levels of risk to those that simply satisfy the baseline requirements for design. The new process gives the owner the ability to make investment decisions on a quantitative decision making approach rather than qualitatively. The qualitative assessments under the current process may yield unknown risk levels that the owner is unwilling to accept. Moreover, designing against the minimum code requirements does not always provide the appropriate levels of risk reduction for the process hazard as determined through a risk assessment. By disclosing this information in the new process, not only are we providing process owners with a system they are more comfortable choosing, but we are providing protection systems that have risk levels appropriate for the environment in which it operates.

The following section demonstrates the methodology of this new process map by designing and optimizing an explosion protection scheme for a case-study example.

6 Demonstrating the Methodology through a Case Study

Introduction

The framework of the new process for providing explosion protection provides the necessary information for design engineers and manufacturers to design to a target Safety Integrity Level while maintaining the design within the boundary conditions of a defined Process Risk Assessment, remaining above the minimum requirements of the governing NFPA documents, and utilizing a specified and limited set of protection equipment. In this section, a case-study process is established and the example will progress through structure of the new process to provide an acceptable option. The case study will not assume what a process owner is willing to accept for the given process, so the testing for this project will focus on the Assessment and Design Phase of the new process.

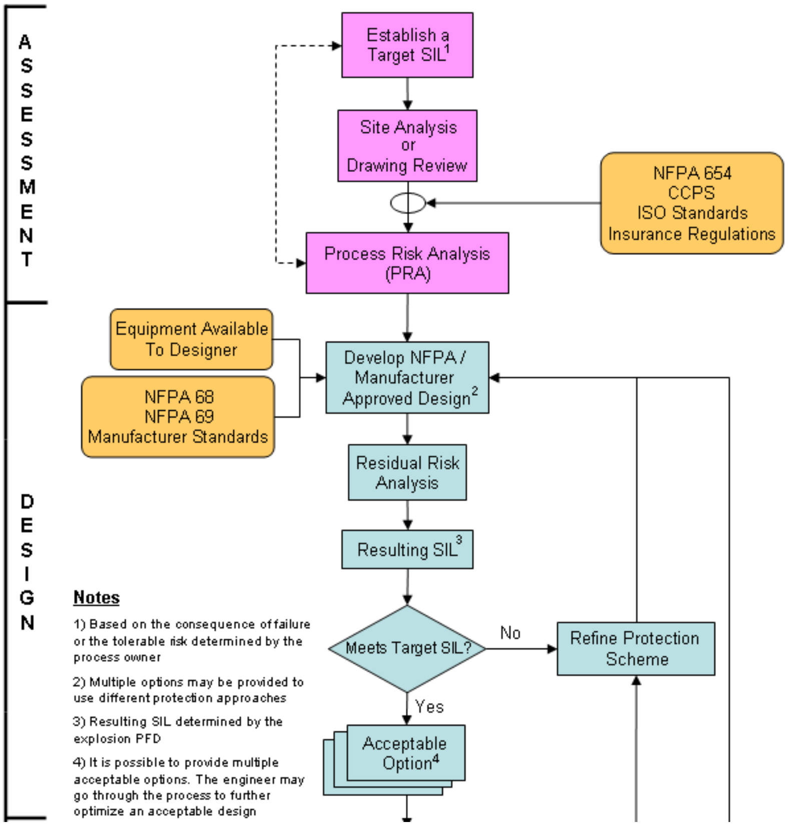

Figure 13: New Process Demonstrated

As the designer, the challenge is to develop a design that meets the governing NFPA documents for the case study process hazard and optimize the residual risk using a specified set of available equipment. To demonstration the new process, a sample explosive hazard is provided, where a baseline design that satisfies the governing NFPA standards for providing explosion protection is calculated using the above process, and is optimized to achieve the target SIL.

Assessment Phase

To demonstrate the methodology through a case study, a system is provided that is both appropriate for testing and representative of real-world processes. The process was modified from an actual explosion protection application, and can be considered a common industrial process.

Safety Integrity Level

As mentioned in the assumptions and limitations section, demonstrating the process to determine a SIL for the case study process is not provided in this document. However, the design SIL required for this system is established to be SIL-2.

Site Analysis or Drawing Review

The first step in the new process is to determine the overall process hazards through an actual walk-through site analysis or through a drawing review. The walk-through will gather information about the overall plant layout as well as the products being moved through the process. Essential to this process is the geometric layout of the process and any crucial components contained within. Provided in this section are the product of explosivity and the hazards presented with the case study process hazard.

Product of Explosivity

Understanding the fuel type and characteristics is a key component in designing an explosion protection design. The fuel for the system will consist of beech wood, which will be processed from large pieces into specific smaller sized pellets. By obtaining the smaller sized wood pellets, the process creates a substantial amount of explosive dust. The characteristics and explosive properties of the dust byproduct come from the BIA-Report Combustion and Explosion Characteristics of Dusts. Below is the table of the explosive characteristics of the material selected (Beck, et. al, 1997):

Material	Mat. No	Median Particle Size [µm]	Moisture Content [% by wt]	Lower Explosivity Limit [g/m^3]	Max Explos. Over-pressure [bar-g]	K_{St} Value	Minimum Ignition Energy [mJ]	Ignition Temperature [ºC]	Glowing Temperature [ºC]
Wood, Beech (flour)	3410	70	11	60	8.0	128	>10	400	320

Table 8: Physical Properties of Material for Testing

Process Overview

The physical process is a four vessel operation which can be considered a single zone. The single zone aspect of the protection means that if one of the system's components initiates, all components of the system will subsequently initiate. Systems the size of the system will sometimes incorporate a zoned approach, where a logic-based control panel will initiate only certain components of the protection scheme (NFPA, 2002). A single zone approach is both appropriate in a real-world application, and provides an easy-to-follow design approach when demonstrating the new process for providing explosion protection.

There are four components to the system; the hammermill, cyclone, silo, and baghouse. All of these separate components are connected by a series of ducts of a particular diameter and length. The hammermill is a pulverizing vessel which takes the wood product in large form and grinds it into much smaller pieces. The $10m^3$ vessel has a single outlet duct of 0.5m diameter, and is able to maintain structural integrity up to 2.0-bar gauge pressure. This vessel's process involves a substantial amount of energy as well as many moving objects within the vessel. Its two main byproducts are the smaller wood pellets and the highly explosive wood dust at the specified size in the table above. Both byproducts are then transported via the 0.5m diameter by 10m duct to a $9m^3$ cyclone, which has the ability to maintain structural integrity up to 0.5-bar gauge pressure. The cyclone then separates the larger pieces from the dust particles, which are then distributed to the dedicated independent storage vessels. The heavier wood pellets are transported via a 0.5m diameter by 6.5m duct directly into a $110m^3$ silo for storage, which can maintain structural integrity up to 0.4-bar gauge pressure. Finer pieces are distributed to the $50m^3$ baghouse through a 0.5m diameter by 8m duct, which primarily holds the highly explosive wood dust. The baghouse, which maintains structural integrity up to 0.2-bar gauge pressure, separates the larger pieces that are within the range of the particular desired size and distributes it back into the silo through a $0.12m^2$ diameter by 8m duct for bulk storage. The transportation of the particle is driven by an air circulation unit. This unit, while crucial to the operation of the process, is not considered to be part of the hazard due to the fact that the explosive dusts will never enter the machine. Overall it is important to note that the process is not 100% efficient, and one may expect that there will be mixing from the cyclone to the dedicated storage units. The diagram below shows the general overview of the process in graphical terms.

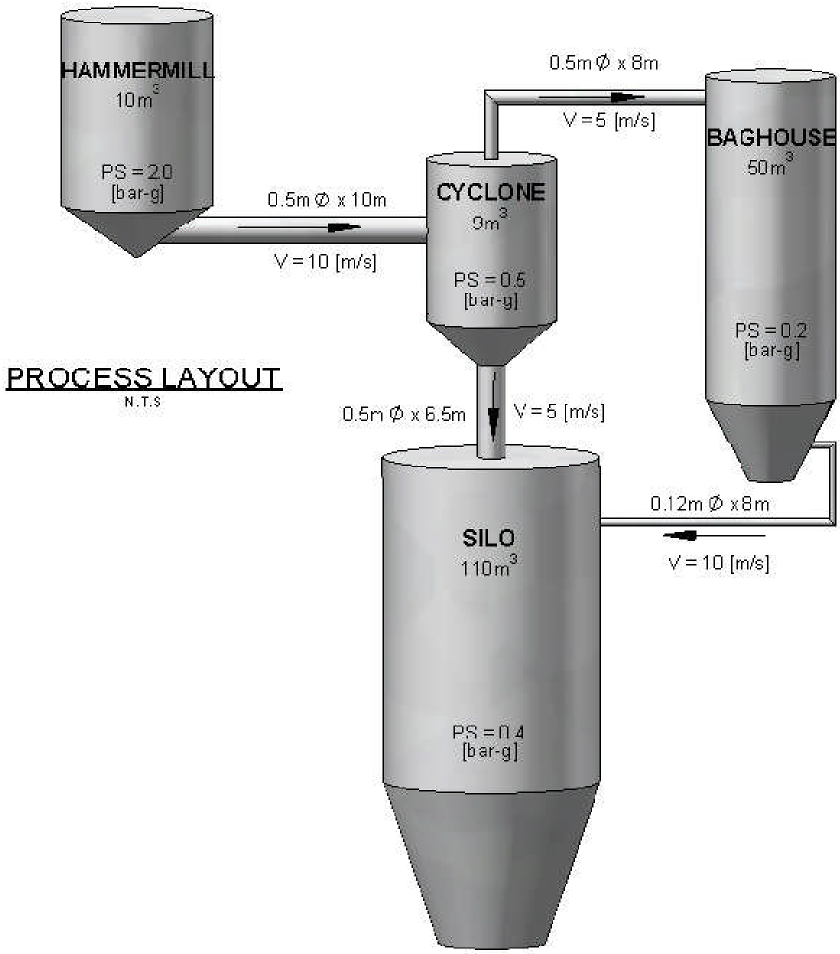

Figure 14: Case Study Process

Below is the description of the process in table format, which gives the important characteristics of the vessels in the explosion protection scheme. To correctly design an explosion protection system, the total volume, strength, and orifice openings need to be outlined.

Vessel Name	Volume [m^3]	Total Strength [bar (g)]	Orifice Openings	Duct Orifice Dia [m]
Hammermill	10	2.0	1	0.5
Cyclone	9	0.5	3	0.5
				0.5
				0.5
Silo	110	0.4	2	0.5
				0.12
Baghouse	50	0.2	2	0.5
				0.12

Table 9: Physical Dimensions of Theoretical System Vertexes

The table below outlines the connecting ducting throughout the whole system. The entire system is driven by air conveying system as mentioned in the process overview section. Understanding the connections between and among vessels is important in deciding the overall protection scheme (Moore and Lade, 2009). The table below gives the connecting vessels' duct diameter, length, and velocity of air movement.

Duct	Connecting Vessel 1	Connecting Vessel 2	Diameter [m]	Length [m]	Air Velocity [m/s]
1	Hammermill	Cyclone	0.5	10	20
2	Cylcone	Silo	0.5	6.5	5
3	Cylcone	Baghouse	0.25	8	20
4	Baghouse	Silo	0.12	8	20

Table 10: Physical Dimensions of System Connections

Process Risk Analysis (PRA)

Using NFPA 654 along with the AIChE's *Center for Chemical Process Safety Guidelines*, appropriate protective measures are determined. Provided below are the typical hazard posed by each plant item and the appropriate methods to protect the types of vessels provided in the case study hazard. Other inputs that may go into a process risk analysis include a Hazard Operability Study (HAZOP) or fault tree analysis, which would itemize protective and preventative measures to develop an overall risk mitigation strategy. The assessment phase will be performed by a third party organization or group to determine the overall risks to the process. For purposes of demonstrating the new process, a general guidance is provided below; a real system process risk analysis may be much more detailed and include more specific guidance for protection.

Hammmermill

(from 5.3.17 Size Reduction Equipment) (CCPS, 2005)
A hammermill effectively reduces the size of the input materials to a smaller more manufacturing-appropriate size. "Size reduction equipment must always be regarded as providing ignition sources because of the presence of friction and hot surfaces arising from the energy used in the comminution process." (CCPS, 2005)

Most hammermills can be protected by vents; however, providing vents on the actual hammermill is not typically done, but rather vents are installed on adjacent vessels receiving the milled product However, a very commonly used protective measure against explosive hammermills is to provide suppression (CCPS, 2005). When doing so, the mill must be designed "for the expected overpressure of generally 0.5 to 1.0 bar(g) in the event of a suppressed explosion." (CCPS, 2005)

Due to the nature of a hammermill's operation, venting is not normally feasible as a protection strategy due to the mechanical working of the vessel. Vent panels are calibrated rupture panels that may be sensitive to the hammermill's product being mechanically worked to a smaller size. Additionally, a hammermill may carry the highest probability of explosion due to its nature, so an active chemical explosion suppressant is an appropriate and more commonly used choice.

Baghouse

(from 5.3.4.3 Fabric Filters) (CCPS, 2005)
The main explosive hazard for equipment of this kind is an electrostatic spark that may discharge within the system. "Dust explosions occur quite frequently in baghouses because the likelihood of an easily ignitable fine dust atmosphere is high and there is high turbulence, which can cause electrostatic charge accumulation on the dust particles." (CCPS, 2005) Another possible source of ignition is the entrance of hot, glowing, particles into the baghouse from an upstream process.

"Fabric filters can be protected from fires and explosions by venting, suppression, or containment." (CCPS, 2005) A major challenge with explosion venting results from dislodged fabric bags causing a blockage of the vent area, which will reduce the effectiveness of the protection scheme (CCPS, 2005).

As mentioned above, venting a baghouse will sometimes prove to be problematic due to the fabric blocking a portion of the vent. However, this can be overcome by sizing your vent near the top of the vessel where the bags are not hanging. Additionally, since this vessel is a thin walled vessel which carries low plant strength, the addition of a chemical suppressant may provide too much pressure on the walls. Venting is likely the best option for this vessel.

Storage Silo

(from 5.3.15 Silos and Hoppers) (CCPS, 2005)
"[I]t is usually necessary to provide protective measures, which can be any of the following: venting, suppression, containment, inerting, and fire protection."

Venting is both the most widely used method and possibly the most economical method of protection. When a silo is located outdoors the vented explosion can be discharged directly to the atmosphere; if indoors, it must be ducted to the atmosphere. Explosion venting design is simplified, as typical low-pressure systems should utilize roof vents for proper explosion protection. Suppression may be utilized, but the designer must determine if the maximum pressure of the suppression is below the design strength of the vessel. Additionally, due to the typical size of a silo, the cost of suppression may greatly exceed the budget of the end user, therefore venting might be a more attractive option when considering cost.

Cyclone

(from 5.3.4.1 Cyclone Separators) (CCPS, 2005)
Less susceptible to explosions when compared to baghouses, but they sometimes occur and should be protected for that possibility. The source of ignition is often electrostatically charged dust from an upstream process. "The most common protective measures for cyclones are venting and suppression" (CCPS, 2005).

The cyclone is a working vessel and therefore it may be difficult to provide successful venting to mitigate an explosive hazard. An active chemical suppressant is typically the first choice of explosion protection measures when considering cyclones. For the cyclone, venting is a valid design option; however, it should only be considered if trying to decrease Pred for a greater safety factor is not achievable with a suppressant.

Design Phase

Available Equipment

As with any real-world problem, the available equipment from which an engineer or designer may select from is limited. This tends to be the biggest constraint when attempting to provide an explosion protection design that satisfies the minimum requirements. While a theoretical solution would provide the correct answer all the time; the challenge is to design within the real-world parameter, which is represented by a limited set of equipment. In the following sections, the selected equipment that is able to be utilized for an acceptable explosion protection scheme is given. This equipment will then be inserted into the design phase to establish a proper explosion protection design.

Mean Time Between Failures

A key component of the residual risk analysis calculation is driven by the mean time between failures (MTBF) as well as the system design. In a real-world design, it would be necessary to have all the information of each component available for design. Since this is a case study model, it is necessary to provide MTBF data for all of the available equipment from which the design may utilize. The MTBF numbers seen in this section is derived directly from Date's work. The basis for these numbers, as per a conversation with Dr. Rob Lade of Kidde Products (UK), comes from field approximations where the number of failure events in the field is known as well as the protection equipment involved in these failures. Below is an overview table of the MTBF data for the equipment types available. A breakdown of each individual component is provided in the following sections

Component	Mean Time Between Failure MTBF
Vent Panel Type 1	50,000
Detector	4,000
Suppressor Type 1	30,000
Suppressor Type 2	50,000
Control Panel	25,000
Valve	2000

Table 11: Date et al., Example Mean Time Between Failure Figures

Control Panel

A control panel will be used in every system design. The function of the control panel is to be the brain of the operation. Information from the detection circuits is relayed to the control panel, which then performs the necessary functions for explosion protection and any other ancillary shutdowns. In a real-world application there may be one or two different control panels available for explosion protection. There is only one control panel that is going to be utilized for the case study, and the information for it is provided below.

Protection Type	**Identification**	**MTBF** *(years)*
Control Panel	Panel 1	25,000

Table 12: Control Panel MBTF

Suppression Hardware

There are two types of explosion suppression hardware, which can be used in a vessel suppression application or as a chemical barrier establishment. As a vessel suppressant, an agent cylinder will discharge the chemical inhibitor microseconds after the control panel registers an explosion event via the detectors. As a chemical barrier, a cylinder will discharge the calculated appropriate amount of agent into the interconnection to suppress any flame jet ignition prior to its arrival. There are various sizes of agent cylinders which will all contain the same fire suppressant materials and is controlled by the control panel and the detection circuit.

Suppression-1 has an explosive charge delivery system to get the agent from the storage container into the vessel it is protecting. Alternatively, Suppression-2 has a piston driven delivery system. Both use the same type of chemical suppressant; however, the only difference is the delivery system. Either option may be used depending on the reduced pressure given with each delivery system. Since each option will get the chemical agent to the vessel with different delivery systems, there are different calculated reduced pressures with each type. These calculations can be seen in the suppression calculations section.

Protection Type	**Identification**	**MTBF** *(years)*
Explosive Charge Suppression Delivery	Suppression 1	30,000
Piston Suppression Delivery	Suppression 2	50,000

Table 13: Explosion Suppression MTBF

Venting Panel Hardware

There is only one vent panel type with regards to the mean time between failures. Vent panels are sized according to the required area for venting to create a reduced pressure less than the plan strength as seen in the venting calculation section. A design engineer would have various explosion vent panel sizes, which can be seen in the appendix of this document. However, there is only one MTBF for all the different sizes. Below is the vent panel MTBF information.

Protection Type	**Identification**	**MTBF** *(years)*
Explosion Venting (all sizes)	Vent 1	50,000

Table 14: Venting Panel MTBF

Detection Hardware

The real world options break down the detection into two major classifications: Vessel detectors used for the chemical suppressant inside a plant item, and interconnection detectors used for the establishment of a barrier in the ducting. Pressure detectors are the only type that is used on the vessels. The detectors have a variable setting of 35 mBar and 52 mBar, which will detect a rapid pressure increase (i.e. explosion) at the mentioned pressure settings. The isolation detectors are a little bit more complicated, as there are several varieties. In isolation barrier establishment, there are the following cases: flame detector only, 35mBar pressure detector, 52 mBar pressure detector, and the combination of flame and either pressure detector. Below is a table detailing the different types of detectors and their application.

Application	**Type**
Vessel	35 mBar Pressure Detector
	52 mBar Pressure Detector
Isolation Barrier	Flame Detector Only
	35 mBar Pressure Detector Only
	52 mBar Pressure Detector Only
	Combo: 35 mBar Pressure & Flame Detector
	Combo: 52 mBar Pressure & Flame Detector

Table 15: Detection Types Offered

Because of the limitations of the MTBF data, only one mean time between failure number is used for all the possible detector types. The calculations for each vertex or isolation barrier seen in the calculations section will establish which one are used in which case. To understand this deficiency, a sensitivity analysis is performed to correlate the MTBF and residual risk associated with the design for the case study.

Protection Type	Classification	**MTBF** *(years)*
All Types	Detection 1	4,000

Table 16: Detection MTBF

Mechanical Isolation Hardware

There are two types of mechanical isolation. The valve is a mechanical barrier that will be established through a detection circuit similar to a chemical isolation barrier. In addition to the active valve barrier, rotary valves pre-installed on the storage vessels have no MBTF as they will be installed in lieu of any active explosion protection measures. NFPA-654 allows them to be used on interconnections as an explosion barrier, and it is advisable to use on vessels with a low probability of ignition (NFPA 654, 2006). It is commonly used in this type of setup to avoid the costly addition of either a chemical or mechanical barrier; as it does not rely on the detection or suppression systems to work. The table below has the information for mechanical isolation hardware.

Protection Type	Classification	**MTBF** *(years)*
Mechanical Isolation	Isolation Valve 1	2,000
	Rotary Valve	N/A

Table 17: Mechanical Isolation MTBF

Calculations for Satisfying Standards

The next step is to calculate satisfying solutions to the standards and the process risk analysis. For this case study, we will utilize NFPA 68 and NFPA 69 to develop venting, suppression, and isolation strategy to satisfy the codes. In demonstrating the case study, it is important to utilize the available equipment, listed above, to determine the proper protection strategy. Provided in this section are the venting and suppression calculation results. Refer to the appendix for a detailed calculation of each parameter.

Venting Calculations

The minimum required level of protection for explosion venting is determined through the calculations from NFPA 68. The first step is to determine the length/diameter correlation, which is assumed as a bottom up flame propagation to resemble a worst-case scenario for the locations of the explosion relief vent panels. Additionally, venting ducts are not used due in this demonstration; it is assumed that all venting can be safely released into the atmosphere without a duct (NFPA 68, 2007). The panels, which come in various pre-determined sizes, are the venting hardware used for this case study model. All reduced pressures (Pred) are a relationship between actual panel areas that would be fitted on each respective vessel, and a total of three options are provided for each vessel (NFPA 68, 2007). All calculations can be seen in the appendix; provided below are the summaries of the calculations performed.

System Geometry		**Hammermill**	**Cyclone**	**Baghouse**	**Silo**
Vessel Type	*Description*	Cylindrical Vessel with Conicle Hopper: Bottom Up	Cylindrical Vessel with Conicle Hopper: Bottom Up	Rectangular with Hopper Extension: Bottom Up	Cylindrical Vessel with Conicle Hopper: Bottom Up
Total Volume	*[m^3]*	10	9	50	110
Storage Ht.	*[m]*	3	3	7.6	8.4
Hopper Ht.	*[m]*	1	1	1	1
Main Dia.	*[m]*	1.92	1.82	2.5	4
Hopper Dia.	*[m]*	0.5	0.5	0.5	0.5
Vent Location	*[m] from top*	0.1	0.1	0.1	0 - Vents on Top

Table 18: System Geometry used for NFPA 68 Venting Calculations

The table above is a summary table of the necessary information needed to perform a venting calculation. As previously mentioned, the vessels are considered to be bottom-up flame propagation, is considered a worst case scenario as the detection is typically placed on the top of the vessel, which is the furthest point from a likely explosion event. Similarly, the vessel types are modeled as either a cylindrical vessel or rectangular vessel depending on the geometry of the process equipment in the case study. All units are metric.

The information from the above table 20 is used to determine the Length/Diameter (L/D) calculations as performed in Annex-A of NFPA 68. The height (H), effective volume (Veff), effective area (Aeff), diameter (Dhe) and the L/D are all directly calculated from the above geometry based on the process equipment analysis. These calculations, provided below, are used in the determination of the effective area required to successfully mitigate an explosion by keeping the reduced pressure below the plant strength.

L/D Calcs		Hammermill	Cyclone	Baghouse	Silo
H	*[m]*	3.9	3.9	8.5	9.4
Veff	*[m³]*	9.67	8.74	49.46	110.3
Aeff	*[m²]*	2.48	2.24	5.82	11.74
Dhe	*[m]*	1.77	1.69	2.41	3.86
L/D	-	2.19	2.31	3.52	2.43

Table 19: Length over Diameter Calculation Results from NFPA 68

Provided below in the Venting Option 1 table are the minimum satisfying venting requirements to successfully provide explosion venting mitigation below the maximum plant strength. Because designers are restricted to real-world panel sizes (see appendix), the reduce pressure (Pred) indicated below is the first available panel size that provides pressures below the maximum plant strength. This would be considered the minimum satisfying option given the limitations of panel sizes. Under the current process for providing explosion protection, this venting requirement would be the baseline design where the engineer would have the option to increase the size of the vent panel based on engineering judgment. Under the new process, we can effectively determine if these vent panel sizes are adequate for the system in relation to the residual risk coordinated with the required SIL. The calculations are performed using NFPA 68, and can be seen in the appendix of this paper. Below are the results of the calculations to meet the minimum requirements.

Venting Option 1		Hammermill	Cyclone	Baghouse	Silo
Panels	*[No.]*	1	1	3	3
Size Panel	*[in x in]*	12 x 24	27 x 40	27 x 66	25.5 x 57.1
Size Panel	*[m²]*	0.152	0.633	1.061	0.904
Panel Mass	*[kg/m²]*	19	19	19	19
Vent Area	*[m²]*	0.152	0.633	3.183	2.712
Pmax	*bar-g*	2.0	0.5	0.2	0.4
Pred	*bar-g*	1.91	0.19	0.17	0.375
Safety Margin	-	0.09	0.31	0.03	0.025

Table 20: Venting Option 1 for Residual Risk Analysis

If we were following the current process, this is where we would stop and provide a design to go to bid. However, with qualitative information from the residual risk calculations, the design engineer now has the information necessary to determine if the minimum satisfying requirements are aligned with the required SIL for the explosion protection scheme. In an effort to keep all the data in one place, additional venting options that exceed the minimum requirements are provided below. In the new process, these would be performed on an iterative as-needed basis, where if the residual risk calculations require a lower Pred venting value for a vessel, a calculation will be performed to provide one. In demonstrating the methodology through a case study, the options provided below are simply options that may or may not be used later in the design process. They were determined by selecting Pred values that were significantly lower than the minimum required reduced pressure. In terms of real-world application, there are many more available options to the design engineer; the tables below are a representative sample of two options that exceed the minimum requirements. As the design phase continues, the actual place where these would be calculated will be explained; for purposes of this paper, it is simpler to show various calculation options in one place for reference.

Venting Option 2		**Hammermill**	**Cyclone**	**Baghouse**	**Silo**
Panels	*[No.]*	1	1	3	3
Size Panel	*[in x in]*	19.7 x 19.7	27.5 x 44.3	44 x 44	36 x 44
Size Panel	*[m^2]*	0.213	0.717	1.165	0.946
Panel Mass	*[kg/m^2]*	19	19	19	19
Vent Area	*[m^2]*	0.213	0.717	3.495	2.838
Pmax	*bar-g*	2.0	0.5	0.2	0.4
Pred	*bar-g*	1.19	0.15	0.14	0.35
Safety Margin	-	0.81	0.35	0.06	0.05

Table 21: Venting Option 2 for Residual Risk Analysis

Venting Option 3		**Hammermill**	**Cyclone**	**Baghouse**	**Silo**
Panels	*[No.]*	1	1	3	3
Size Panel	*[in x in]*	18 x 24	30 x 44	27.5 x 51.6	27 x 66
Size Panel	*[m^2]*	0.239	0.781	1.301	1.061
Panel Mass	*[kg/m^2]*	19	19	19	19
Vent Area	*[m^2]*	0.239	0.781	3.903	3.183
Pmax	*bar-g*	2.0	0.5	0.2	0.4
Pred	*bar-g*	1.0	0.12	0.115	0.28
Safety Margin	-	1.0	0.38	0.085	0.12

Table 22: Venting Option 3 for Residual Risk Analysis

Suppression Calculations

The minimum required level of protection for explosion suppression and vessel isolation is determined through from NFPA 69. Four total options are provided for each type of vessel, which is a function of matching the various types of suppression equipment with the available detection equipment. Both an explosive charge and a piston delivery system of the suppression agents were tested as well as detectors that have sensitivities at 35mbar and 52mbar respectively. As a design engineer, calculating all options is not necessary for a real world application. Similar to the venting options above, the information from the residual risk calculations will determine whether or not a new option is needed to provide a Pred adequate for achieving a certain SIL. In this demonstration every option for the given protection equipment available is calculated, some of which are not going to be used in this calculation. Since the calculation options are performed up-front, the actual place in the design phase where these would be calculated will be explained in the appropriate section.

All calculations can be seen in the appendix; provided below are the summaries of the calculations performed. The tables have been highlighted to show if the reduced pressure is greater or less than the maximum allowed pressure (Pmax) in each vessel. It was shown that suppression is not feasible on the baghouse which may limit the total options of the protection scheme.

		Hammermill	Cyclone	Baghouse	Silo	Pass
Pmax	*bar-g*	2.0	0.5	0.2	0.4	Fail

Table 23: Maximum Pressures per Each Vessel

Below are the results from the calculations (see appendix) for the Hammermill vessel. For Suppression-1, which is an explosive charge delivery system, the required amount of cylinders is two regardless of the detection setting. Each detection setting (either 35 or 52 mbar) results in a different reduced explosive pressure of the vessel based on the dust's characteristics and the time it takes to chemically inhibit the explosive atmosphere. The reduced explosion pressure is the resulting pressure inside the vessel due to the incipient phases of the explosion. However, the rapid discharge of the extinguishing agent will also produce a contributing pressure increase. The reduced explosion pressure and suppressor contribution are added to give a total reduced pressure, which must be below the maximum pressures allowed for each vessel.

In the Hammermill calculations, it was necessary to reduce the overall pressure to below 2.0 bar (gauge pressure), which was easily feasible for both suppression delivery options. Depending on the option provided below, the delta value between the Pred and Ps is varied below this threshold. Suppression 1 paired with a 52mbar detector has the highest Pred value (smallest delta between Ps), and Suppression 2 paired with a 35mbar detector has the lowest Pred value (largest delta between Ps). To determine which suppression option is most appropriate for the protection will be determined by the residual risk calculations and the need for a lower Pred based on achieving the desired SIL. If using suppression, the baseline design will use the highest value of Pred, which represents the lowest value over the minimum requirements.

Hammermill	**Suppression 1**		**Suppression 2**	
Detection Setting	35 mbar	52 mbar	35 mbar	52 mbar
Equipment Quantity	2	2	2	2
Reduced Explosion Pressure *(barg)*	0.143	0.192	0.111	0.149
Suppressor Contribution *(barg)*	0.093	0.093	0.105	0.105
Total Reduced Pressure *(barg)*	0.236	0.285	0.216	0.254

Table 24: Hammermill Suppression Calculations

Similar to the hammermill, the cyclone calculations are all considered acceptable to satisfy the minimum requirements for explosion protection as determined by NFPA 69. The maximum plant strength for the cyclone is 0.5 bar, which is maintained because the total reduced pressure for all the options provided in the table below do not exceed the plant strength.

Cyclone	**Suppression 1**		**Suppression 2**	
Detection Setting	35 mbar	52 mbar	35 mbar	52 mbar
Equipment Quantity	2	2	2	2
Reduced Explosion Pressure *(barg)*	0.139	0.189	0.108	0.146
Suppressor Contribution *(barg)*	0.104	0.104	0.117	0.117
Total Reduced Pressure*(barg)*	0.243	0.293	0.225	0.263

Table 25: Cyclone Suppression Calculations

The baghouse is a thin-walled vessel, which can only withstand a maximum pressure increase of 0.2 bar-g. Regardless of the delivery type and the quantity of suppressors, the system could not inhibit the explosion prior to it exceeding the maximum pressure of the vessel. Therefore the designer can conclude that for this particular case-study, suppression is not an option with the available equipment, and venting should be used as a viable explosion protection option.

Baghouse	**Suppression 1**		**Suppression 2**	
Detection Setting	35 mbar	52 mbar	35 mbar	52 mbar
Equipment Quantity	4	4	4	4
Reduced Explosion Pressure *(barg)*	0.259	0.34	0.108	0.17
Suppressor Contribution *(barg)*	0.085	0.085	0.117	0.084
Total Reduced Pressure*(barg)*	0.344	0.425	0.225	0.254

Table 26: Baghouse Suppression Calculations

Similar to the baghouse described above, not all options of suppressants will work in reducing the pressure below the plant strength. The only system that did not conform to the requirement of providing an explosive atmosphere below its plant strength is the Suppression 2 delivery system paired with a 52 mbar detector. This limits the total available suppression options to three for this case study.

Silo	**Suppression 1**		**Suppression 2**	
Detection Setting	35 mbar	52 mbar	35 mbar	52 mbar
Equipment Quantity	5	5	6	6
Reduced Explosion Pressure *(barg)*	0.098	0.136	0.3	0.39
Suppressor Contribution *(barg)*	0.048	0.048	0.057	0.057
Total Reduced Pressure *(barg)*	0.146	0.184	0.357	0.447

Table 27: Silo Suppression Calculations

Isolation Calculations

The final calculations that are needed to be performed are the isolation between every interconnection in the system layout. A design engineer will take the equipment available to provide a barrier and provide the required gas if (a suppressor is used) and the proper location of the device. This is a function of the time to barrier establishment versus the time to detection which is described in the background sections. Unlike the above venting and suppression calculations, the calculations will be performed on the as-needed basis as determined by the residual risk calculations. Below is a summary table of all possible (16 in total) isolation options, which include a passive rotary valve, suppressor 1, suppressor 2, and a mechanical valve paired with the various offered detection detailed in the sections above.

	Vessel Isolation Options															
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Detector Type	-	Flame Detector	35mbar pressure	52mbar Pressure	35mbar Pressure & Flame	52mbar pressure & Flame	Flame Detector	35mbar pressure	52mbar Pressure	35mbar Pressure & Flame	52mbar pressure & Flame	Flame Detector	35mbar pressure	52mbar Pressure	35mbar Pressure & Flame	52mbar pressure & Flame
Isolator Type	Rotary Valve	Suppressor 1					Suppressor 2					Mechanical Valve				

Table 28: Available Isolation Options

Rotary valves are not calculated, as they are a passive explosion protection inhibitor in duct connections. Mechanical valves and chemical isolation barriers are calculated based on distance required from each vessel and the quantity (suppressors) of cylinders needed. The calculations will give a range of acceptable locations along the interconnection path, which satisfies the minimum NFPA requirements. Moving outside the allowable range with respect to each vessel connection will violate the minimum requirements, which translates to an unacceptable design. The calculations of each used in each scenario can be seen in the appendix.

Residual Risk Calculations

To determine the residual risk, a beta calculation tool from Kidde Products-UK, has been developed that incorporates the information from the system and utilizes the equations and directed graphical representation to develop a per-vertex residual risk, and a per-ignition residual risk figure. To understand the engineering approach to the calculations, please refer to the hand calculated example in section 3. The tool used to calculate the residual risk is cumbersome at this point in time, where a single calculation takes 30-45 minutes to program and calculate depending on the complexity of the design. The residual risk equations can be programmed into any calculation engine; therefore a design engineer has the ability to analyze complicated systems without performing difficult hand calculations.

When looking at the total risk of the entire explosion protection system, there are two differing terminologies. The per-vertex residual risk assumes ignition in all of the plant items. It then calculates the risk of ignition in one plant item from its probability of occurrence and from a connecting plant items by its probability of ignition. Conversely, a per-ignition residual risk assumes ignition in a single plant item, and calculates the risk of failure from the plant item that had the ignition as well as the connected plant items. It is important to reiterate that the total residual risk of the system is calculated by taking the sum of the individual components (either per-ignition or per-vertex residual risk), where the per-ignition and per-vertex additive residual risk are equal for all cases in this case study due to the equal probability of ignition in the vessels. For applications where the total residual risk differs, the worst case scenario total risk (per-ignition or per-vertex) will govern.

Baseline Design

The preceding sections have walked through the Assessment Phase of the new process. Additionally the above sections beings the Design Phase by determining minimum required explosion protection requirements as dictated by NFPA 68 and NFPA 69, and the limited available equipment to provide a design that meets the governing standards and the Process Risk

Assessment. As the designer, the aim is to provide a system that satisfies the minimum governing requirements as the baseline design. Listed below is a protection overview of each vessel and each interconnection for the baseline design, which conforms to the minimum requirements of NFPA 654, NFPA 68, and NFPA 69.

The selection of the equipment is not arbitrary; selecting the baseline design is similar to the current process for providing explosion protection. The design engineer must consider the PRA as well as the initial calculations. The design below is considered a satisfying design because it meets all the criteria of the governing documents as seen in the calculation sections of 5.4.2. In using the new process for providing explosion protection, a residual risk calculation is performed to quantify the residual risk assumed with the current design. Once calculated the engineering will check it against the benchmark SIL-2 to determine if the protection scheme needs to be optimized.

Vessel Protection

Hammermill: (2) Suppression-1 paired with (1) 52 mbar Detector, *[Pred = 0.285 bar]*
Cyclone: (1) Suppression-1 paired with (1) 52 mbar Detector, *[Pred = 0.293 bar]*
Baghouse: (3) 44”x44” Vent-1, *[Pred = 0.17 bar]*
Silo: (3) 36”x44” Vent-1, *[Pred = 0.375 bar]*

Interconnection Protection

Hammermill – Cyclone: (1) Suppression-1 paired with (1) Flame Detector
Cyclone – Baghouse: (1) Suppression-1 paired with (1) Flame Detector
Cyclone – Silo: (1) Passive Rotary Valve
Baghouse – Silo: (1) Passive Rotary Valve

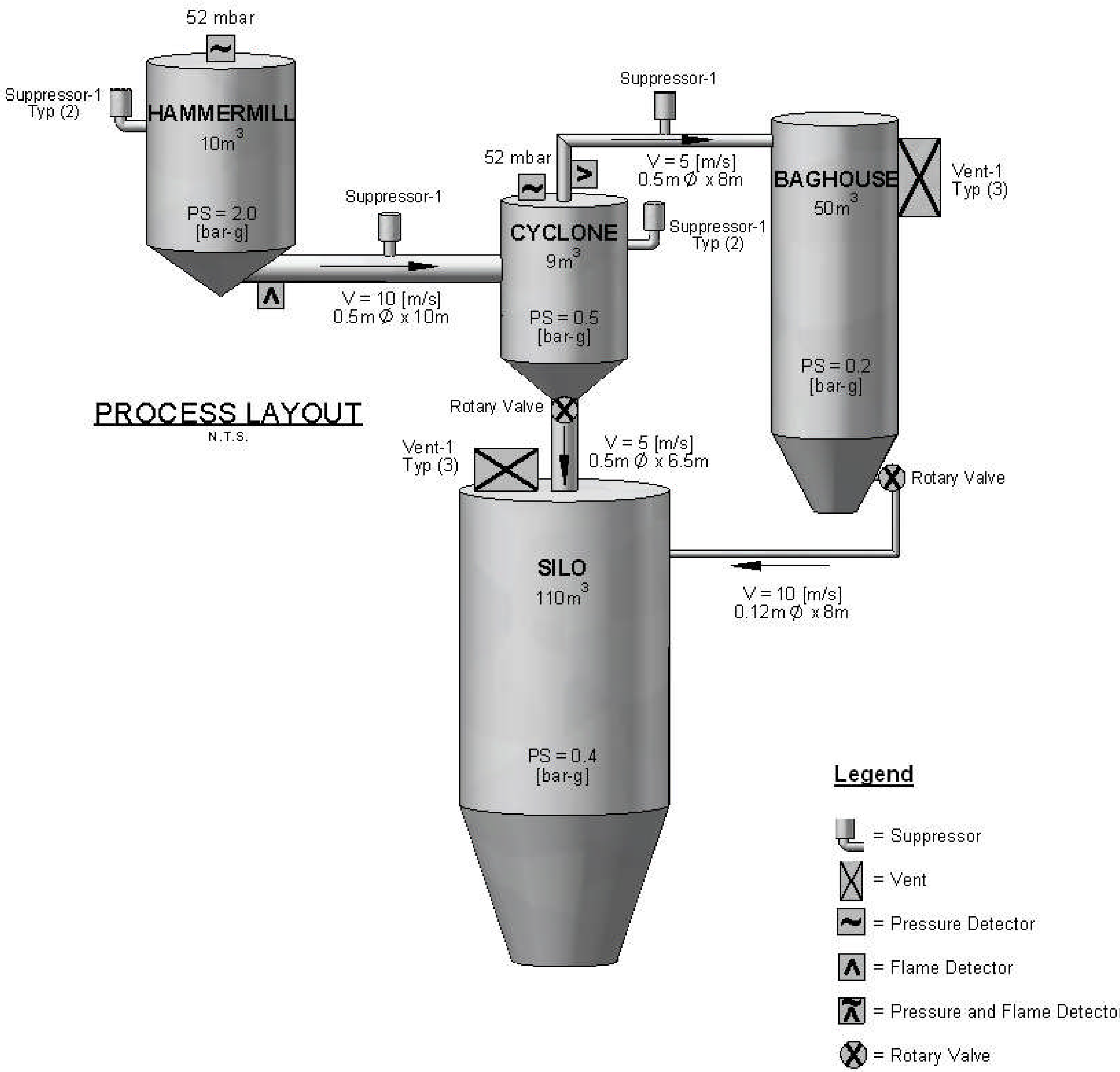

Figure 15: Graphical Protection Scheme for Baseline Design

Residual Risk Input Variables Overview

All the information provided in this section is necessary for the baseline design for the residual risk calculations. In addition provided information, it is also crucial to import the geometry and characteristics of the system, which includes: the size of each vessel; the size and length of the interconnections; the velocity of the air movement of the interconnections; and the explosive characteristics of the dust. With all the information, we can accurately calculate the overall residual risk of the explosion protection design. The table below provides the variable, description, and the value of all the input values for the residual risk calculation. Anything shown with a value is going to hold true for all systems shown in this case study. Since the designs will vary with the different proposals, the input values that are changing will be displayed in each section. Please refer to the hand calculations for a more detailed explanation on how the equations interact for residual risk.

Variable	Description	Value
$Q_E(i)$	Ignition probability in vertex i.	1.0 for all vessels. Worst case scenario is to assume that each vessel has a 100% probability of ignition.
$P_{red}(i,j)$	Reduced explosion pressure in vertex i following an ignition in vertex j.	See Each Vessel (bar-g)
$P_s(i)$	Pressure shock resistance of vertex i.	See Each Vessel (bar-g)
$Q_f^s(i,j)$	Probability of flame propagation between connected vessels i and j which then leads to an enhanced explosion in vertex j.	1.0 for all connections. Worst case scenario is to assume that each duct has a 100% probability of propagation.
$D_{min(i)}$	Minimum distance of installed isolation barrier from vertex *i*	See Each Interconnection (m)
$D_{max(i)}$	Maximum distance of installed isolation barrier from vertex *i*	See Each Interconnection (m)
$D_{min(j)}$	Minimum distance of installed isolation barrier from vertex *j*	See Each Interconnection (m)
$D_{installed(i)}$	Installed distance of installed isolation barrier from vertex *i*	See Each Interconnection (m)
π_1	Reciprocal Mean Time Between Failures (MTBF) value for the Suppressor 1	3.3×10^{-5}
π_2	Reciprocal MTBF value for the Suppressor 2	2.0×10^{-5}
π_3	Reciprocal MTBF value for the vent panel	2.0×10^{-5}
π_4	Reciprocal MTBF value for the Detector	2.5×10^{-4}
π_5	Reciprocal MTBF value for the Control Panel	4.0×10^{-5} : (1) Unit for the entire System
π_6	Reciprocal MTBF value for the Mechanical Isolation Valve	5.0×10^{-4}

Table 29: Input Parameters for Residual Risk Calculations

Vessel Input

Provided in the following sections are the component-by-component and interconnections inputs that are used to calculate the residual risk of the system. For an example on how to hand-calculate a system; refer to the example calculation to understand the interactions of the equations. The inputs to the residual risk calculations shown in this section are representative of the entire explosion protection design. The inputs and results are given in table format for convenience. Please refer to the Input Parameters for Residual Risk Calculations table above for the description of the nomenclature and the units of measure associated with each value.

Hammermill Vessel (V_1):

Item	Value or Quantity
π_1	2
π_2	0
π_3	0
π_4	1
$P_s(i)$	2.0
$P_{red}(i, j)$	0.285

Cyclone Vessel (V_2):

Item	Value or Quantity
π_1	2
π_2	0
π_3	0
π_4	1
$P_s(i)$	0.5
$P_{red}(i, j)$	0.293

Baghouse Vessel (V_3):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.2
$P_{red}(i, j)$	0.17

Silo Vessel (V_4):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.4
$P_{red}(i, j)$	0.375

Inter-Connections Input

Hammermill – Cyclone (V_1-V_2):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	3.2
$D_{max(i)}$	8.2
$D_{min(j)}$	2.1
$D_{installed(i)}$	5.0

Cyclone – Baghouse (V_2-V_3):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	3.1
$D_{max(i)}$	8.1
$D_{min(j)}$	2.1
$D_{installed(i)}$	5.0

Cyclone – Silo (V_2-V_4): No Active Explosion Protection (Rotary Valve)
Baghouse – Silo (V_3-V_4): No Active Explosion Protection (Rotary Valve)

Baseline Residual Risk Calculation Results

Per Vertex Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	1.24E-02	81	98.760%	SIL-1
Hammermill	1.42E-04	7019	99.986%	SIL-3
Cyclone	3.18E-02	31	96.820%	SIL-1
Baghouse	5.34E-05	18877	99.995%	SIL-4
Total Risk	4.44E-02	23	95.560%	SIL-1

Table 30: Baseline Per Vertex Residual risk

Per Ignition Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	7.55E-03	132	99.245%	SIL-2
Hammermill	1.42E-04	7019	99.986%	SIL-3
Cyclone	1.15E-02	87	98.850%	SIL-1
Baghouse	2.52E-02	40	97.480%	SIL-1
Total Risk	4.44E-02	23	95.561%	SIL-1

Table 31: Baseline Per Ignition Residual risk

Residual Risk Analysis and Resulting SIL

The total calculated risk in this entire system is shown above as 4.44×10^{-2}, which is the addition of all the process components individual risk via the residual risk calculations. This number corresponds to a failure rate of 1-in-23 event, where if the system were to be called on 23 times, it would be likely that there would be an explosion mitigation failure in one of those instances. When corresponding this to Safety Integrity Levels, the overall availability of the system, which is equal to one-minus-residual risk, is just over 95.5% system availability. Quantitatively, this would conclude that the system corresponds directly to a SIL-1 rating, which does not meet the minimum requirement initial set in the PRA.

Target SIL

In selecting the individual protection components that reduce the pressures below the plant strength and provide connection barriers, this system conforms to the NFPA code literature, and is considered as code-satisfying design. However, it is clear through the residual risk calculations that the baseline design, while code compliant, does not meet the SIL required for an explosion protection system. The importance of residual risk calculations is clear; the ability to provide a code-satisfying design with risk levels unacceptable to a PRA is possible. Under the current process, it is conceivable to provide this design without being concerned with its consequence of failure. However, in utilizing and demonstrating the new process, the design engineer has the ability to optimize the system by reducing the overall risk associated with the design.

To optimize this design, the design engineer must analyze the residual risk of each component and as an entire system to see which vessels or interconnects are the problem areas. For this baseline design, it is clear that there are a few vessels that carry a high amount of risk with this design. The cyclone carries particularly high risk for each calculation, which is of immediate concern. Additionally, downstream of the cyclone the silo and baghouse carry elevated levels of risk, which may be due to the heightened risk of the cyclone. To optimize the system the design engineer might start by changing the explosion protection strategy on the cyclone, and possibly the interconnections to the downstream vessels.

Optimized Design – Option 1

In the baseline design of this case study, the SIL-1 rating is not satisfactory, and it providing a minimum SIL-2 protection strategy is required. Since residual risk calculations are new in the industry, there is no guideline to effectively mitigating risk with specific design strategies. Using engineering judgment in a trial and error basis is the only method currently available to minimize risk. By analyzing the residual risk calculations of the initial baseline design for the case study, the cyclone provides the biggest level of concern. Therefore the design strategy for the optimized design solution will focus around this vessel and possible downstream vessels.

At this point in the new process, a calculation would be performed on various types of protection equipment to deliver reduced pressures via venting or isolation. In a real-world design, the engineer would calculate based on the equipment available and the needs for the system. From the results of the residual risk calculations, the design engineer would calculate for the following options: reduced Pred using available suppression and detection on the cyclone; reduced Pred using venting with available vent panels for the cyclone; and active isolation barriers using the available equipment for the cyclone-baghouse, hammermill-cyclone and cyclone-silo interconnections. With the exception of the interconnection calculations (seen in the appendix) the venting and suppression options are provided in the equipment calculation section, and are the designs from which this demonstration will select.

Failed Design Option 1

Providing an optimized solution is an iterative process; the baseline system is modified, a residual risk calculation is performed, and the total residual risk is checked the required SIL for system acceptability. Provided in section 5.4.5.2 is a detailed description of successfully achieving a code satisfying system that meets the risk level of the associated SIL. The design engineer may test and trial multiple potentially viable solutions until one is met. Provided below is a brief description of design changes that did not reduce the risk enough to achieve a SIL-2 classification.

The cyclone was the primary vessel of concern for the optimization to SIL-2. A calculation for utilizing each suppression option seen in the equipment calculations section was performed, but did not achieve SIL-2 residual risk levels. The new Pred of each option did not provide a large enough delta to the cyclone Ps to statistically reduce the likelihood of a failure of the vessel failing. Therefore a venting solution was the next course of action.

Changing the protection scheme to the venting option-1 value for the cyclone reduced the residual risk of failure for the cyclone and for the downstream silo but not to SIL-2 risk levels. Additionally, the downstream baghouse was protected at a SIL-1 rating, which governed the total risk. Therefore it was necessary to change the design of two interconnections protection types: the upstream hammermill, and the cyclone and baghouse connection. Changing only one of the two interconnections did not reduce the total risk to a SIL-2, thus it is necessary to change both. The following section provides a detailed description of the design that successfully achieves SIL-2 for the total system residual risk.

Successful Design Option 1

In the baseline design, the cyclone was protected with two explosion protection suppressors to create a reduced pressure of 0.293 bar. In order to significantly reduce the Pred, it is necessary to move to a venting solution. In this scenario we are providing a 27 x 44 inch vent to reduce the Pred value to 0.19 bar. By reducing the Pred value, the new design statistically decreases the likelihood (see hand calculations for an example) that the Ps is exceeded. In addition to providing protection at the cyclone itself, removing the rotary valve and am substituting it with an active isolation suppressant paired with a pressure detector at the prescribed location. The final design change to achieve SIL-2 is to replace the detection on the interconnection between the hammermill and cyclone with a 35mbar pressure and flame detector, which reduces the likelihood of violating the time to barrier establishment. Below is an itemized list of the protection strategy, as well as a diagram for ease of understanding.

<u>Vessel Protection</u>

Hammermill: (2) Suppression-1 paired with (1) 52 mbar Detector *[Pred = 0.285 bar]*
Cyclone: (1) 27” x 40” Vent-1 *[Pred = 0.19 bar]*
Baghouse: (3) 44”x44” Vent-1 *[Pred = 0.17 bar]*
Silo: (3) 36”x44” Vent-1 *[Pred = 0.375 bar]*

Interconnection Protection

Hammermill – Cyclone: (1) Suppression-1 with (1) 35 mBar pressure & flame detector
Cyclone – Baghouse: (1) Suppression-1 with (1) 52 mBar detector
Cyclone – Silo: (1) Suppression-1 with (1) 52 mBar detector
Baghouse – Silo: (1) Passive Rotary Valve

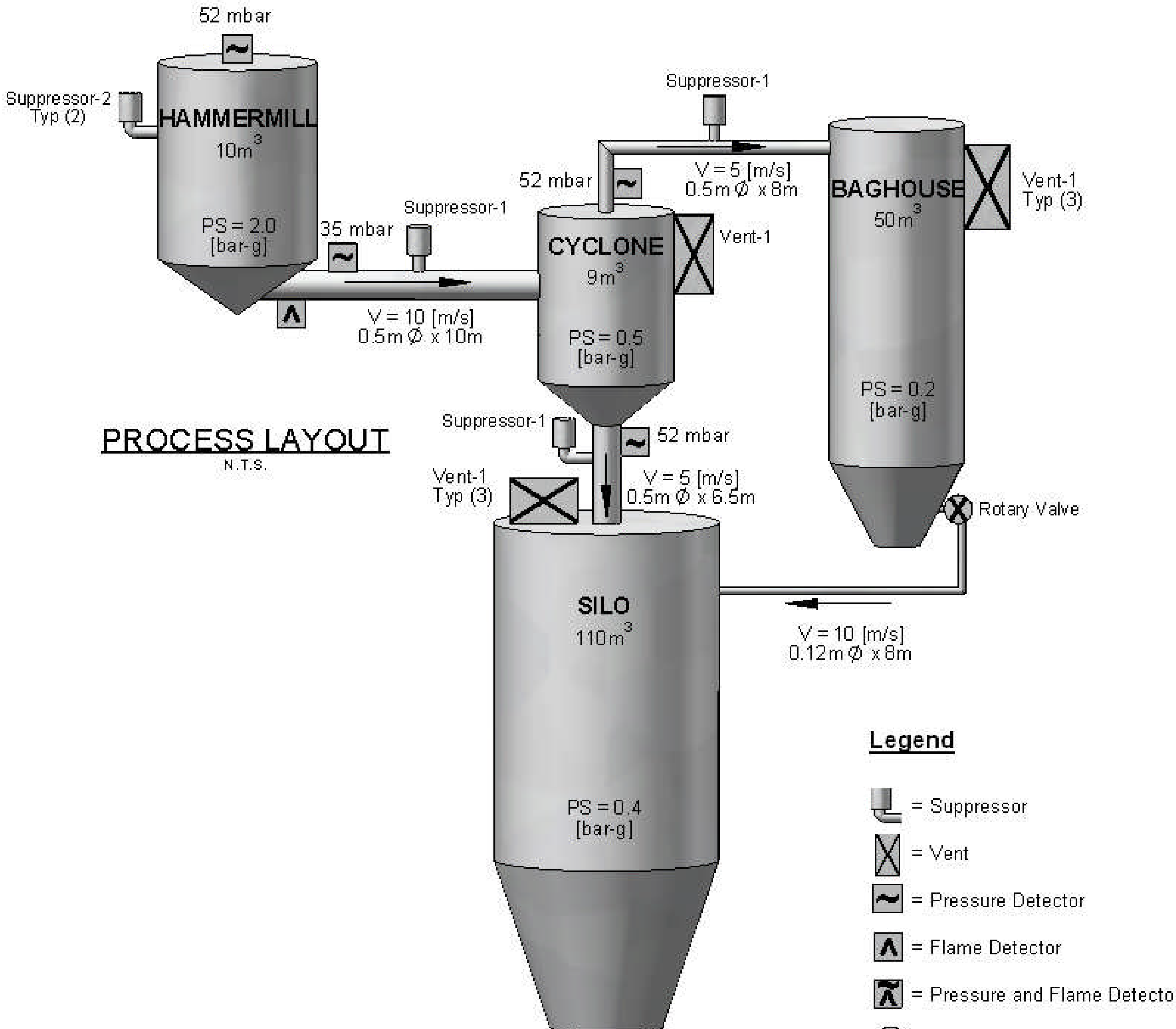

Figure 16: Graphical Protection Scheme for Optimized Design Option 1

Residual Risk Input Variables Overview

Vessel Input

Provided in the following sections are the component-by-component and interconnections inputs that are used to calculate the residual risk of the system. The references for each item can be seen in the table 31 above. For an example on how to hand-calculate a system; refer to the example calculation to understand the interactions of the equations. The inputs and results are given in table format for convenience.

Hammermill Vessel (V_1):

Item	Value or Quantity
π_1	2
π_2	0
π_3	0
π_4	1
$P_s(i)$	2.0
$P_{red}(i,j)$	0.285

Cyclone Vessel (V_2):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.5
$P_{red}(i,j)$	0.19

Baghouse Vessel (V_3):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.2
$P_{red}(i,j)$	0.17

Silo Vessel (V_4):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.4
$P_{red}(i,j)$	0.375

Inter-Connections Input

Hammermill – Cyclone (V_1-V_2):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	1.3
$D_{max(i)}$	6.3
$D_{min(j)}$	2.1
$D_{installed(i)}$	2.0

Cyclone – Baghouse (V_2-V_3):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	3.1
$D_{max(i)}$	8.1
$D_{min(j)}$	2.1
$D_{installed(i)}$	4.0

Cyclone – Silo (V_2-V_4):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	2.5
$D_{max(i)}$	7.5
$D_{min(j)}$	2.0
$D_{installed(i)}$	2.5

Baghouse – Silo (V_3-V_4): No Active Explosion Protection (Rotary Valve)

Option-1 Residual Risk Calculation Results

Per Vertex Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	1.24E-03	806	99.876%	SIL-2
Hammermill	1.61E-04	6211	99.984%	SIL-3
Cyclone	4.77E-03	210	99.523%	SIL-2
Baghouse	5.59E-05	17889	99.994%	SIL-4
Total Risk	**6.23E-03**	**161**	**99.377%**	**SIL-2**

Table 32: Option-1 Per Vertex Residual Risk

Per Ignition Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	1.17E-03	855	99.883%	SIL-2
Hammermill	1.43E-04	6993	99.986%	SIL-3
Cyclone	2.98E-04	3356	99.970%	SIL-3
Baghouse	4.61E-03	217	99.539%	SIL-2
Total Risk	**6.22E-03**	**161**	**99.378%**	**SIL-2**

Table 33: Option-1 Per Ignition Residual Risk

Option-1 Residual Risk Analysis and Resulting SIL

The total calculated risk for the optimized solution – option 1 above is 6.22 x 10^{-3}, which is the addition of all the process components individual risk via the residual risk calculations. This number corresponds to a failure rate of 1-in-161 events, where if the system were to be called on 161 times, it would be likely that there would be an explosion mitigation failure in one of those instances. When corresponding this to Safety Integrity Levels, the overall availability of the system, which is equal to one-minus-residual risk, is just over 99.3% system availability. On a purely quantitative level, this would conclude that the system corresponds directly to a SIL-2 rating, which successfully meets the requirement of the PRA.

For this solution, it was possible to achieve the required benchmarked SIL-2 by modifying the design approach around the cyclone vessel. In lowering the Pred value via the installation of a vent panel as well as installing an active isolation suppressant, there were immediate impacts on the risk levels. This is not the only design that will satisfy the minimum requirements of the code and maintain residual risk below SIL-2. For the design engineer, there are many viable options which can provide risk levels acceptable to the process owner while maintaining NFPA system requirements. However, without the ability to calculate the residual risk of the system there is no way to determine how much risk is associated with each design approach. In demonstrating the case study, redesigning the baseline design and analyzing the changes in risk is an iterative process, which mixes trial and error with engineering judgment. The solution detailed in this case study is one of many potential designs that will meet the entire criterion for a minimum code and risk satisfying system.

Target SIL

Since the system meets all the qualifications of an "acceptable" system, the design engineer may decide that this option can be placed to bid. However, one may want to further reduce the risk associated with the design, to achieve the next highest SIL classification; SIL-3. In analyzing the residual risk calculations, the vessels of concern are the cyclone and downstream baghouse, which carry the highest amount of risk.

Optimized Design – Option 2

The optimized design in the above section provided a design that meets the minimum requirements of the NFPA standards as well as meeting the target risk requirements. To achieve SIL-3, which is above the requirement for this case study, please refer to Appendix 2.

7 Assumptions and Limitations

Mean Time Between Failures Sensitivity Analysis

The residual risk calculations derived from Date's work relies heavily on five major principles:

- the layout of the process;
- the Mean Time Between Failures (MTBF) of hardware,
- the reduced pressure in relation to the plant strength,
- the time to barrier establishment in relation to the propagating flame jet; and
- the probability of an explosion.

Currently, MTBF data must be estimated because there is a lack of published data from any manufacturer; therefore, the MTBF numbers seen in the following sections are derived directly from Date's work. The basis for these numbers, as per a conversation with Dr. Rob Lade of Kidde Products (UK), comes from field approximations where the number of failure events in the field is known as well as the protection equipment involved in these failures. From this field information, the MTBF figures were then back-calculated to approximate MTBF figures for certain types of components, which Date uses in his work. These data is taken from a small sample size of protection equipment and are essentially field approximation. It is assumed that while the numbers are approximations, the component MTBF data is on the correct order of magnitude. Because of this, a sensitivity analysis is performed on the case study process to provide a correlation factor to understand how MTBF for hardware used affects the overall residual risk numbers. The remaining governing principles that are used in residual risk calculations are either calculated based on input data or are reasonably assumed.

Methodology

To test the residual risk calculation's sensitivity to varying MTBF data, it is important to establish controls. The overall residual risk can vary widely by changing any of the major contributors listed above. Because of this, the baseline design is going to be used as the basis for the testing, where the process will not change, the protection equipment is exactly the same, and only the MTBF numbers will vary. The table below contains the baseline figures for the MTBF that are used in the initial calculation. To see basis of the protection scheme, please refer to the section 5 for the methodology of establishing the baseline protection design.

Component	Mean Time Between Failure MTBF
Vent Panel Type 1	50,000
Detector	4,000
Suppressor Type 1	30,000
Suppressor Type 2	50,000
Control Panel	25,000
Valve	2000

Table 34: Date et al., Example Mean Time Between Failure Figures

There is no way to determine if these numbers are true approximations of actual explosion protection components because there are no published MTBF data from any manufacturer at this time. Even approximating certain components such as similar detectors or solenoids will not give a true representation of the actual MTBF. Therefore, to determine the impact of MTBF on the overall residual risk, the baseline figures in the table above were varied positively and negatively 1%, 5%, 10%, 25%, and 50%. Even though Date's numbers are approximations, varying the figures in increments from one to fifty percent in the positive and negative direction will the overall sensitivity to these changes. Below are the changes in MTBF.

	-50%	-25%	-10%	-5%	-1%	BASELINE	1%	5%	10%	25%	50%
Vent 1	25000	37500	45000	47500	49500	50000	50500	52500	55000	62500	75000
Suppression 1	15000	22500	27000	28500	29700	30000	30300	31500	33000	37500	45000
Suppression 2	25000	37500	45000	47500	49500	50000	50500	52500	55000	62500	75000
Detection 1	2000	3000	3600	3800	3960	4000	4040	4200	4400	5000	6000
Control Panel	12500	18750	22500	23750	24750	25000	25250	26250	27500	31250	37500
Valve 1	1000	1500	1800	1900	1980	2000	2020	2100	2200	2500	3000

Table 35: Varied MTBF Figures for a Sensitivity Analysis

Results

The initial baseline calculations seen in section 6 are reprinted below for ease of reference.

Per Vertex Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	1.24E-02	81	98.760%	SIL-1
Hammermill	1.42E-04	7019	99.986%	SIL-3
Cyclone	3.18E-02	31	96.820%	SIL-1
Baghouse	5.34E-05	1877	99.995%	SIL-4
Total Risk	4.44E-02	23	95.560%	SIL-1

Table 36: SIL-1 Per Vertex Residual risk

Per Ignition Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	7.55E-03	132	99.245%	SIL-2
Hammermill	1.42E-04	7019	99.986%	SIL-3
Cyclone	1.15E-02	87	98.850%	SIL-1
Baghouse	2.52E-02	40	97.480%	SIL-1
Total Risk	4.44E-02	23	95.561%	SIL-1

Table 37: SIL-1 Per Ignition Residual risk

The residual risk calculation inputs were identical to the inputs seen in section 5 – the baseline design, with the exception of the varied MTBF changes for each option. Below is a summary table for the sensitivity analysis of MTBF on overall residual risk.

	-50%	-25%	-10%	-5%	-1%	1%	5%	10%	25%	50%
Silo	1.24E-02	1.24E-02	1.24E-02	1.24E-02	1.24E-02	1.24E-02	1.24E-02	1.24E-02	1.24E-02	1.24E-02
Hammermill	2.85E-04	1.90E-04	1.58E-04	1.50E-04	1.44E-04	1.41E-04	1.36E-04	1.30E-04	1.14E-04	9.50E-05
Cyclone	3.19E-02	3.19E-02	3.18E-02	3.18E-02	3.18E-02	3.18E-02	3.18E-02	3.18E-02	3.18E-02	3.18E-02
Baghouse	1.01E-04	6.94E-05	5.88E-05	5.60E-05	5.39E-05	5.30E-05	5.12E-05	4.91E-05	4.39E-05	3.75E-05
Total Risk	4.47E-02	4.46E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.43E-02

Table 38: Per-Vertex Residual Risk Sensitivity Analysis Results

	-50%	-25%	-10%	-5%	-1%	1%	5%	10%	25%	50%
Silo	7.57E-03	7.56E-03	7.56E-03	7.56E-03	7.55E-03	7.55E-03	7.55E-03	7.55E-03	7.55E-03	7.55E-03
Hammermill	2.85E-04	1.90E-04	1.58E-04	1.50E-04	1.44E-04	1.41E-04	1.36E-04	1.30E-04	1.14E-04	9.50E-05
Cyclone	1.16E-02	1.16E-02	1.15E-02	1.15E-02	1.15E-02	1.15E-02	1.15E-02	1.15E-02	1.15E-02	1.15E-02
Baghouse	2.52E-02	2.52E-02	2.52E-02	2.52E-02	2.52E-02	2.52E-02	2.52E-02	2.52E-02	2.52E-02	2.52E-02
Total Risk	4.47E-02	4.46E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.44E-02	4.43E-02

Table 39: Per-Ignition Residual Risk Sensitivity Analysis Results

Analysis

From the results provided in the previous section, a linear correlation coefficient of **-0.88** is established between the MTBF and the overall residual risk. This suggests that there is a very strong inverse relationship between the MTBF input and the residual risk output, which means when the reliability of the equipment increases, the overall risk decreases. However, in looking at the total risk, it is clear that there is not much variation from the baseline design. While there is a strong correlation between these two values, it is important to remember the scale of its impact. The inverse MTBF (used in the calculations) are on the order of 10^{-4} to 10^{-5} in magnitude, which will have an impact more strongly on residual risk values much closer to that scale. For instance, the per-vertex residual risk sensitivity analysis results show that the Silo and Cyclone do not vary at all even with a +/- 50% MTBF change; however, the Hammermill and Baghouse have noticeable changes because the risk numbers are so low. Essentially, the scale is not the same. In the per vertex risk, the total risk is governed by higher risk vessels (Silo and Cyclone). The lower risk vessels (Hammermill and Baghouse) are affected by the variance in MTBF, but do not have an impact on the total residual risk.

The sensitivity of the accuracy of the MTBF numbers is dependent on the target risk levels trying to be achieved. If the goal is to develop a protection scheme that is SIL-1 or SIL-2, it is likely that variations in the MTBF figures similar to the above data will not greatly impact the overall risk. However, if the target SIL rating is SIL-3 or SIL-4, the impact of varying MTBF data may have a much larger impact and the accuracy is much more significant when compared to other aspects that drive the residual risk calculations.

Safety Integrity Levels

The challenge with using residual risk analysis for explosion protection systems in direct correlation with SIL ratings and PFD is that IEC 61511 does not specifically provide the requirements for these types of systems. While IEC 61511 focuses its attention on the process sector, it does not provide the requirements of other instrumented safety systems such as fire, safety alarms, safety controls, and gas systems (IEC 61508, 2004). This model is an extrapolation of known residual risk, PFD, and SIL relationships in the application towards explosion protection and mitigation.

Additionally, this work will not assume ancillary qualifying factors such as consequence of failure, preventative measures, and other information that is used by qualified risk assessors to determine the appropriate SIL for demonstrating the case study. For purposes of demonstration, it is assumed that the explosion protection system must deliver a residual risk aligned with SIL-2 or greater.

8 Contributions of Thesis and Future Work

Contributions

The problem of unmitigated explosions from inadequate explosion protection poses serious threats to the processes in operation, personnel who work around the process, and the community in which it surrounds. To overcome this, the hypothesis of this thesis is that residual risk analysis is a tool for process owners and design engineers to make investment and design decisions based on improvements in an explosive hazard or process' risk position. In exploring the solution to the industry problem, this thesis:

- Documents the current procedure for explosion protection system design which satisfies the minimum governing requirements;
- Introduces the residual risk analysis work of Date et, al. (2009), as a quantitative calculation tool for the mitigation of an explosion occurrence;
- Considers the relationship between residual risk analysis and Safety Integrity Levels for analyzing explosion designs against appropriately benchmarked risk levels;

- Proposes an updated design methodology for explosion protection which utilizes residual risk analysis, safety integrity levels, and system optimization;
- Demonstrates the use of the proposed methodology to present the owner with a quantified residual risk associated with a particular design; and
- Upgrades the discussions between stakeholders from “meets code” to “probability of failure on demand”

While this work brings the theory of residual risk analysis closer to a practical implementation into real world applications, future work is needed before widespread use of the methodology can be brought to the design engineer and the code bodies. The following section outlines the future work for this topic.

Future Work

The most important future work is to gather and document data from manufacturers regarding appropriate Mean Time Between Failures for a variety of process equipment and associated protection measures. A sensitivity analysis shows that there is a strong correlation between MTBF and total system risk associated with a particular design, depending on the order of magnitude. The MTBF numbers used in this work are approximations. The sensitivity analysis conducted demonstrated an increased importance of accurate MTBF numbers as a design attempts to meet a higher SIL. With availability of reliable and accurate MTBF numbers, the new process for providing explosion protection could be implemented into applicable codes and standards governing explosion protection, providing a more complete and concrete design approach based on actual risk.

A key component of this work is the use of Safety Integrity Levels,   to describe and quantify the level of protection provided in an explosion system design even though use of SIL for this application has not been formally established through IEC code requirements. It remains to be explored if the boundary values for each safety integrity level are socially acceptable, i.e., if the qualitative consequence of failure aligns with the quantitative system availability. Future studies may want to investigate the correlation between established risk levels and process failures lead.

For the purposes of this paper, the values set forth by the governing bodies are assumed to be socially acceptable and appropriate for this application.

Another future area of study is to investigate redundancy with the residual risk calculations. The calculations do not take into consideration any redundancies provided in an explosion protection system. For instance, it may be appropriate to assume that adding an additional venting panel as a back-up to the primary would lower the residual risk. However, there is no way to model this in the current equations, as the probability of hardware failure assumes that all components are necessary for explosion protection. For instance, one may have two suppressors, where only one is required but two actuate to ensure that there is proper suppression in the case that there is a hardware failure. The current equations assume that both would need to be discharged to provide adequate explosion protection, even though one piece of hardware is solely for redundancy. Currently there is no way to model this with the equations given by Date.

There is future work in developing this process to include or incorporate a performance based design approach. This work currently only considers a prescriptive code-based approach, which is reflected by the design steps needed to successfully delivery a satisfying system. To bring this process into code reality, seamless integration to the accepted performance based design outlined in NFPA 654 must be considered. With acceptance of residual risk analysis in the code community, there is opportunity that the risk-based design can supplant some of the performance based aspects required by the code. This thesis does not explore that possibility, and there is a great deal of work to solidify a performance based option.

Finally, there is future work in establishing a best practices methodology for determining the proper course of action to optimize a system within the new process for providing explosion protection. Using engineering judgment in a trial and error basis is the only method currently available to minimize risk, where the design engineer will analyze the residual risk calculations and start with the worst case vessel. A trial and error basis is not good engineering, and a best practices methodology would go a long way in solidifying the new process. Future work would consist of looking at multiple zoned systems as well as systems with fewer vessels than the case study example.

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10 Appendix 1 – IEC Standards

IEC 61508

The International Electrotechnical Commission (IEC) published IEC 65108 *Functional Safety of Electrical/Electronic/Programmable Electronic Safety-Related Systems* in 1998, which establishes standards for which safety-related system hardware and software must be designed (IEC 65108, 1998). This document is a general standard which first established that hazards posed by certain plant items and its associated control systems must be identified and that a risk assessment be performed (IEC 65108, 1998). Any risk determined through an assessment must be mitigated until the risk is considered tolerable with both its functional and safety integrity requirements. IEC-61508 defines safety integrity as the "likelihood of a safety-related system satisfactorily performing the required safety functions under all the stated conditions, within a stated period of time" (Redmill, 1999). This safety integrity level is benchmarked to certain levels (SIL) which are defined as, "a discrete level (one of 4) for specifying the safety integrity requirements of safety functions" (Redmill, 1999). Furthermore, the safety integrity levels are associated with probabilities of unsafe failures, and are broken down into two major categories: low demand operation, and continuous operation (IEC 65108, 1999). From this, the Probability of Failure on Demand (PFD) for different SIL ratings are provided below.

Safety Integrity Level	Probability of Failure to Perform its Safety Function on Demand
4	$\geq 10^{-5}$ to 10^{-4}
3	10^{-4} to 10^{-3}
2	10^{-3} to 10^{-2}
1	10^{-2} to 10^{-1}

Table 40: Safety Integrity Levels of Low Demand Operation (IEC 61508, 1998)

Safety Integrity Level	Probability of Failure to Perform its Safety Function on Demand
4	$\geq 10^{-9}$ to 10^{-8}
3	10^{-8} to 10^{-7}
2	10^{-7} to 10^{-6}
1	10^{-6} to 10^{-5}

Table 41: Safety Integrity Levels of Continuous Operation (IEC 61508, 1998)

IEC 61511/ ISA 84.00.01

While IEC 61508 was created to serve as a generic function safety standard, it became clear that there was a need for sector-specific standards. Using the framework established in the IEC 61508 document, IEC 61511, the *Function safety – Safety instrumented systems for the process industry sector* was developed to ensure the safety of industrial processes with the use of some type of instrumentation (IEC 61508, 2004). Much of the same language is the same as IEC 61508, but has been tailored for the process industry sector, which includes petrochemical, hazardous goods, and chemical industries (IEC 61508, 2004). The document contains three major parts, which establishes the application of the standard as well as the guidance for the determination of the required Safety Integrity Levels. After its publication, the European Standards Body, European Committee for Electrotechnical Standardization (CENELEC), adopted the standard as EN 61511, which gives each state in the EU a specific published national standard identical to the original IEC standard (CENELEC, 2010).

The United States has an ANSI/ISA 84.00.01, issued in 2004, which is largely based on the IEC 61511 language with the exception of some grandfathering clauses (ISA 84.01, 2004). Like its parent document, both the ISA 84.01 and IEC 61511 utilize benchmarking Safety Integrity Levels to establish the necessary performance of a system. A risk analysis is performed on a process hazard to identify the required safety functions and risk reductions for certain specified process events, where the design of the control systems are to meet the required SIL (ISA 84.01, 2004).

11 Appendix 2 – Optimized Design – Option 2 Calculations

For this Option-2, it is the aim to design a system that maintains the protection requirements set by the risk analysis, and associated NFPA requirements, and further reduces the risk. The aim for this design is to achieve a SIL-3 system availability to exceed the hypothetical requirements of SIL-2.

At this point in the new process, a calculation would be performed on various types of protection equipment to deliver reduced pressures via venting or isolation. In a real-world design, the engineer would calculate based on the equipment available and the needs for the system. From the results of the residual risk calculations, the design engineer would calculate for the following options: reduced Pred using available suppression and detection on the cyclone; reduced Pred using venting with available vent panels for the cyclone; active isolation barriers using the available equipment for the cyclone-baghouse and cyclone-silo interconnections; reduced Pred using available suppression and detection on the hammermill. With the exception of the interconnection calculations (seen in the appendix) the venting and suppression options are provided in the equipment calculation section, and are the designs from which this demonstration will select.

Failed Design Option 2

Providing an optimized solution is an iterative process; the baseline system is further modified, a residual risk calculation is performed, and the total residual risk is checked the required SIL for system acceptability. Provided in section 5.4.6.2 is a detailed description of successfully achieving a code satisfying system that meets the risk level of SIL-3. The design engineer may test and trial multiple potentially viable solutions until one is met. Provided below is a brief description of design changes that did not reduce the risk enough to achieve a SIL-3 classification.

After analyzing the residual risk calculations, it is clear that the cyclone needs to have its Pred reduced to give the system less of a likelihood of exceeding the Ps value. Even with the most aggressive venting option (option 3) the overall cyclone residual risk did not move above SIL-3. To address this, the upstream hammermill Pred value is reduced by utilizing a different delivery and detection combination calculated in the equipment calculations, which moved the residual risk of the cyclone to achieve SIL-3.

However, the downstream baghouse and silo were governing the risk level and inhibiting the protection system from achieving SIL-3. Therefore it was necessary to change the design of Baghouse-Silo interconnection by removing the passive rotary valve and replacing it with an active suppressant isolation barrier. The following section provides a detailed description of the design that successfully achieves SIL-3 for the total system residual risk.

Successful Design Option 2

In this option we are going to modify the design to bring the cyclone's risk position to a SIL-3 rating. To do this, the first protection change is to modify the hammermill by providing a Pred reduction by utilizing the suppression-2 delivery paired with the high-sensitivity 35mbar pressure detector. Additionally, the cyclone vent size was increased to provide a Pred of 0.15bar by using a larger sized venting panel as noted in the calculations sections. For the interconnections, the isolation detection was changed to a 35mBar pressure and flame detector to allow for a greater ratio between the time to detection and time to barrier establishment. Finally, the rotary valve was removed in the baghouse-silo connection and replaced with an isolation suppressor paired with a 52mbar pressure detector. Below is an itemized list of the protection strategy as well as a diagram for ease of understanding.

<u>Vessel Protection</u>

> **Hammermill:** (2) Suppression-2 with (1) 35 mbar pressure detector *[Pred = 0.216 bar]*
> **Cyclone:** (1) 27.5" x 44.3" Vent-1 *[Pred = 0.15bar]*
> **Baghouse:** (3) 44"x44" Vent-1 *[Pred = 0.17bar]*
> **Silo:** (3) 36"x44" Vent-1 *[Pred = 0.375 bar]*

<u>Interconnection Protection</u>

> **Hammermill – Cyclone:** (1) Suppression-1 with (1) 35 mBar pressure & flame detector
> **Cyclone – Baghouse:** (1) Suppression-1 with (1) 52 mBar pressure detector

Cyclone – Silo: (1) Suppression-1 with (1) 52 mBar pressure detector
Baghouse – Silo: (1) Suppression-1 with (1) 52 mBar pressure detector

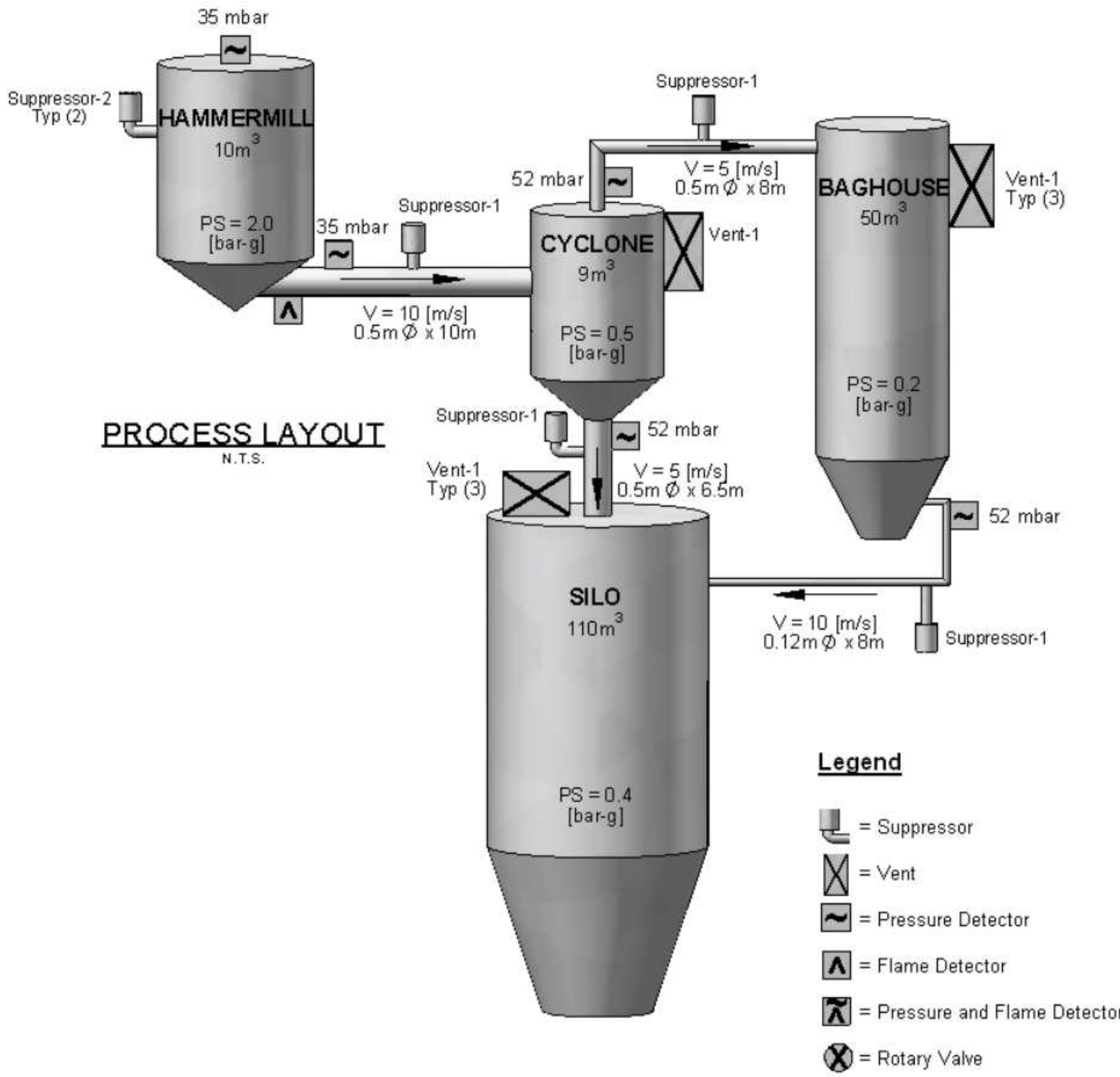


Figure 17: Graphical Protection Scheme for Optimized Design Option 2

Vessel Input

Provided in the following sections are the component-by-component and interconnections inputs that are used to calculate the residual risk of the system. The references for each item can be seen in the table 31 above. For an example on how to hand-calculate a system; refer to the example calculation to understand the interactions of the equations. The inputs and results are given in table format for convenience.

Hammermill Vessel (V_1):

Item	Value or Quantity
π_1	0
π_2	2
π_3	0
π_4	1
$P_s(i)$	2.0
$P_{red}(i,j)$	0.216

Cyclone Vessel (V_2):

Item	Value or Quantity
π_1	0
π_2	0
π_3	1
π_4	0
$P_s(i)$	0.5
$P_{red}(i,j)$	0.15

Baghouse Vessel (V_3):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.2
$P_{red}(i,j)$	0.17

Silo Vessel (V_4):

Item	Value or Quantity
π_1	0
π_2	0
π_3	3
π_4	0
$P_s(i)$	0.4
$P_{red}(i,j)$	0.375

Inter-Connections Input

Hammermill – Cyclone (V_1-V_2):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	3.2
$D_{max(i)}$	8.2
$D_{min(j)}$	2.1
$D_{installed(i)}$	5.0

Cyclone – Baghouse (V_2-V_3):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	3.1
$D_{max(i)}$	8.1
$D_{min(j)}$	2.1
$D_{installed(i)}$	5.0

Cyclone – Silo (V_2-V_4):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	2.5
$D_{max(i)}$	7.5
$D_{min(j)}$	2.0
$D_{installed(i)}$	2.5

Baghouse – Silo (V_3-V_4):

Item	Value or Quantity
π_1	1
π_2	0
π_4	1
π_6	0
$D_{min(i)}$	3.5
$D_{max(i)}$	7.5
$D_{min(j)}$	4.0
$D_{installed(i)}$	3.5

Option-2 Residual Risk Calculation Results

Per Vertex Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	3.14E-04	3185	99.969%	SIL-3
Hammermill	1.47E-04	6803	99.985%	SIL-3
Cyclone	1.42E-05	70423	99.999%	SIL-4
Baghouse	4.83E-05	20704	99.995%	SIL-4
Total Risk	**5.24E-04**	**1910**	**99.948%**	**SIL-3**

Table 42: SIL-1 Per Vertex Residual risk

Per Ignition Risk				
Process	**Residual Risk**	**Failure One In**	**Availability**	**SIL**
Silo	4.40E-05	22727	99.996%	SIL-4

Hammermill	1.30E-04	7692	99.987%	SIL-3
Cyclone	2.94E-04	3401	99.971%	SIL-3
Baghouse	5.43E-05	18416	99.995%	SIL-4
Total Risk	**5.22E-04**	**1915**	**99.948%**	**SIL-3**

Table 43: SIL-1 Per Ignition Residual risk

Option-2 System Analysis

The total calculated risk for the optimized solution – option 2 above is 5.22 x 10^{-4}, which is the addition of all the process components individual risk via the residual risk calculations. This number corresponds to a failure rate of 1-in-1915 events, where if the system were to be called on 1915 times, it would be likely that there would be an explosion mitigation failure in one of those instances. When corresponding this to Safety Integrity Levels, the overall availability of the system, which is equal to one-minus-residual risk, is just over 99.94% system availability. On a purely quantitative level, this would conclude that the system corresponds directly to a SIL-3 rating, which exceeds the requirement of the PRA.

This system not only meets the requirements of the minimum governing standards and documents, but it also provides risk mitigation to levels at two orders of magnitude when compared to the baseline design. In this case study system, there are many viable options which can provide risk levels of SIL-3 while maintaining NFPA system requirements; this system is just one representation in achieving lower residual risk than the minimum requirements. Without the ability to calculate the residual risk of the system there is no way to determine how much risk, and thus the benefit, is associated with each design approach. As the design engineer or manufacturer, one would have the ability to give the end user or process owner multiple options where he/she could select based on cost-benefit parameters.

12 Appendix 3 – Calculations

Hammermill Calculations

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	10.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	**********
Quantity	2
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	18.7
Head Diameter, mm	75.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.143 bar(a)

+ 0.093 bar contribution from suppressor(s)

Suppressant Delivery Curve

Engineer: AT

Time 19:21

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	10.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	18.7
Head Diameter, mm	75.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.192 bar(a)

\+ 0.093 bar contribution from suppressor(s)

Suppressant Delivery Curve

Engineer: AT

Time 19:20

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	10.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

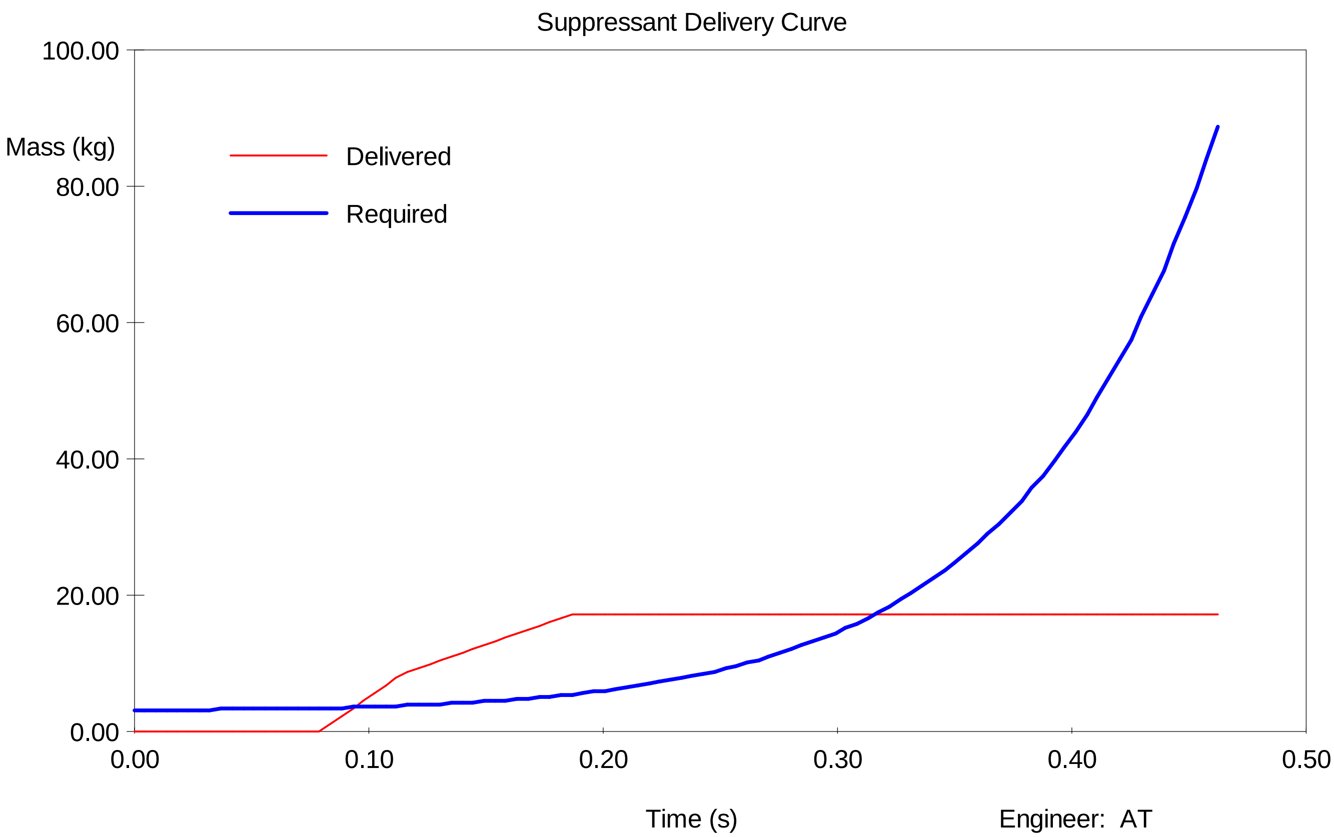
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	13.5
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.111 bar(a)

+ 0.105 bar contribution from suppressor(s)



Engineer: AT

Time 19:23

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	10.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition			
Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

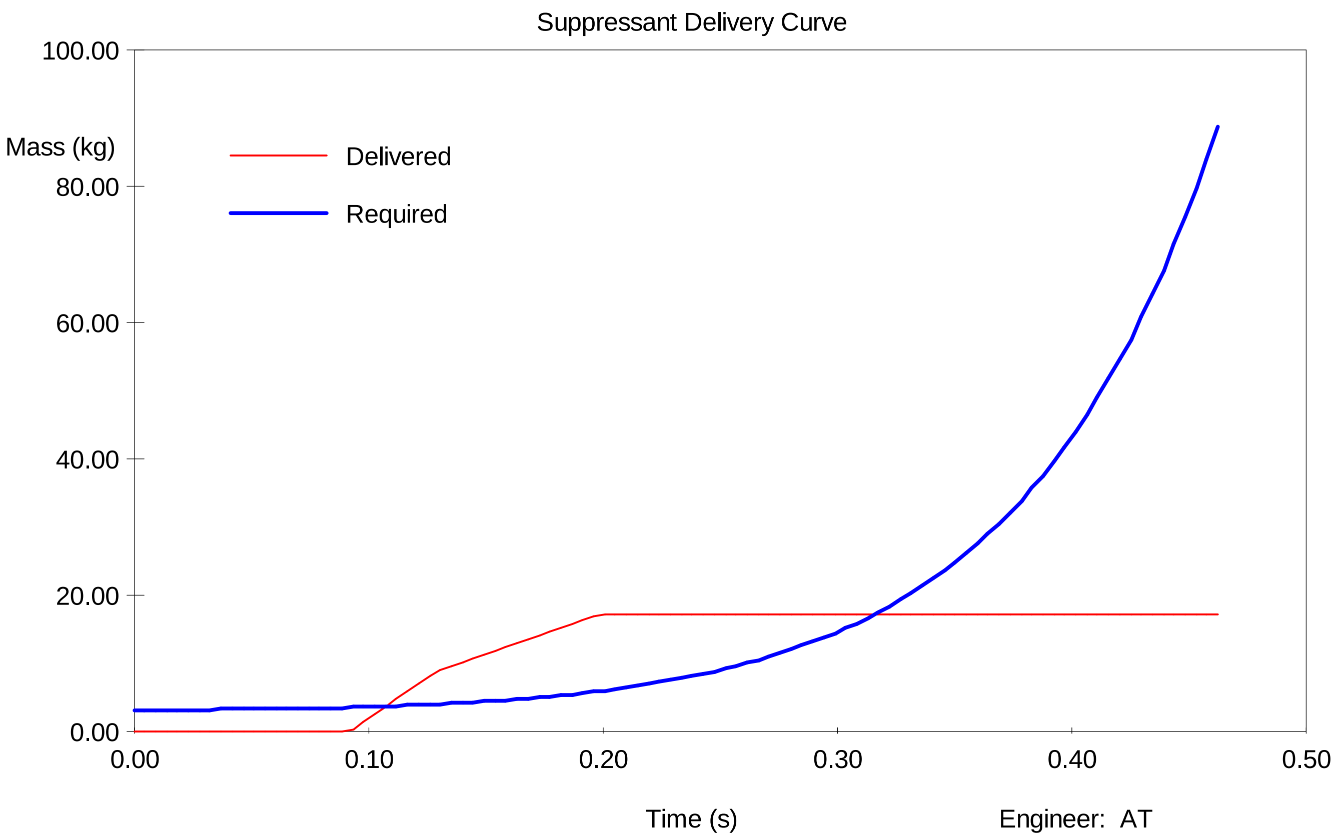
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	13.5
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.149 bar(a)

+ 0.105 bar contribution from suppressor(s)



Engineer: AT

Time 19:25

Date: October 21 2009

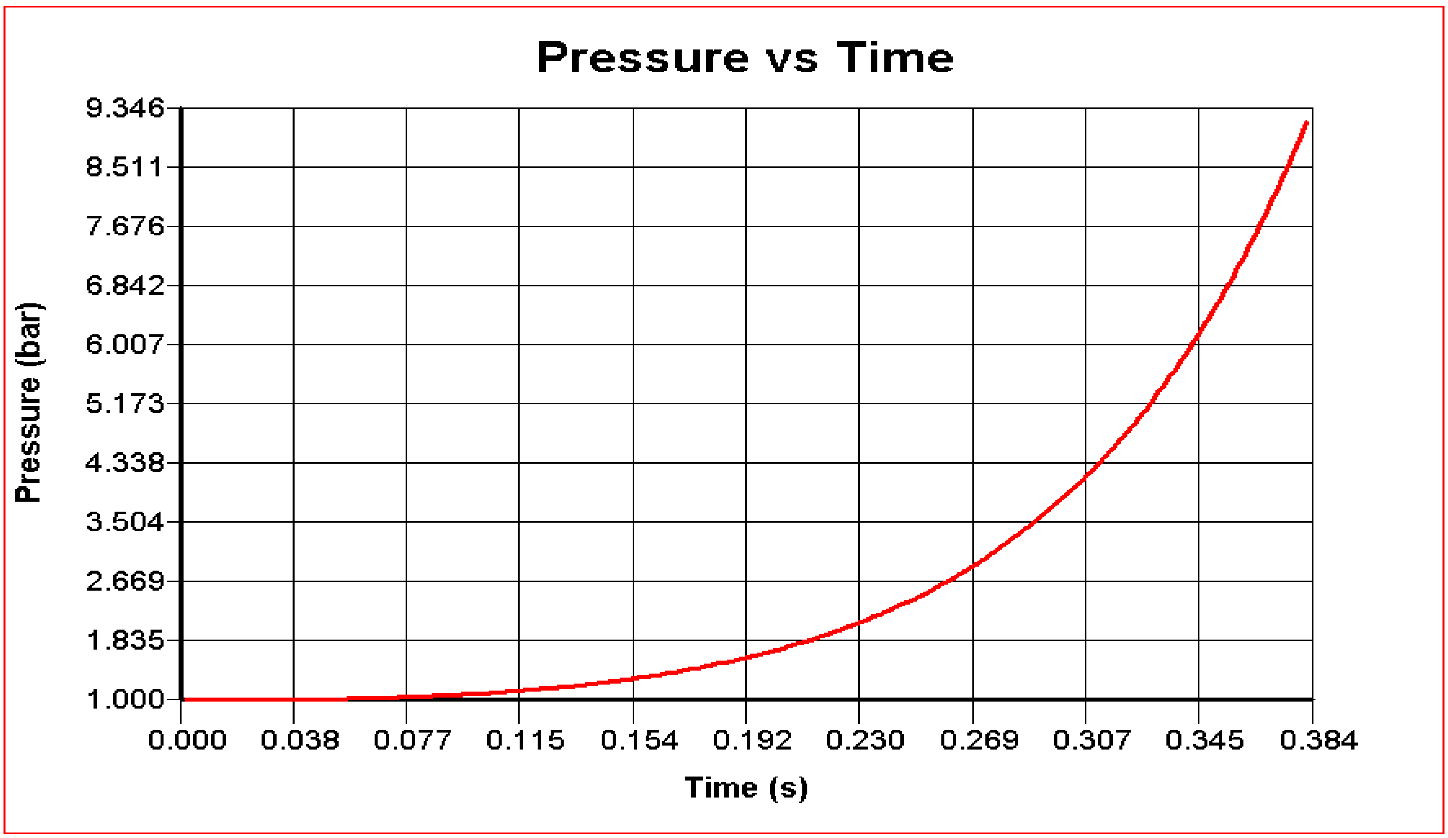
Industrial Explosion Protection Model (Nagy Output) v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	10.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	



Engineer: AT

Time 19:17

Date: October 21 2009

Cylindrical Vessels with Conical Hopper Extension

Calculate L/D for Bottom-up Flame Propogation

Volume above Top of Vent (not included in Effective Volume for L/D)

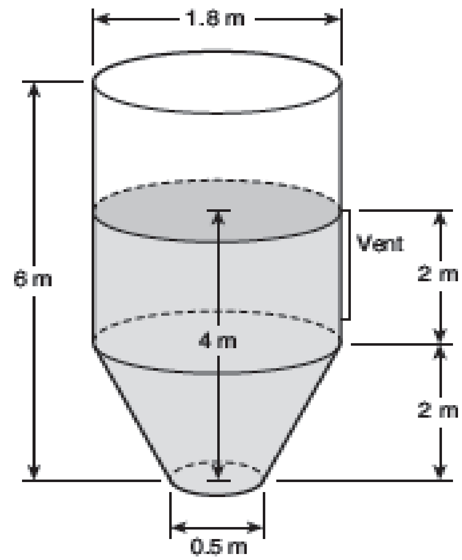
Length	0.1	meters	The distance from the top of the vessel to the top of the vent.
Diam 1	1.92	meters	Diameter of larger cylindrical cross-section
Volume	0.289529	cubic meters	Volume of Cylindrical Section

Length	2.9	meters	The distance from the top of the Conical Hopper to the opposite end of the vent.
Diam 1	1.92	meters	Diameter of larger cylindrical cross-section
Volume	8.396346	cubic meters	Volume of Cylindrical Section

Height h	1	meters	The distance from the top to the bottom of the the Conical Hopper
Diam 2	0.5	meters	Diameter at bottom of Conical Hopper
Volume	1.281875	cubic meters	Volume of Cylindrical Hopper Section

9.96775

$$V = \pi \cdot (h) \frac{\left[(D_1)^2 + (D_1 \cdot D_2) + (D_2)^2\right]}{12}$$



The effective area, A_{eff}, shall be determined by dividing V_{eff} by H (based on the longest central axis flame length). With only one vent, enter the longest distance from one end of the vessel to the opposite end of the vent.

H	3.9	meters
Veff	9.678221	cubic meters
Aeff	2.481595	sq meters

The effective hydraulic diameter, D_{he}, for the enclosure shall be determined based upon the general shape of the enclosure taken normal to the central axis.

$D_{he} = 4 * A_{eff} / p$, Where p is the perimeter of the general shape above the hopper

Dhe	1.777545	meters
L/D	2.194038	**Top-Down**

L/D can not be less than 1, by definition

NFPA 68-2007 Dust in Equipment
Hammermill Option 1

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	3.9	meters	
Volume (V)	10	cubic meters	(This is total volume, not Veff)
Area (Aeff)	2.482	square meters	
Diameter (Dhe)	1.77	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	1.911905	bar			
Pmax	8	bar			
Pstat	0.2	bar	US	0.5	psig
Π	0.238988	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{\max}}{P_{red}} - 1}$$

Av0= 0.151581 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.20339 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2})\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 0.152435 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)
		Av2/Av1=	0

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	7.8	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air}\cdot L}{V}$$

$$A_{v2} = \left[1+\frac{Max\cdot(v_{axial},v_{tan})-20}{36}\cdot 0.7\right]\cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2\cdot(3.14)\cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 0.152435 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 1 number of panels

$$M_T = \left[6.67\cdot\left(P_{red}^{0.2}\right)\cdot\left(n^{0.3}\right)\cdot\left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	24.03143	kg/m^2	
MT	24.03143	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M	19	kg/m^2	US	44	lb/sq ft	If panel density is in US units, enter here and enter metric units at left
			Metric =	215.3	kg/m^2	If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3} = \left[1+\frac{(0.0075)\cdot M^{0.6}\cdot K_{St}^{0.5}}{n^{0.3}\cdot V\cdot P_{red}^{0.2}}\right]\cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 0.152435 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs}c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss}Vc_w} + \frac{M_e}{Vc_w}$$

where:

$\overline{X}_r$ = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
$\overline{H}$ = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	3.9	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	10	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	9.620513	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}$$

Av4= 0.152435 sq meters 236 sq inches

NFPA 68-2007 Dust in Equipment
Hammermill Option 2

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	3.9	meters	
Volume (V)	10	cubic meters	(This is total volume, not Veff)
Area (Aeff)	2.482	square meters	
Diameter (Dhe)	1.77	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	1.191819	bar			
Pmax	8	bar			
Pstat	0.2	bar	US	0.5	psig
Π	0.148977	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{max}}{P_{red}} - 1}$$

Av0= 0.203023 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.20339 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2})\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 0.212593 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)

Av2/Av1= 0

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	7.8	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 0.212593 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass < 40 kg/m^2

n 1 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	20.52221	kg/m^2	
MT	20.52221	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2 US 44 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 215.3 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 0.212593 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
H = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	3.9	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	10	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	9.620513	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4= 0.212593 sq meters 330 sq inches

NFPA 68-2007 Dust in Equipment
Hammermill Option 3

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	3.9	meters	
Volume (V)	10	cubic meters	(This is total volume, not Veff)
Area (Aeff)	2.482	square meters	
Diameter (Dhe)	1.77	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	1.01	bar			
Pmax	8	bar			
Pstat	0.2	bar	US	0.5	psig
Π	0.12625	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{\ 4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{\max}}{P_{red}} - 1}$$

Av0= 0.223467 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.20339 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}\left(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{\ 2}\right)\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 0.238875 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)

Av2/Av1= 0

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	7.8	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 0.238875 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass $\leq$ 40 kg/m^2

n 1 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	19.41839	kg/m^2	
MT	19.41839	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density
M 19 kg/m^2 US 44 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 215.3 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 0.238875 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction
Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
H = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	3.9	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	10	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	9.620513	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4=	0.238875	sq meters	370	sq inches

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	10.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.3	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	1.9	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	1	
	Min. Distance - Vessel to Barrier	d_{min}	=	3.2	m
	Max. Distance - Vessel to Barrier	d_{max}	=	8.2	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.29	bar(g)
	Detection Time	t_a	=	246	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	219	ms
	Flame Velocity at Barrier	V_{fb}	=	86.2	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	104.9	m/s

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	10.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.3	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	1.9	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Flame**				
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	1	
	Min. Distance - Vessel to Barrier	d_{min}	=	1.3	m
	Max. Distance - Vessel to Barrier	d_{max}	=	6.3	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.29	bar(g)
	Detection Time	t_a	=	82	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	82	ms
	Flame Velocity at Barrier	V_{fb}	=	102.9	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	102.9	m/s

Adam Tracy / Hammermill Cylcone.sis / 12/18/09

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	10.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	1.9	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Flame**				
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	1	
	Min. Distance - Vessel to Barrier	d_{min}	=	1.3	m
	Max. Distance - Vessel to Barrier	d_{max}	=	6.3	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.22	bar(g)
	Detection Time	t_a	=	82	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	82	ms
	Flame Velocity at Barrier	V_{fb}	=	102.9	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	102.9	m/s

Cyclone Calculations

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	9.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

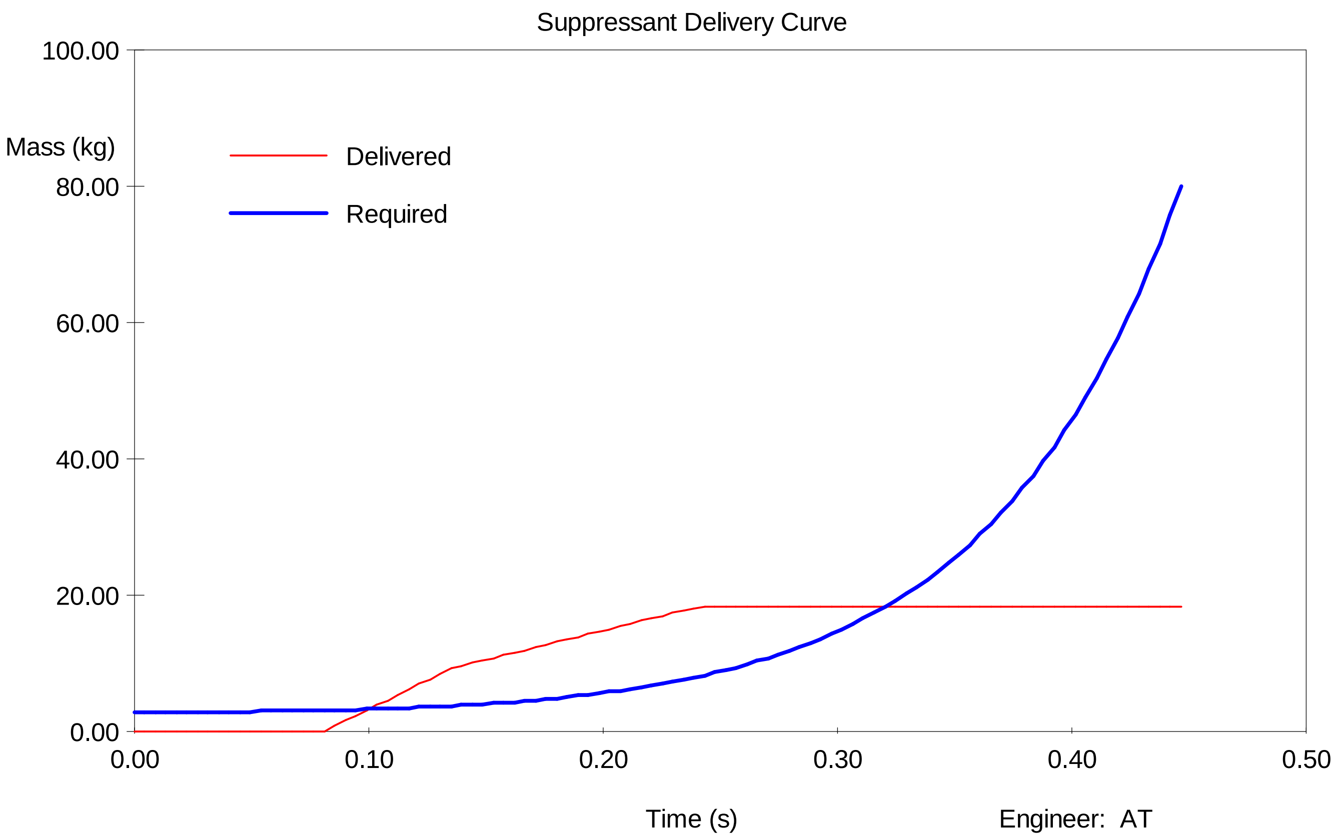
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	18.7
Head Diameter, mm	75.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.139 bar(a)

+ 0.104 bar contribution from suppressor(s)



Engineer: AT

Time 19:34

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	9.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

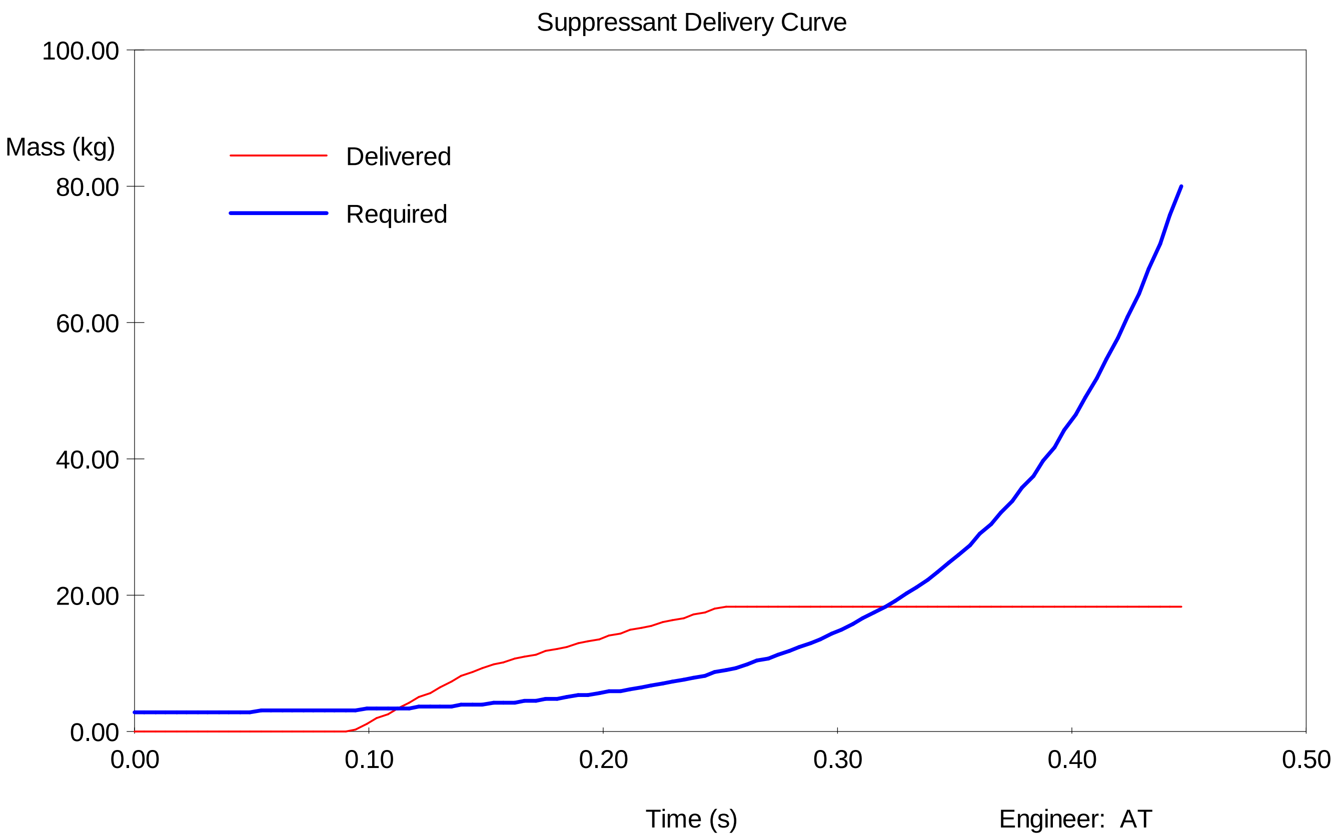
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	18.7
Head Diameter, mm	75.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.189 bar(a)

\+ 0.104 bar contribution from suppressor(s)



Engineer: AT

Time 19:33

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	9.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

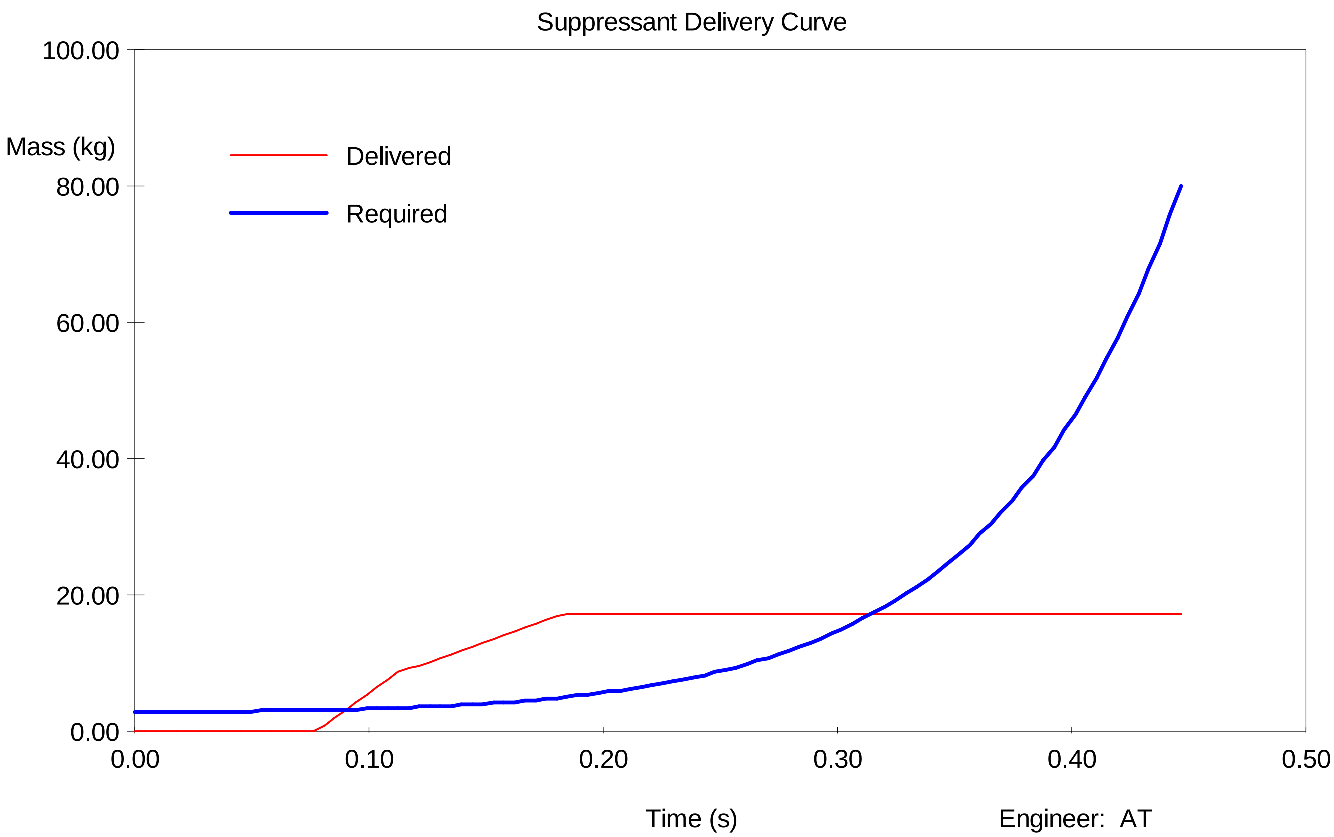
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	13.5
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.108 bar(a)

+ 0.117 bar contribution from suppressor(s)



Engineer: AT

Time 19:31

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	9.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

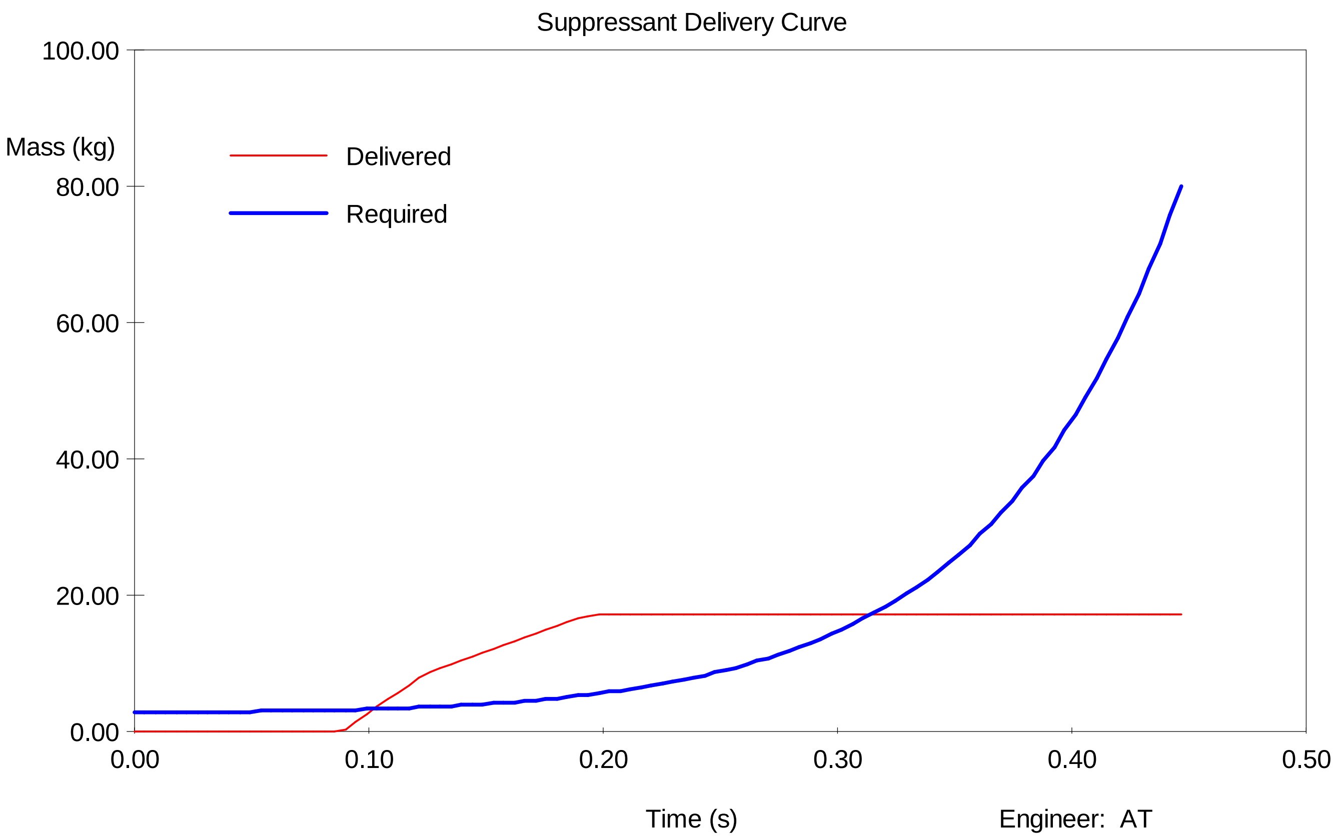
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	13.5
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.146 bar(a)

+ 0.117 bar contribution from suppressor(s)



Engineer: AT

Time 19:32

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	9.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

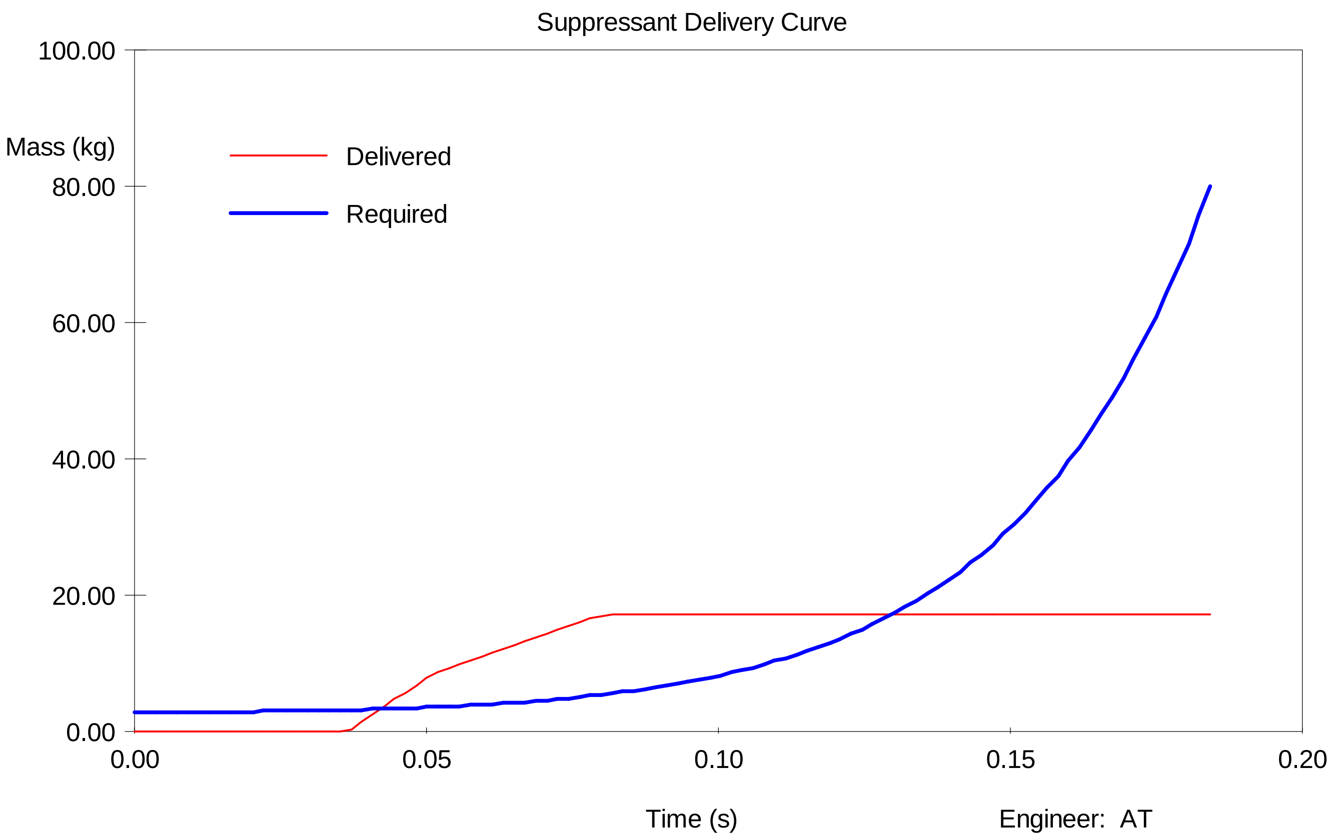
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	2
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	13.5
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.146 bar(a)

\+ 0.117 bar contribution from suppressor(s)



Engineer: AT

Time 19:49

Date: October 21 2009

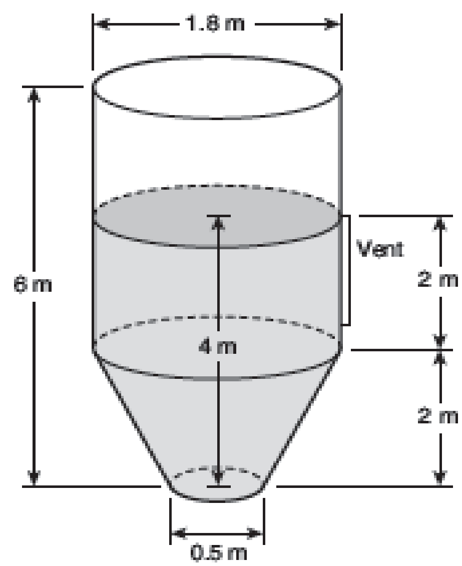
Cylindrical Vessels with Conical Hopper Extension

Calculate L/D for Bottom-up Flame Propogation

Volume above Top of Vent (not included in Effective Volume for L/D)

Length	0.1	meters	The distance from the top of the vessel to the top of the vent.
Diam 1	1.822572	meters	Diameter of larger cylindrical cross-section
Volume	0.260891	cubic meters	Volume of Cylindrical Section
Length	2.9	meters	The distance from the top of the Conical Hopper to the opposite end of the vent.
Diam 1	1.822572	meters	Diameter of larger cylindrical cross-section
Volume	7.565843	cubic meters	Volume of Cylindrical Section
Height h	1	meters	The distance from the top to the bottom of the the Conical Hopper
Diam 2	0.5	meters	Diameter at bottom of Conical Hopper
Volume	1.173661	cubic meters	Volume of Cylindrical Hopper Section
	9.000395		

$$V = \pi \cdot (h) \frac{\left[(D_1)^2 + (D_1 \cdot D_2) + (D_2)^2\right]}{12}$$



The effective area, A_{eff}, shall be determined by dividing V_{eff} by H (based on the longest central axis flame length). With only one vent, enter the longest distance from one end of the vessel to the opposite end of the vent.

H	3.9	meters
Veff	8.739504	cubic meters
Aeff	2.240898	sq meters

The effective hydraulic diameter, D_{he}, for the enclosure shall be determined based upon the general shape of the enclosure taken normal to the central axis.

$D_{he} = 4 * A_{eff} / p$, Where p is the perimeter of the general shape above the hopper

Dhe	1.689142	meters	
L/D	2.308865	**Top-Down**	L/D can not be less than 1, by definition

NFPA 68-2007 Dust in Equipment

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	3.9	meters	
Volume (V)	9	cubic meters	(This is total volume, not Veff)
Area (Aeff)	2.24	square meters	
Diameter (Dhe)	1.69	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.19	bar			
Pmax	8	bar			
Pstat	0.14	bar	US	0.5	psig
Π	0.02375	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{max}}{P_{red}} - 1}$$

Av0= 0.474161 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.307692 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2})\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 0.587733 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building? N YES/NO Correction factor of 1.7 if a building (occupiable)

Av2/Av1= 0

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	8.666667	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 0.587733 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 1 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	9.320914	kg/m^2	
MT	9.320914	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2 US 0.75 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}$$

OSECO PANEL MASSES	
CRP	13.4 kg/m^2
CRV	13.4 kg/m^2
CRVC	19 kg/m^2
RNDCC	15.5 kg/m^2
MVC	19 kg/m^2
GLV	7.2 kg/m^2

Av3 0.632929 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction
Calculate the worst-case building partial volume fraction, Xr, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction

$\overline{M}_f$ = average mass (gram) of floor samples

A_{fs} = measured floor areas

c_w = worst-case dust concentration

H = ceiling height of the building

$\overline{M}_s$ = average mass (gram) of surface samples

A_{sur} = total area of surfaces with dust deposits

A_{ss} = measured sample areas of surfaces with dust deposits

V = building volume

M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	3.9	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	9	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	10.59829	fill fraction

If Xr is less then Π, then no venting is required

If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4= 0.632929 sq meters 981 sq inches

NFPA 68-2007 Dust in Equipment
Silo Option 2

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	3.9	meters	
Volume (V)	9	cubic meters	(This is total volume, not Veff)
Area (Aeff)	2.24	square meters	
Diameter (Dhe)	1.69	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.15	bar			
Pmax	8	bar			
Pstat	0.14	bar	US	0.5	psig
Π	0.01875	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}{}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{\max}}{P_{red}} - 1}$$

Av0= 0.535016 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.307692 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}(-0.95 \cdot \mathbf{P}_{\mathbf{red}}{}^{2})\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 0.66483 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)
			Av2/Av1= 0

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	8.666667	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 0.66483 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 1 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	8.613294	kg/m^2	
MT	8.613294	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M	19	kg/m^2	US	0.75	lb/sq ft	If panel density is in US units, enter here and enter metric units at left
			Metric =	3.7	kg/m^2	If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 0.71843 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
$\overline{H}$ = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO		
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	3.9	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	9	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	10.59829	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4= 0.71843 sq meters 1114 sq inches

NFPA 68-2007 Dust in Equipment
Cyclone Option 3

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	3.9	meters	
Volume (V)	9	cubic meters	(This is total volume, not Veff)
Area (Aeff)	2.24	square meters	
Diameter (Dhe)	1.69	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.13	bar			
Pmax	8	bar			
Pstat	0.14	bar	US	0.5	psig
Π	0.01625	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{\max}}{P_{red}} - 1}$$

Av0= 0.57543 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.307692 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}(\ -0.95 \cdot \mathbf{P}_{\mathbf{red}}^{\ 2})\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 0.715796 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)
			Av2/Av1= 0

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	8.666667	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 0.715796 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 1 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	8.211299	kg/m^2	
MT	8.211299	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M	19	kg/m^2	US	0.75	lb/sq ft	If panel density is in US units, enter here and enter metric units at left
			Metric =	3.7	kg/m^2	If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2

MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 0.77518 sq meters — If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

- X_r = worst-case building partial fraction
- $\overline{M}_f$ = average mass (gram) of floor samples
- A_{fs} = measured floor areas
- c_w = worst-case dust concentration
- H = ceiling height of the building
- $\overline{M}_s$ = average mass (gram) of surface samples
- A_{sur} = total area of surfaces with dust deposits
- A_{ss} = measured sample areas of surfaces with dust deposits
- V = building volume
- M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	3.9	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	9	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	10.59829	fill fraction

If Xr is less then Π, then no venting is required

If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4= 0.77518 sq meters 1202 sq inches

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.3	bar(g)
	Initial Pressure	P_{i}	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	2.6	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_{a}	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	3.1	m
	Max. Distance - Vessel to Barrier	d_{max}	=	8.1	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_{b}	=	0.29	bar(g)
	Detection Time	t_{a}	=	237	ms
	Reaction Time of Barrier	t_{b}	=	14	ms
	Time of Entry of Flame into Duct	t_{e}	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	86.4	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	105.4	m/s

Adam Tracy / Cylcone baghouse.sis / 12/18/09

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	2.6	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	3.1	m
	Max. Distance - Vessel to Barrier	d_{max}	=	8.1	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.24	bar(g)
	Detection Time	t_a	=	237	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	86.4	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	105.4	m/s

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	2.6	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	3.1	m
	Max. Distance - Vessel to Barrier	d_{max}	=	8.1	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.19	bar(g)
	Detection Time	t_a	=	237	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	86.4	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	105.4	m/s

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	20.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	2.6	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	3.1	m
	Max. Distance - Vessel to Barrier	d_{max}	=	8.1	m
	Downstream Safety Distance	d_{safety}	=	2.1	m
	Mass of Suppressant Required	M_{supp}	=	5.0	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.15	bar(g)
	Detection Time	t_a	=	237	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	86.4	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	105.4	m/s

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	5.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	1.8	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	2.5	m
	Max. Distance - Vessel to Barrier	d_{max}	=	7.5	m
	Downstream Safety Distance	d_{safety}	=	2.0	m
	Mass of Suppressant Required	M_{supp}	=	4.7	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.24	bar(g)
	Detection Time	t_a	=	237	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	70.9	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	89.5	m/s

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	5.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	1.8	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	2.5	m
	Max. Distance - Vessel to Barrier	d_{max}	=	7.5	m
	Downstream Safety Distance	d_{safety}	=	2.0	m
	Mass of Suppressant Required	M_{supp}	=	4.7	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.19	bar(g)
	Detection Time	t_a	=	237	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	70.9	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	89.5	m/s

Explosion Protection

Equipment:	Source Vessel Volume	V_{vessel}	=	9.0	m3
	Reduced Explosion Pressure	P_{red}	=	0.2	bar(g)
	Initial Pressure	P_i	=	0.00	bar(g)
	Duct Area	A_{duct}	=	0.196	m2
	Duct Diameter	D_{duct}	=	500	mm
	Duct Height	H_{duct}	=		mm
	Duct Width	W_{duct}	=		mm
	Process Air Velocity	V_{air}	=	5.00	m/s
	Air Temperature	T_{air}	=	20	°C
	Source Vessel Diameter	D_{vessel}	=	1.8	m
	Downstream Protection		=	yes	
Fuel:	**Dust**				
	Explosion Overpressure	P_{max}	=	8.0	bar(g)
	Explosion Rate Constant	K_{max}	=	128	bar·m/s
Sensor:	**Pressure**				
	Detection Pressure	P_a	=	0.04	bar(g)
Barrier:	**Extinguishing**				
	Type of HRD's				
	Opposite	no			
	Elbow	no			
Results:	**calculation for an extinguishing barrier**				
	Number of HRD's	n_{HRD}	=	2	
	Min. Distance - Vessel to Barrier	d_{min}	=	2.5	m
	Max. Distance - Vessel to Barrier	d_{max}	=	7.5	m
	Downstream Safety Distance	d_{safety}	=	2.0	m
	Mass of Suppressant Required	M_{supp}	=	4.7	kg
	Max. Pressure in Duct and at Barrier	P_b	=	0.15	bar(g)
	Detection Time	t_a	=	237	ms
	Reaction Time of Barrier	t_b	=	14	ms
	Time of Entry of Flame into Duct	t_e	=	211	ms
	Flame Velocity at Barrier	V_{fb}	=	70.9	m/s
	Max. Flame Velocity at Barrier	V_{fbm}	=	89.5	m/s

Baghouse Calculations

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	50.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

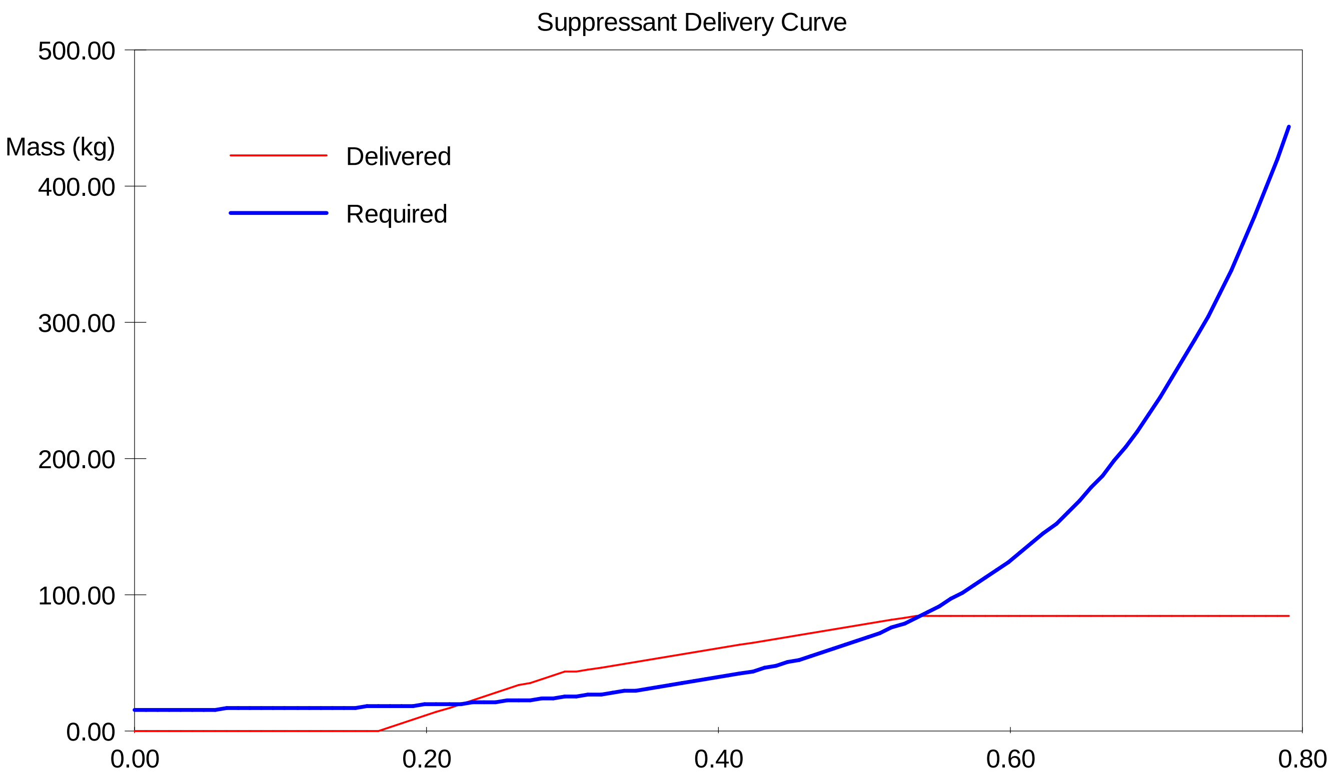
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	4
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	42.5
Head Diameter, mm	75.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.259 bar(a)

+ 0.085 bar contribution from suppressor(s)



Engineer: AT

Time 19:40

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	50.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

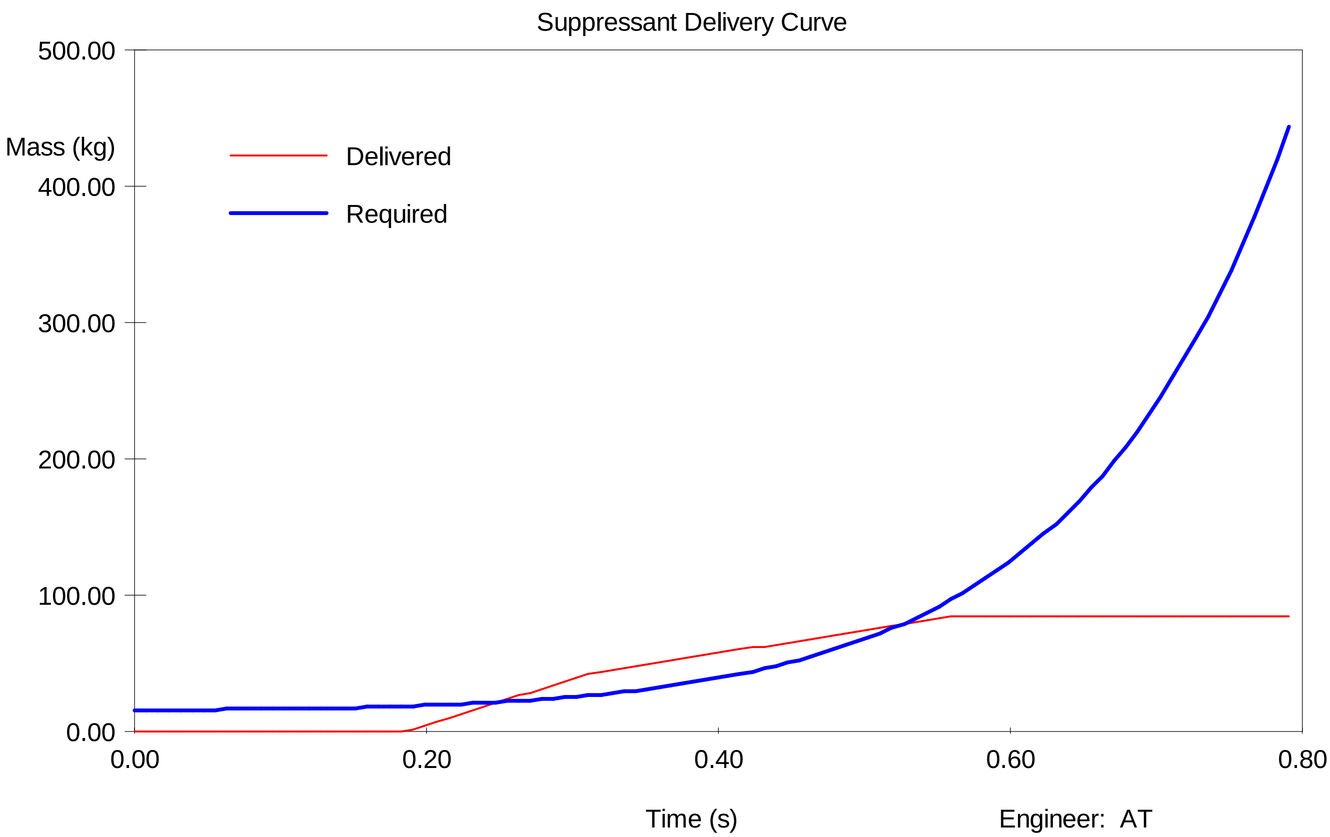
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	4
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	42.5
Head Diameter, mm	75.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.340 bar(a)

+ 0.085 bar contribution from suppressor(s)



Engineer: AT

Time 19:41

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	50.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

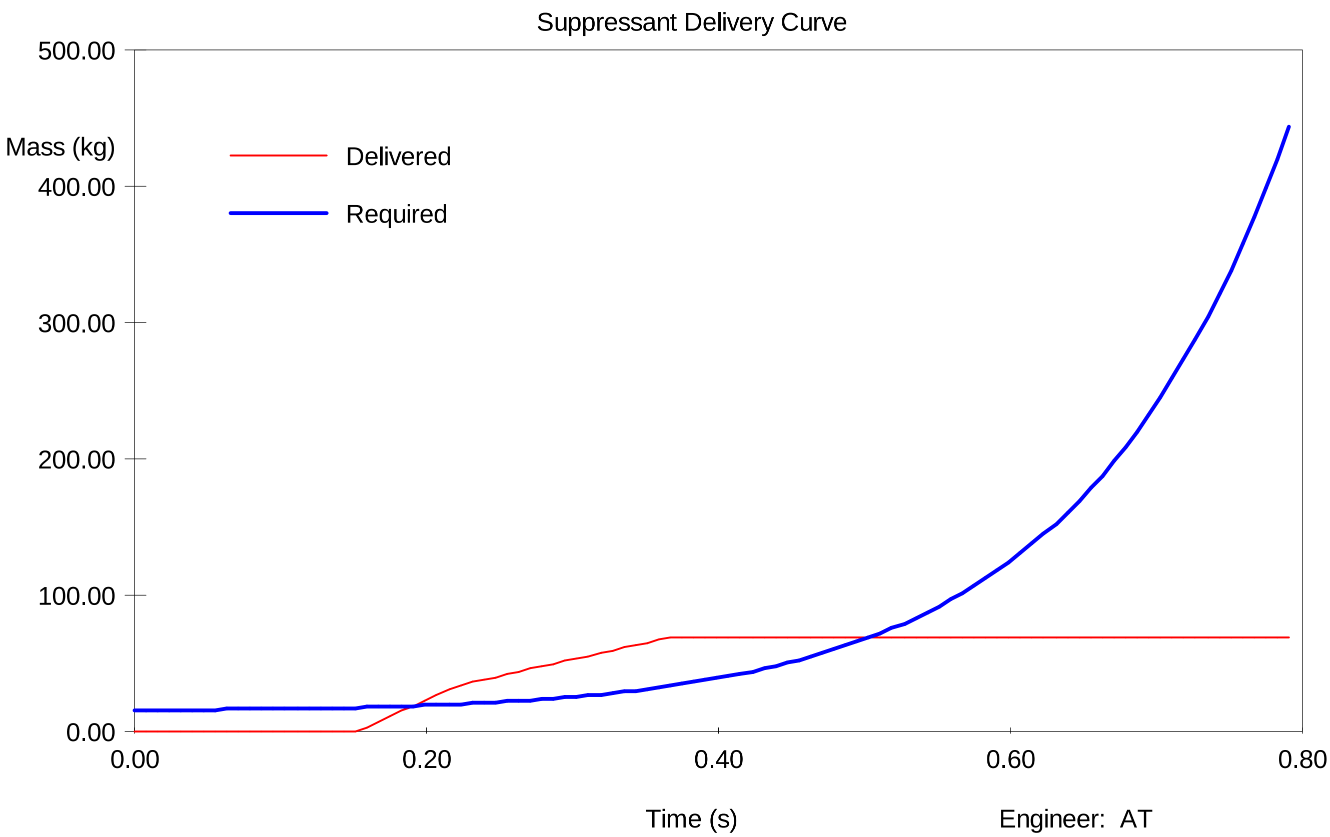
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	4
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	27.0
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.170 bar(a)

+ 0.084 bar contribution from suppressor(s)



Engineer: AT

Time 19:43

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	50.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

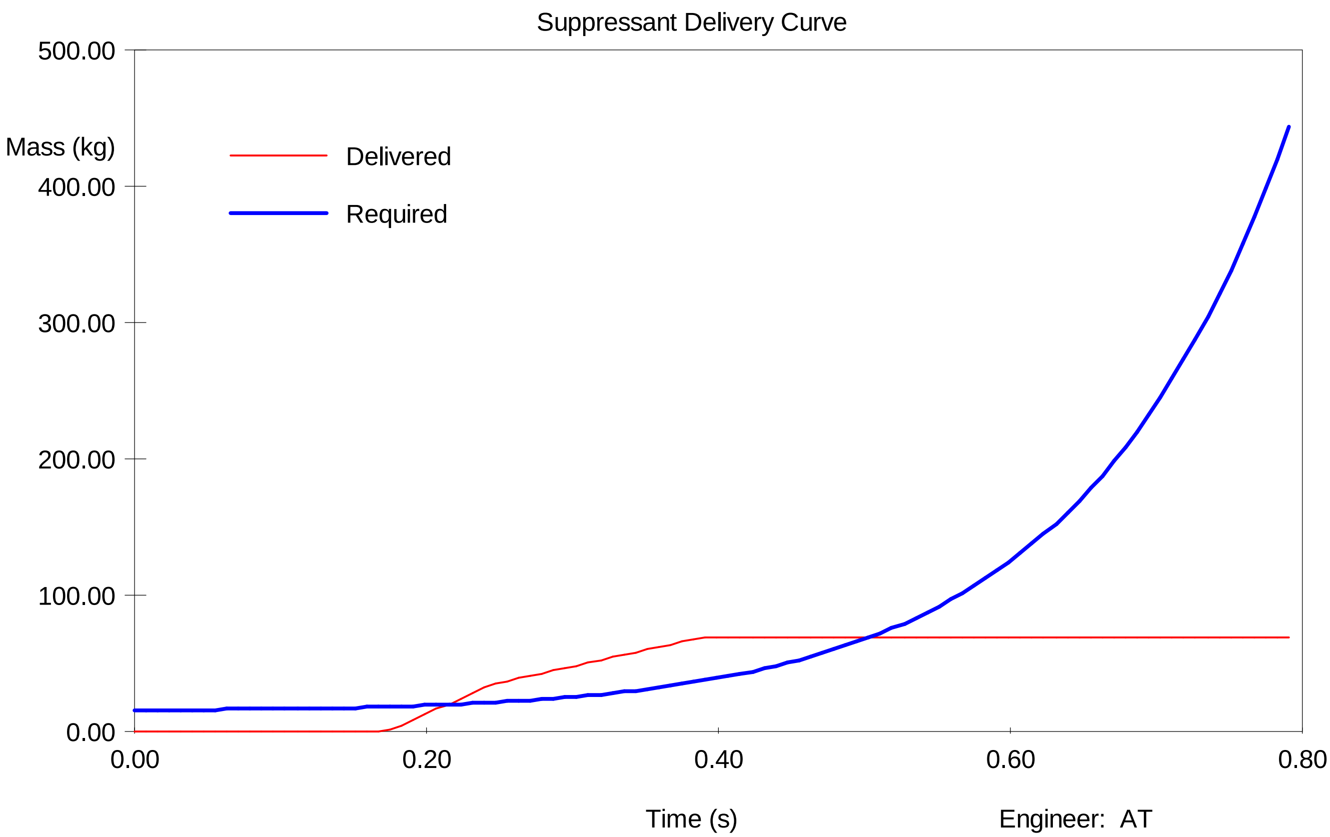
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	4
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	27.0
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.226 bar(a)

+ 0.084 bar contribution from suppressor(s)



Engineer: AT

Time 19:42

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	50.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

Suppressant = Suppressant-X

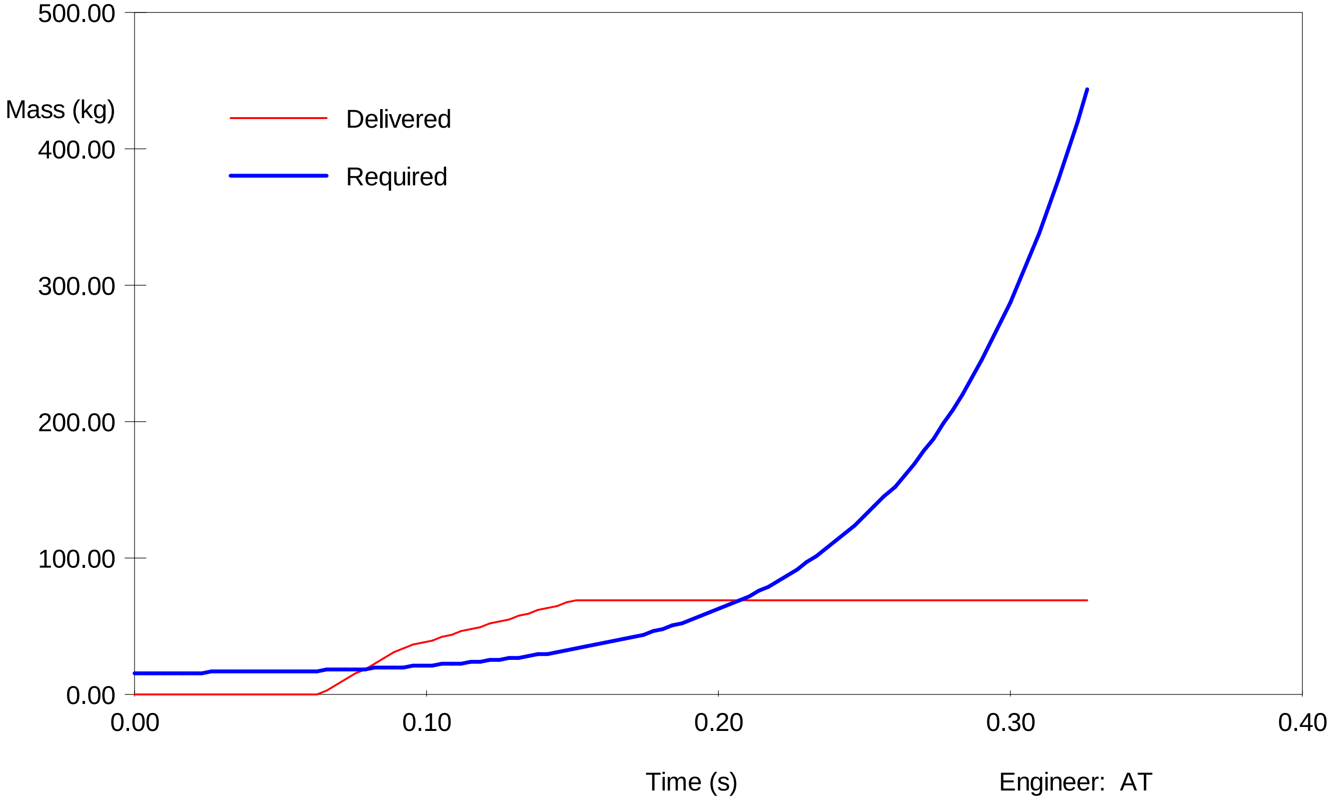
Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	4
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	27.0
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.170 bar(a)

+ 0.084 bar contribution from suppressor(s)

Suppressant Delivery Curve



Engineer: AT

Time 19:50

Date: October 21 2009

Rectangular Vessels with Rectangular Hopper Extension

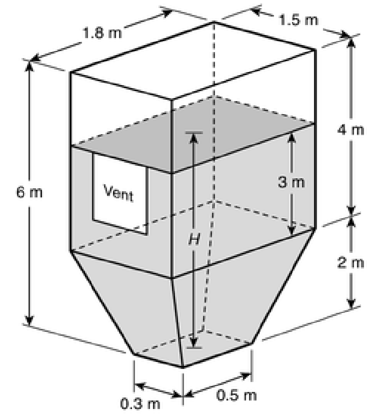
Calculate L/D for Bottom-up Flame Propogation

Volume above Top of Vent (not included in Effective Volume for L/D)

Length	0.1	meters	The distance from the top of the vessel to the top of the vent.
Width a2	2.5	meters	Dimension of straight side
Width b2	2.5	meters	Dimension of opposite straight side
Volume	0.625	cubic meters	Volume of Rectangular Section above the vent

Length	7.5	meters	The distance from the top of the Rectangular Hopper to the opposite end of the vent.
Width a2	2.5	meters	Dimension of straight side
Width b2	2.5	meters	Dimension of opposite straight side
Volume	46.875	cubic meters	Volume of Rectangular Section

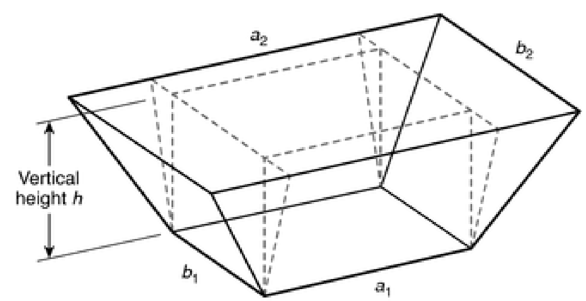
Height h	1	meters	The distance from the top to the bottom of the the Rectangular Hopper
Width a1	0.5	meters	Dimension of Rectangular Hopper bottom (same side as B28)
Width b1	0.5	meters	Dimension of Rectangular Hopper bottom (same side as B29)
Volume	2.583333	cubic meters	Volume of Rectangular Hopper Section



The effective area, A_{eff}, shall be determined by dividing V_{eff} by H (based on the longest central axis flame length). With only one vent, enter the longest distance from one end of the vessel to the opposite end of the vent.

50.08333

H	8.5	meters
Veff	49.45833	cubic meters
Aeff	5.818627	sq meters



The effective hydraulic diameter, D_{he}, for the enclosure shall be determined based upon the general shape of the enclosure taken normal to the central axis.

$D_{he} = 4 * A_{eff} / p$, Where p is the perimeter of the general shape above the hopper

Aspect	1	ratio a/b	This uses the aspect ratio of the above rectangular section
Side a	2.412183	meters	
Side b	2.412183	meters	
p	9.648733	meters	
Dhe	2.412183	meters	

$$V = \frac{(a_1)\cdot(h)\cdot(b_2-b_1)}{2} + \frac{(b_1)\cdot(h)\cdot(a_2-a_1)}{2} + \frac{(h)\cdot(a_2-a_1)\cdot(b_2-b_1)}{3} + (a_1)\cdot(b_1)\cdot h$$

L/D	3.523779	**Bottom-up**	L/D can not be less than 1, by definition

NFPA 68-2007 Dust in Equipment
Baghouse Option 1

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	8.5	meters	
Volume (V)	50	cubic meters	(This is total volume, not Veff)
Area (Aeff)	5.82	square meters	
Diameter (Dhe)	2.412	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.169749	bar			
Pmax	8	bar			
Pstat	0.11	bar	US	0.5	psig
Π	0.021219	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{\max}}{P_{red}} - 1}$$

Av0= 1.767314 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 3.524046 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}\left(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2}\right)\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 3.182538 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building? N YES/NO Correction factor of 1.7 if a building (occupiable)
Av2/Av1= 0

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	3.4	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7 \right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 3.182538 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass $\leq$ 40 kg/m^2

n 3 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}} \right) \right]^{1.67}$$

Mformula	272.7994	kg/m^2	
MT	40	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2

US 0.75 lb/sq ft — If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 — If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}} \right] \cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 3.182538 sq meters — If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction

$\overline{M}_f$ = average mass (gram) of floor samples

A_{fs} = measured floor areas

c_w = worst-case dust concentration

$\underline{H}$ = ceiling height of the building

$\overline{M}_s$ = average mass (gram) of surface samples

A_{sur} = total area of surfaces with dust deposits

A_{ss} = measured sample areas of surfaces with dust deposits

V = building volume

M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO		
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	8.5	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	50	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	2.136471	fill fraction

If Xr is less then Π, then no venting is required

If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4=	3.182538	sq meters	4933	sq inches

NFPA 68-2007 Dust in Equipment
Baghouse Option 2

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	8.5	meters	
Volume (V)	50	cubic meters	(This is total volume, not Veff)
Area (Aeff)	5.82	square meters	
Diameter (Dhe)	2.412	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.142285	bar			
Pmax	8	bar			
Pstat	0.11	bar	US	0.5	psig
Π	0.017786	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{max}}{P_{red}} - 1}$$

Av0= 1.933745 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 3.524046 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}\left(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2}\right)\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 3.494902 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)
			Av2/Av1= 0

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	3.4	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 3.494902 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 3 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	257.183	kg/m^2	
MT	40	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2 US 0.75 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 3.494902 sq meters If M < MT, then there is no area correction for inertia
3.495

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
$\overline{H}$ = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	8.5	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	50	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	2.136471	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}$$

Av4=	3.494902	sq meters	5417	sq inches

NFPA 68-2007 Dust in Equipment
Baghouse Option 3

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	8.5	meters	
Volume (V)	50	cubic meters	(This is total volume, not Veff)
Area (Aeff)	5.82	square meters	
Diameter (Dhe)	2.412	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.115013	bar			
Pmax	8	bar			
Pstat	0.11	bar	US	0.5	psig
Π	0.014377	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{max}}{P_{red}} - 1}$$

Av0= 2.154553 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 3.524046 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2})\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 3.905607 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)
			Av2/Av1= 0
Flow-Created?	N	YES/NO	

$$v = \frac{Q_{air} \cdot L}{}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

36

Inlet Air	20	m^3/sec	
Inlet Pipe Diam	1	m	
Outlet Pipe Diam	1	m	
Vaxial	3.4	meter/sec	
Vtangential	12.7324	meter/sec	(0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO	
Rotational Radius	0.5	meter	
Rotational Speed	1000	RPM	
Vtangential	26.17994	meter/sec	(0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 3.905607 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 3 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	239.5394	kg/m^2	
MT	40	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2 US 0.75 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 3.905607 sq meters If M < MT, then there is no area correction for inertia
3.906

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation:

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

where:

X_r = worst-case building partial fraction
$\bar{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
H = ceiling height of the building
$\bar{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	8.5	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	50	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm

Xr 2.136471 fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}$$

Av4= 3.905607 sq meters 6054 sq inches

Silo Calculations

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	110.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

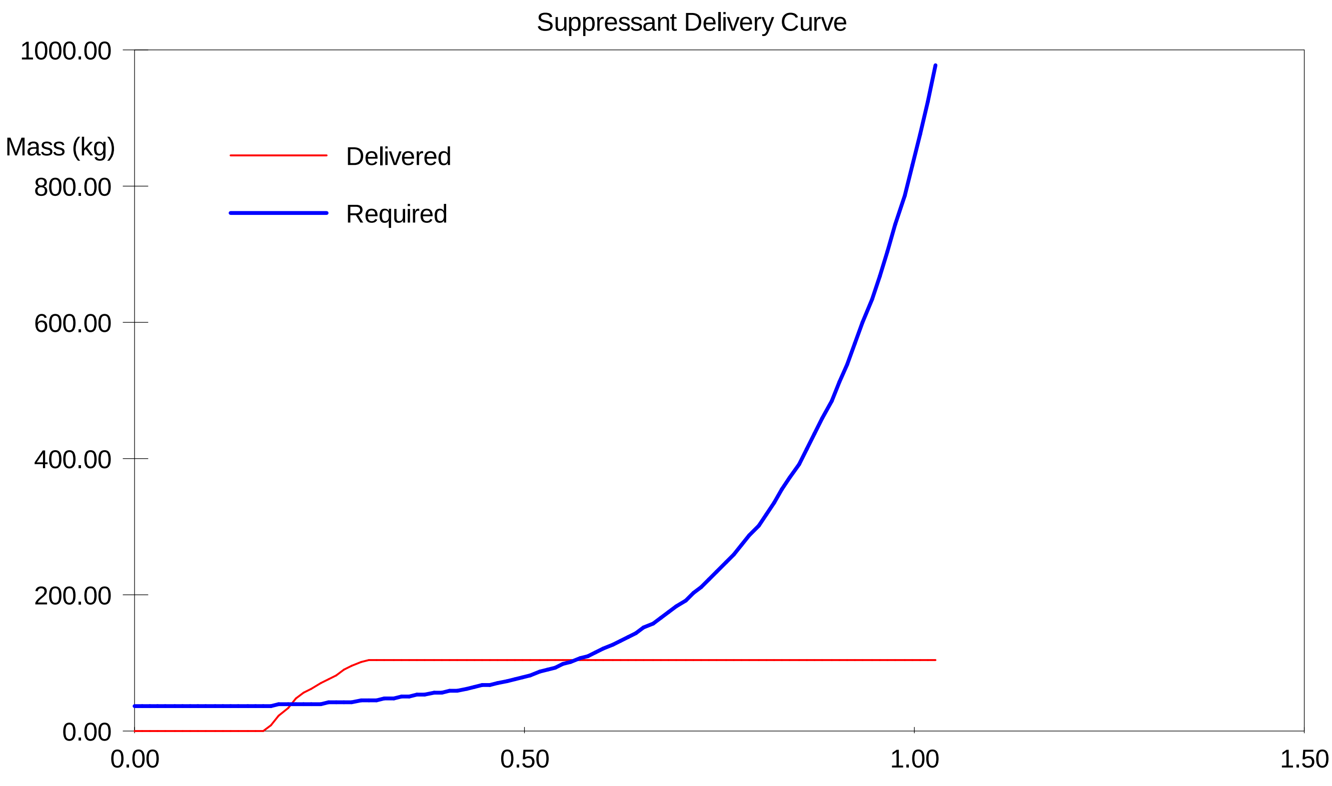
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	5
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	42.5
Head Diameter, mm	125.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.098 bar(a)

+ 0.048 bar contribution from suppressor(s)



Engineer: AT

Time 19:47

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	110.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

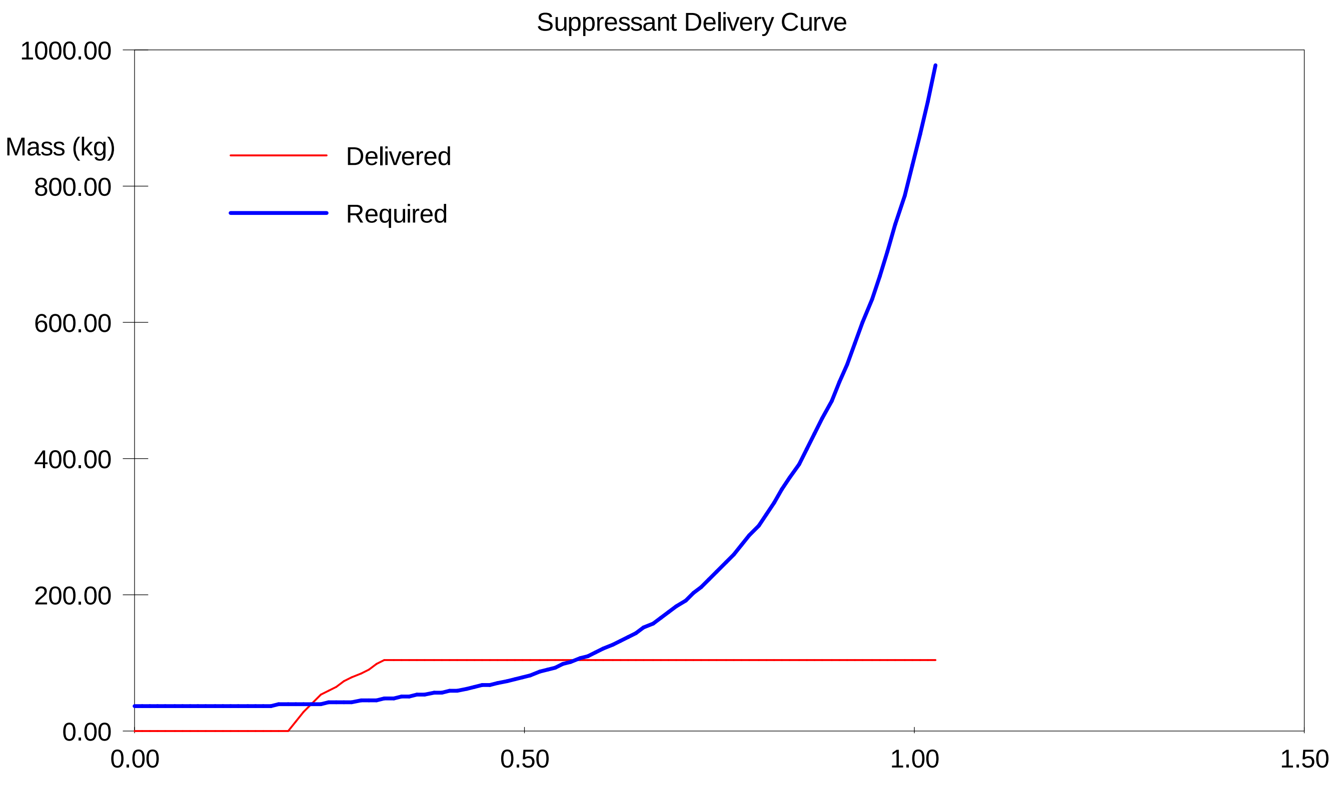
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	5
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	42.5
Head Diameter, mm	125.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.136 bar(a)

+ 0.048 bar contribution from suppressor(s)



Engineer: AT

Time 19:47

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	110.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

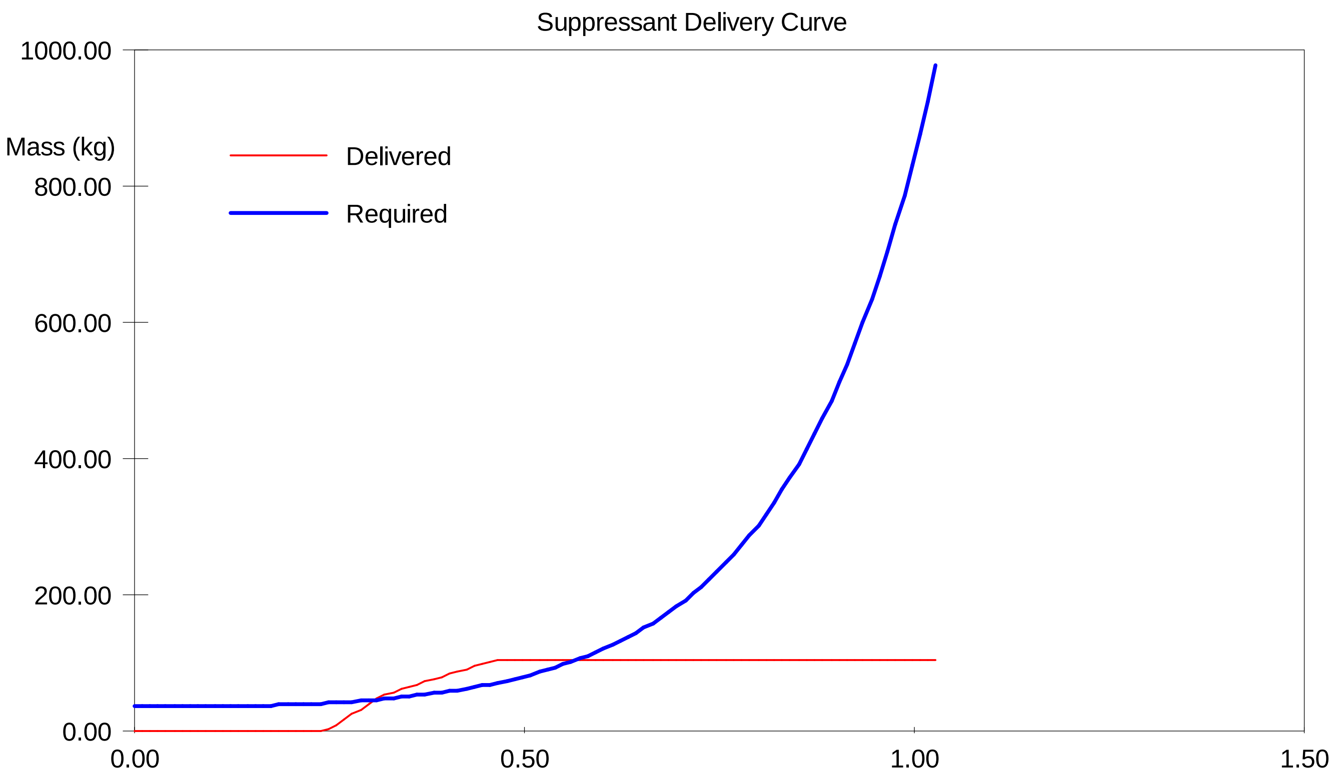
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	6
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	27.0
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.300 bar(a)

+ 0.057 bar contribution from suppressor(s)



Engineer: AT

Time 19:45

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	110.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	52.0	mbar(g)	

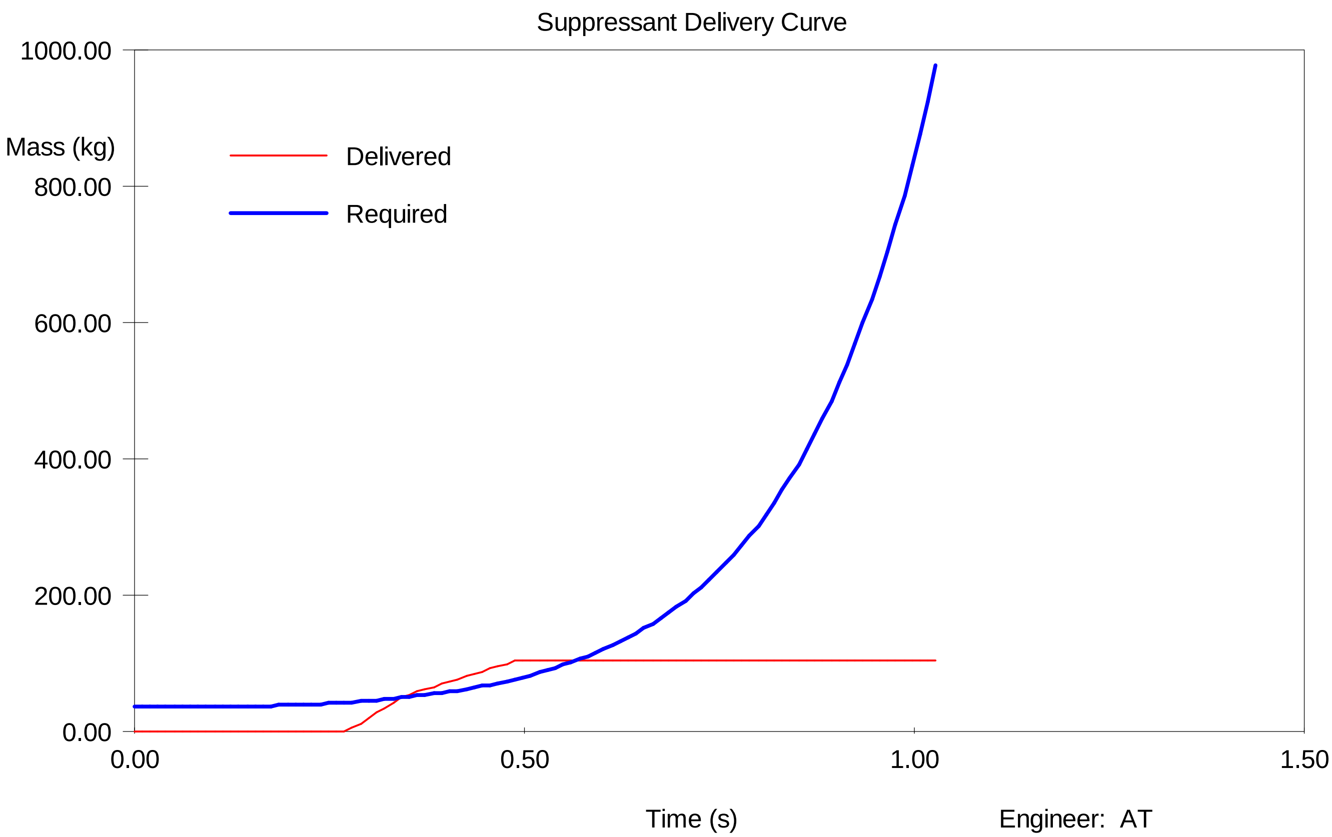
Suppressant = Suppressant-X

Suppressor Requirements

Description	PistonFire
Part Number	
Quantity	6
Suppressor Pressure, bar(g)	62.0
Suppressor Volume, litres	27.0
Head Diameter, mm	78.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.390 bar(a)

\+ 0.057 bar contribution from suppressor(s)



Engineer: AT

Time 19:45

Date: October 21 2009

Industrial Explosion Protection Model v 4.2.0

Customer: Ref:

Description:

Explosion Hazard

Vessel Volume =	110.00	cubic metres	Compact Vessel Assumption
Initial Pressure =	1.000	bar(a)	
Maximum Pressure =	9.000	bar(a)	
K value =	128	bar m/s	
Auto Ignition Temperature =	400	degrees Celsius	
Detection Pressure =	35.0	mbar(g)	

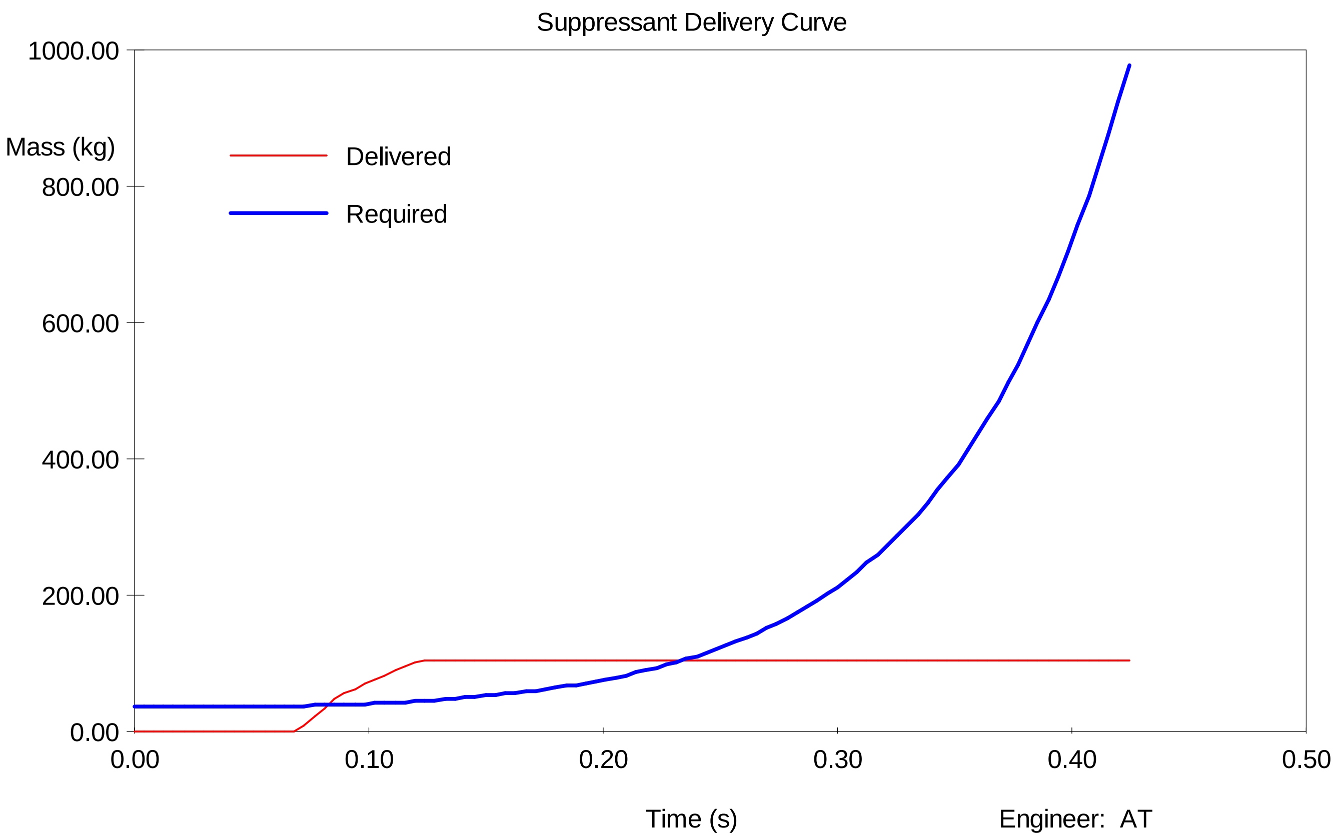
Suppressant = Suppressant-X

Suppressor Requirements

Description	EHRD
Part Number	
Quantity	5
Suppressor Pressure, bar(g)	35.0
Suppressor Volume, litres	42.5
Head Diameter, mm	125.0
Elbow	Yes
Spreader	Flat

Reduced Pressure = 1.098 bar(a)

+ 0.048 bar contribution from suppressor(s)



Engineer: AT

Time 19:51

Date: October 21 2009

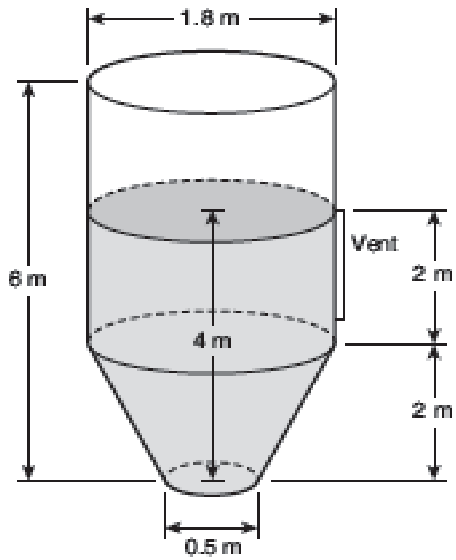
Cylindrical Vessels with Conical Hopper Extension

Calculate L/D for Bottom-up Flame Propogation

Volume above Top of Vent (not included in Effective Volume for L/D)

Length	0	meters	The distance from the top of the vessel to the top of the vent.
Diam 1	4	meters	Diameter of larger cylindrical cross-section
Volume	0	cubic meters	Volume of Cylindrical Section
Length	8.4	meters	The distance from the top of the Conical Hopper to the opposite end of the vent.
Diam 1	4	meters	Diameter of larger cylindrical cross-section
Volume	105.5575	cubic meters	Volume of Cylindrical Section
Height h	1	meters	The distance from the top to the bottom of the the Conical Hopper
Diam 2	0.5	meters	Diameter at bottom of Conical Hopper
Volume	4.777839	cubic meters	Volume of Cylindrical Hopper Section
	110.3354		

$$V = \pi \cdot (h) \frac{\left[(D_1)^2 + (D_1 \cdot D_2) + (D_2)^2\right]}{12}$$



The effective area, A_{eff}, shall be determined by dividing V_{eff} by H (based on the longest central axis flame length). With only one vent, enter the longest distance from one end of the vessel to the opposite end of the vent.

H	9.4	meters
Veff	110.3354	cubic meters
Aeff	11.7378	sq meters

The effective hydraulic diameter, D_{he}, for the enclosure shall be determined based upon the general shape of the enclosure taken normal to the central axis.

$D_{he} = 4 * A_{eff} / p$, Where p is the perimeter of the general shape above the hopper

Dhe	3.865881	meters
L/D	2.431529	**Top-Down**

L/D can not be less than 1, by definition

NFPA 68-2007 Dust in Equipment
Silo Option 1

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	9.4	meters	
Volume (V)	110	cubic meters	(This is total volume, not Veff)
Area (Aeff)	11.74	square meters	
Diameter (Dhe)	3.866	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.375023	bar			
Pmax	8	bar			
Pstat	0.11	bar	US	0.5	psig
Π	0.046878	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{\;4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{max}}{P_{red}} - 1}$$

Av0= 2.119521 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.431454 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp(}\; -0.95 \cdot \mathbf{P}_{\mathbf{red}}^{\;2}\,\mathbf{)}\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 2.711851 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building? N YES/NO Correction factor of 1.7 if a building (occupiable)

Av2/Av1= 0

$$A_{v2}=\left[1+\frac{Max\cdot(v_{axial},v_{tan})-20}{36}\cdot 0.7\right]\cdot A_{v1}$$

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	1.709091	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial}=\frac{Q_{air}\cdot L}{V}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max}=\frac{2\cdot(3.14)\cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 2.711851 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 2 number of panels

$$M_T=\left[6.67\cdot\left(P_{red}^{0.2}\right)\cdot\left(n^{0.3}\right)\cdot\left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	1082.547	kg/m^2	
MT	40	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2

US 0.75 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3}=\left[1+\frac{(0.0075)\cdot M^{0.6}\cdot K_{St}^{0.5}}{n^{0.3}\cdot V\cdot P_{red}^{0.2}}\right]\cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 2.711851 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation: 0.90395

2.712

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
H = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	9.4	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	110	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm

Xr	1.140426	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4=	2.711851	sq meters	4203	sq inches

NFPA 68-2007 Dust in Equipment
Silo Option 2

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H)	9.4	meters	
Volume (V)	110	cubic meters	(This is total volume, not Veff)
Area (Aeff)	11.74	square meters	
Diameter (Dhe)	3.866	meters	

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt	128	bar-m/sec			
Pred	0.35	bar			
Pmax	8	bar			
Pstat	0.11	bar	US	0.5	psig
Π	0.04375	Pred/Pmax	Metric =	0.034474	barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{max}}{P_{red}} - 1}$$

Av0= 2.197576 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.431454 L/D≤6 (8 for silos)

$$\mathbf{A_{v1}} = \mathbf{A_{v0}}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp(}\ -0.95 \cdot \mathbf{P_{red}}^{2}\mathbf{)}\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 2.822395 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building?	N	YES/NO	Correction factor of 1.7 if a building (occupiable)

Av2/Av1= 0

Flow-Created?	N	YES/NO
Inlet Air	20	m^3/sec
Inlet Pipe Diam	1	m
Outlet Pipe Diam	1	m
Vaxial	1.709091	meter/sec
Vtangential	12.7324	meter/sec (0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air} \cdot L}{V}$$

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO
Rotational Radius	0.5	meter
Rotational Speed	1000	RPM
Vtangential	26.17994	meter/sec (0.5 Vtan_max)

$$v_{tan\_max} = \frac{2 \cdot (3.14) \cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 2.822395 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 2 number of panels

$$M_T = \left[6.67 \cdot \left(P_{red}^{0.2}\right) \cdot \left(n^{0.3}\right) \cdot \left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	1057.865	kg/m^2	
MT	40	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2 US 0.75 lb/sq ft If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 If greater than 40 kg/m^2, consult an expert

$$A_{v3} = \left[1 + \frac{(0.0075) \cdot M^{0.6} \cdot K_{St}^{0.5}}{n^{0.3} \cdot V \cdot P_{red}^{0.2}}\right] \cdot A_{v2}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 2.822395 sq meters If M < MT, then there is no area correction for inertia

Partial Volume Correction
Calculate the worst-case building partial volume fraction, X_r, from the following equation: 2.838

$$X_r = \frac{\overline{M}_f}{A_{fs} c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss} V c_w} + \frac{M_e}{V c_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
H = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	9.4	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	110	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm

Xr	1.140426	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4= 2.822395 sq meters 4375 sq inches

NFPA 68-2007 Dust in Equipment
Silo Option 3

Enclosure Section Dimensions
(see L_D Tab to calculate these terms)

Length (H) 9.4 meters
Volume (V) 110 cubic meters (This is total volume, not Veff)
Area (Aeff) 11.74 square meters
Diameter (Dhe) 3.866 meters

KSt is the deflagration index
Pred is the maximum pressure developed during the vented explosion
Pmax is the maximum pressure developed in a closed explosion test
Pstat is the static release pressure of the vent panel
Π is the ratio of Pred/Pmax

KSt 128 bar-m/sec
Pred 0.28 bar
Pmax 8 bar
Pstat 0.11 bar US 0.5 psig
Π 0.035 Pred/Pmax Metric = 0.034474 barg

$$A_{v0} = 1 \cdot 10^{-4} \cdot \left[1 + 1.54 \cdot P_{stat}^{4/3}\right] \cdot K_{St} \cdot V^{3/4} \cdot \sqrt{\frac{P_{\max}}{P_{red}} - 1}$$

Av0= 2.46818 sq meters

Check for L/D less than 2
(Use inputs above)

L/D (H/Dhe) 2.431454 L/D≤6 (8 for silos)

$$\mathbf{A}_{\mathbf{v}1} = \mathbf{A}_{\mathbf{v}0}\left[1 + 0.6 \cdot \left(\frac{\mathbf{L}}{\mathbf{D}} - 2\right)^{0.75} \cdot \mathbf{exp}\left(-0.95 \cdot \mathbf{P}_{\mathbf{red}}^{2}\right)\right]$$

If L/D >2, increase vent area, else Av1=Av0
Av1= 3.199963 sq meters

Turbulence Correction
Select as many options as applicable for the enclosure and this picks the highest correction.

Building? N YES/NO Correction factor of 1.7 if a building (occupiable)
Av2/Av1= 0

$$A_{v2} = \left[1 + \frac{Max \cdot (v_{axial}, v_{tan}) - 20}{36} \cdot 0.7\right] \cdot A_{v1}$$

$$A_{v2} = \left[1 + \frac{Max\cdot(v_{axial}, v_{tan}) - 20}{36}\cdot 0.7\right]\cdot A_{v1}$$

Flow-Created?	N	YES/NO	
Inlet Air	20	m^3/sec	
Inlet Pipe Diam	1	m	
Outlet Pipe Diam	1	m	
Vaxial	1.709091	meter/sec	
Vtangential	12.7324	meter/sec	(0.5 Vtan_max)

$$v_{axial} = \frac{Q_{air}\cdot L}{V}$$

Correction for Flow-Created Turbulence (uses the maximum Axial or Tangential Turbulence)
This would be typical for a cyclone
Av2/Av1= 0

Rotating Equip?	N	YES/NO	
Rotational Radius	0.5	meter	
Rotational Speed	1000	RPM	
Vtangential	26.17994	meter/sec	(0.5 Vtan_max)

$$v_{tan\_max} = \frac{2\cdot(3.14)\cdot Nr}{60}$$

Correction factor if Rotating Equipment
This would be typical for a grinder or hammermill
Av2/Av1= 0

Pick highest value of selected "YES" options above
Highest Av2/Av1= 0 No adjustment made if calculated Av2/Av1 is <1

If Velocities are less than 20 meters/sec, then Av2=Av1.

Av2= 3.199963 sq meters

For Panel Mass > 40 kg/m^2, NFPA-68 recommends use of the Annex F (not included here)
Based on the Task Group Activities, the inertia equations are applicable up to KSt limit of the basic equation (i.e. KSt=800 bar-m/sec)

Inertia Correction for Panel Mass ≤ 40 kg/m^2

n 2 number of panels

$$M_T = \left[6.67\cdot\left(P_{red}^{0.2}\right)\cdot\left(n^{0.3}\right)\cdot\left(\frac{V}{K_{St}^{0.5}}\right)\right]^{1.67}$$

Mformula	981.889	kg/m^2	
MT	40	kg/m^2	MT is minimum of 40 kg/m^2 or the formula above.

Vent area is increased if panel density exceeds the threshold or 40 kg/m^2, whichever is smaller. The total mass
Intended Vent Panel Density

M 19 kg/m^2 | US 0.75 lb/sq ft | If panel density is in US units, enter here and enter metric units at left
Metric = 3.7 kg/m^2 | If greater than 40 kg/m^2, consult an expert

$$\boxed{A_{v3} = \left[1 + \frac{(0.0075)\cdot M^{0.6}\cdot K_{St}^{0.5}}{n^{0.3}\cdot V\cdot P_{red}^{0.2}}\right]\cdot A_{v2}}$$

OSECO PANEL MASSES		
CRP	13.4	kg/m^2
CRV	13.4	kg/m^2
CRVC	19	kg/m^2
RNDCC	15.5	kg/m^2
MVC	19	kg/m^2
GLV	7.2	kg/m^2

Av3 3.199963 sq meters — If M < MT, then there is no area correction for inertia

Partial Volume Correction

Calculate the worst-case building partial volume fraction, X_r, from the following equation: 2.838

$$X_r = \frac{\overline{M}_f}{A_{fs}c_w H} + \frac{\overline{M}_s A_{sur}}{A_{ss}Vc_w} + \frac{M_e}{Vc_w}$$

where:

X_r = worst-case building partial fraction
$\overline{M}_f$ = average mass (gram) of floor samples
A_{fs} = measured floor areas
c_w = worst-case dust concentration
$\underline{H}$ = ceiling height of the building
$\overline{M}_s$ = average mass (gram) of surface samples
A_{sur} = total area of surfaces with dust deposits
A_{ss} = measured sample areas of surfaces with dust deposits
V = building volume
M_e = total mass of combustible dust that could be released from the process equipment in the building

Mf	148	gm	Estimate Fill Fraction		YES	YES or NO	
Afs	0.37	sq meters	If YES	Mf/Afs =	640	gm/m2	**Assumed Dust on Floor of Operational Room**
Cw	500	gm/m^3		Ms/Ass =	640	gm/m2	
H	9.4	meters		Cw =	200	gm/m2	
Ms	100	gm					
Asur	20	sq meters	Calculated from Inputs at Left				
Ass	0.37	sq meters	If NO	Mf/Afs =	400	gm/m2	
V	110	m^3		Ms/Ass =	270.2703	gm/m2	
				Cw =	500	gm/m2	

Always Enter the mass of combustibles that could be released from equipment or storage below:

Me	4.8	kg
	4800	gm
Xr	1.140426	fill fraction

If Xr is less then Π, then no venting is required
If Xr is greater than 1, partial volume does not apply and Av4=Av3

$$\boxed{A_{v4} = A_{v3} \cdot X_r^{-1/3} \cdot \sqrt{\frac{X_r - \Pi}{1 - \Pi}}}$$

Av4= 3.199963 sq meters 4960 sq inches

Venting Panels

OSECO®

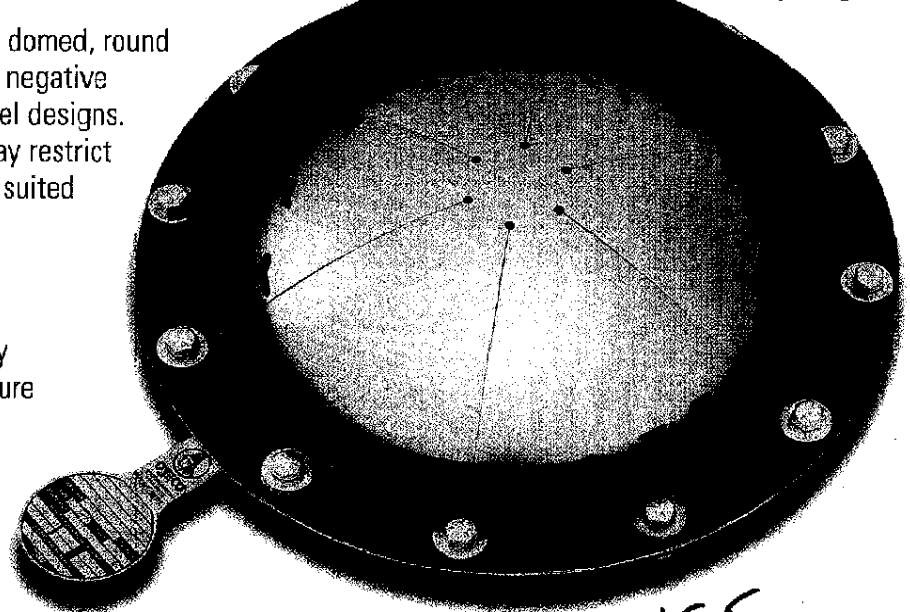
RNDCC Rupture Panels

perimeter cut –

OSECO's Model RNDC Rupture Panel is a domed, round design that withstands higher positive or negative system operating pressures than flat panel designs. Panel frame vacuum support bars that may restrict flow are not required. The RNDC is well suited for high cycling service.

Utilizing our in-house laser and specially designed hydraulic presses, OSECO has dedicated an entire manufacturing facility solely for the development and manufacture of rupture panels.

Should your application require an advanced crowned (RNDC) rupture panel, let OSECO meet your needs.



Recommended maximum ratio of system operating pressure to rated rupture pressure is:

- 70% for Crowned Panels

Rupture Tolerances:

- ± 1/2 PSI for Panels Rated Above 2 psig.
- ± 1/4 PSI for Panels Rated 2 psig or Less.

TÜV CERT ISO 9001

15.5

RNDCC - 31.6kg/m²

RNDC RUPTURE PANEL SPECIFICATIONS

SIZE INCHES (mm)	STANDARD ANGLE FRAME I.D.	RELIEF AREA in² (m²)	RUPTURE PRESSURES AVAILABLE: TEFLON SEAL PSIG @ 72° F MIN (bar)	TEFLON SEAL MAX (bar)	ALUMINUM SEAL PSIG @ 72° F MIN (bar)	ALUMINUM SEAL MAX (bar)
8" (200mm)	8.125" (206mm)	45 (0.029)	4.5 (0.310)	10.0 (0.689)	N/A	N/A
10" (250mm)	10.125" (257mm)	72 (0.046)	3.6 (0.248)	10.0 (0.689)	N/A	N/A
12" (300mm)	12.1875" (310mm)	101 (0.065)	3.0 (0.207)	10.0 (0.689)	N/A	N/A
14" (350mm)	14.1875" (360mm)	135 (0.087)	2.6 (0.179)	10.0 (0.689)	N/A	N/A
16" (400mm)	16.25" (413mm)	176 (0.113)	2.3 (0.159)	10.0 (0.689)	N/A	N/A
18" (450mm)	18.25" (464mm)	227 (0.146)	2.0 (0.138)	10.0 (0.689)	N/A	N/A
20" (500mm)	20.25" (514mm)	279 (0.180)	1.8 (0.124)	10.0 (0.689)	N/A	N/A
24" (600mm)	24.25" (616mm)	415 (0.267)	1.5 (0.103)	10.0 (0.689)	7.7 (0.531)	10.0 (0.689)
28" (700mm)	28.25" (718mm)	530 (0.342)	1.5 (0.103)	10.0 (0.689)	6.7 (0.462)	10.0 (0.689)
30" (750mm)	30.25" (768mm)	650 (0.419)	1.5 (0.103)	10.0 (0.689)	6.2 (0.427)	10.0 (0.689)
32" (800mm)	32.25" (819mm)	754 (0.486)	1.5 (0.103)	10.0 (0.689)	5.8 (0.400)	10.0 (0.689)
34" (850mm)	34.25" (870mm)	855 (0.551)	1.5 (0.103)	10.0 (0.689)	5.4 (0.372)	10.0 (0.689)
36" (900mm)	36.25" (921mm)	955 (0.616)	1.5 (0.103)	9.7 (0.669)	5.1 (0.352)	9.7 (0.669)

For Sizes Or Pressures Not Listed, Please Consult Factory

19kg/m²

CRVC RUPTURE PANEL SPECIFICATIONS

SIZE INCHES (mm)	RELIEF AREA in² (m²)	RUPTURE PRESSURES AVAILABLE TEFLON SEAL PSIG @ 72° F MIN (bar)	TEFLON SEAL MAX (bar)	ALUMINUM SEAL PSIG @ 72° F MIN (bar)	ALUMINUM SEAL MAX (bar)
8" x 8" (203mm x 203mm)	42 (0.035)	N/A	N/A	N/A	N/A
9" x 12" (229mm x 305mm)	78 (0.050)	2.4 (0.165)	10.0 (0.689)	N/A	N/A
12" x 12" (305mm x 305mm)	110 (0.071)	2.4 (0.165)	10.0 (0.689)	N/A	N/A
12" x 18" (305mm x 457mm)	173 (0.111)	2.3 (0.159)	10.0 (0.689)	N/A	N/A
12" x 24" (305mm x 610mm)	236 (0.152)	2.3 (0.159)	10.0 (0.689)	N/A	N/A
16" x 18" (406mm x 457mm)	239 (0.154)	2.3 (0.159)	10.0 (0.689)	7.8 (0.538)	10.0 (0.689)
17-1/8" x 17 1/8" (435mm x 435mm)	244 (0.157)	2.3 (0.159)	10.0 (0.689)	7.7 (0.531)	10.0 (0.689)
13.9" x 22.8" (354mm x 580mm)	265 (0.171)	2.25 (0.155)	10.0 (0.689)	7.5 (0.517)	10.0 (0.689)
19.7" x 19.7" (500mm x 500mm)	330 (0.213)	2.25 (0.155)	10.0 (0.689)	7.5 (0.517)	10.0 (0.689)
18" x 24" (457mm x 610mm)	371 (0.239)	1.9 (0.131)	10.0 (0.689)	7.5 (0.517)	10.0 (0.689)
20" x 24" (508mm x 610mm)	416 (0.268)	1.9 (0.131)	10.0 (0.689)	7.5 (0.517)	10.0 (0.689)
18" x 28" (457mm x 711mm)	437 (0.282)	1.9 (0.131)	10.0 (0.689)	7.5 (0.517)	10.0 (0.689)
18" x 30" (457mm x 762mm)	470 (0.303)	1.9 (0.131)	10.0 (0.689)	7.0 (0.483)	10.0 (0.689)
24" x 24" (610mm x 610mm)	506 (0.326)	1.9 (0.131)	10.0 (0.689)	6.8 (0.469)	10.0 (0.689)
18" x 35" (457mm x 889mm)	552 (0.356)	1.9 (0.131)	10.0 (0.689)	6.5 (0.448)	10.0 (0.689)
18" x 36" (457mm x 914mm)	569 (0.367)	1.9 (0.131)	10.0 (0.689)	6.5 (0.448)	10.0 (0.689)
20.5" x 32.3" (520mm x 820mm)	584 (0.376)	1.9 (0.131)	10.0 (0.689)	6.5 (0.448)	10.0 (0.689)
18" x 37-3/4" (457mm x 959mm)	598 (0.385)	1.9 (0.131)	10.0 (0.689)	6.5 (0.448)	10.0 (0.689)
24" x 30" (610mm x 762mm)	641 (0.413)	1.9 (0.131)	10.0 (0.689)	6.2 (0.427)	10.0 (0.689)
27" x 27.5" (686mm x 698mm)	663 (0.427)	1.9 (0.131)	10.0 (0.689)	6.2 (0.427)	10.0 (0.689)
27.6" x 27.6" (700mm x 700mm)	679 (0.438)	1.8 (0.124)	10.0 (0.689)	6.2 (0.427)	10.0 (0.689)
23.1" x 36.2" (586mm x 920mm)	748 (0.483)	1.8 (0.124)	10.0 (0.689)	5.6 (0.386)	10.0 (0.689)

SIZE INCHES (mm)	RELIEF AREA in² (m²)	RUPTURE PRESSURES AVAILABLE TEFLON SEAL PSIG @ 72° F MIN (bar)	TEFLON SEAL MAX (bar)	ALUMINUM SEAL PSIG @ 72° F MIN (bar)	ALUMINUM SEAL MAX (bar)
24" x 36" (610mm x 914mm)	776 (0.500)	1.8 (0.124)	10.0 (0.689)	5.6 (0.386)	10.0 (0.689)
30" x 30" (762mm x 762mm)	812 (0.524)	1.8 (0.124)	10.0 (0.689)	5.5 (0.379)	10.0 (0.689)
20.5" x 44" (521mm x 1118mm)	807 (0.521)	1.8 (0.024)	10.0 (0.689)	5.5 (0.379)	10.0 (0.689)
20" x 48" (508mm x 1219mm)	860 (0.555)	1.8 (0.124)	10.0 (0.689)	5.4 (0.372)	10.0 (0.689)
27" x 36" (686mm x 914mm)	879 (0.567)	1.8 (0.124)	10.0 (0.689)	5.3 (0.365)	10.0 (0.689)
25" x 40" (635mm x 1016mm)	904 (0.583)	1.8 (0.124)	10.0 (0.689)	5.2 (0.359)	10.0 (0.689)
27" x 39" (686mm x 991mm)	956 (0.616)	1.8 (0.124)	9.6 (0.662)	5.1 (0.352)	9.6 (0.662)
24" x 44" (610mm x 1118mm)	956 (0.616)	1.8 (0.124)	9.6 (0.662)	5.1 (0.352)	9.6 (0.662)
25" x 43" (635mm x 1092mm)	975 (0.629)	1.8 (0.124)	9.5 (0.655)	5.0 (0.345)	9.5 (0.655)
27" x 40" (686mm x 1016mm)	981 (0.633)	1.8 (0.124)	9.5 (0.655)	5.0 (0.345)	9.5 (0.655)
30" x 36" (762mm x 914mm)	983 (0.634)	1.8 (0.124)	9.5 (0.655)	5.0 (0.345)	9.5 (0.655)
25.5" x 43.1" (648mm x 1096mm)	1000 (0.645)	1.7 (0.117)	9.5 (0.655)	5.0 (0.345)	9.5 (0.655)
24" x 47" (610mm x 1194mm)	1023 (0.660)	1.7 (0.117)	9.2 (0.634)	4.9 (0.338)	9.2 (0.634)
24" x 48" (610mm x 1219mm)	1046 (0.675)	1.7 (0.117)	9.1 (0.627)	4.8 (0.331)	9.1 (0.627)
27" x 43" (686mm x 1092mm)	1058 (0.682)	1.7 (0.117)	9.1 (0.627)	4.8 (0.331)	9.1 (0.627)
25" x 47" (635mm x 1194mm)	1069 (0.689)	1.7 (0.117)	9.1 (0.627)	4.8 (0.331)	9.1 (0.627)
27.5" x 44.3" (698mm x 1126mm)	1112 (0.717)	1.7 (0.117)	9.1 (0.627)	4.8 (0.331)	9.1 (0.627)
25" x 49" (635mm x 1245mm)	1116 (0.720)	1.7 (0.117)	9.1 (0.627)	4.8 (0.331)	9.1 (0.627)
31.5" x 39.4" (800mm x 1000mm)	1135 (0.732)	1.7 (0.117)	9.1 (0.627)	4.8 (0.331)	9.1 (0.627)
36" x 36" (914mm x 914mm)	1190 (0.767)	1.7 (0.117)	8.6 (0.593)	4.7 (0.324)	8.6 (0.593)
30" x 44" (762mm x 1118mm)	1211 (0.781)	1.7 (0.117)	8.6 (0.593)	4.7 (0.324)	8.6 (0.593)
27" x 52" (686mm x 1321mm)	1287 (0.830)	1.7 (0.117)	8.4 (0.579)	4.6 (0.317)	8.4 (0.579)

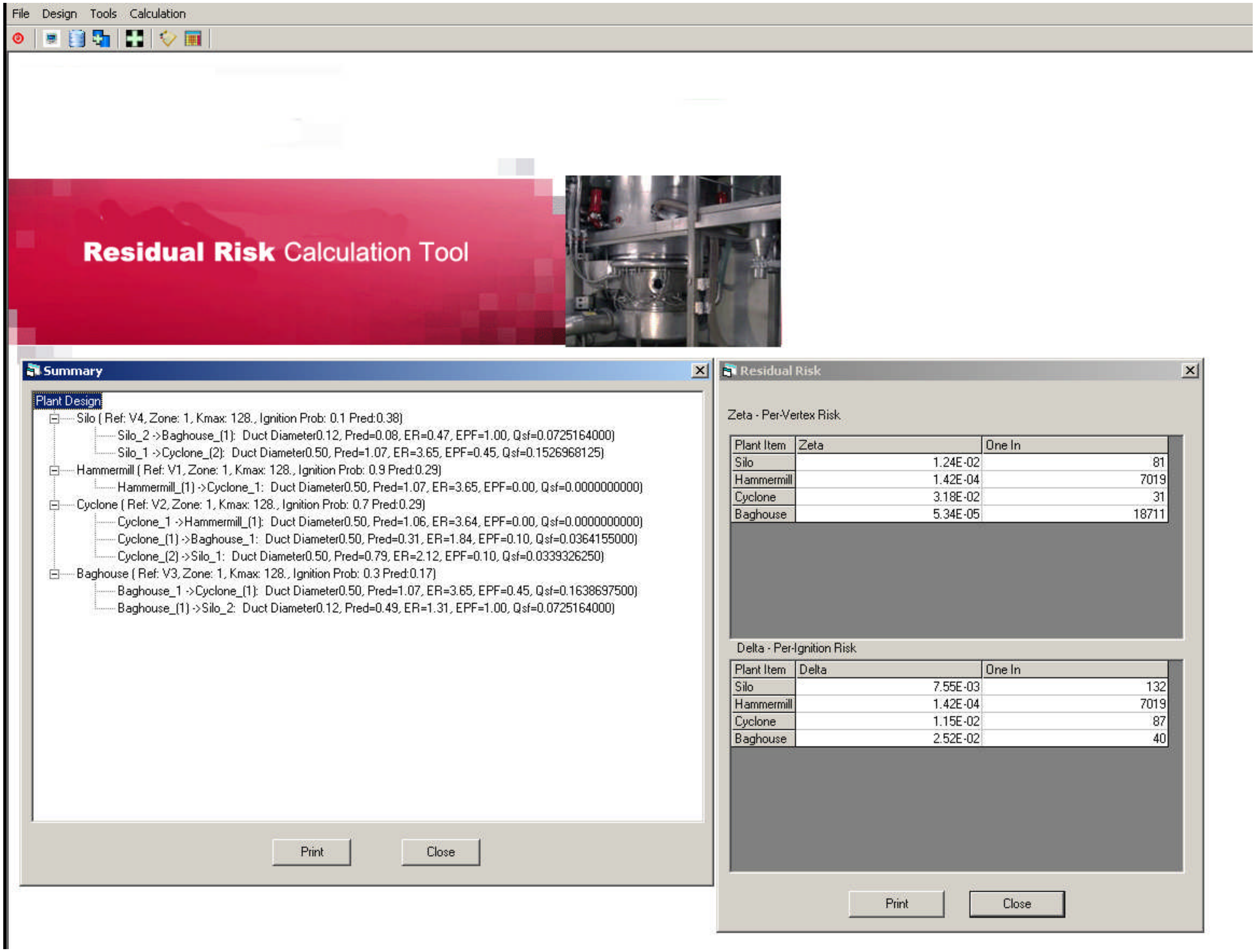
SIZE INCHES (mm)	RELIEF AREA in² (m²)	RUPTURE PRESSURES AVAILABLE TEFLON SEAL PSIG @ 72° F MIN (bar)	TEFLON SEAL MAX (bar)	ALUMINUM SEAL PSIG @ 72° F MIN (bar)	ALUMINUM SEAL MAX (bar)
27.5" x 51.6" (698mm x 1310mm)	1300 (0.839)	1.7 (0.117)	8.4 (0.579)	4.6 (0.317)	8.4 (0.579)
30" x 48" (762mm x 1219mm)	1325 (0.855)	1.7 (0.117)	8.4 (0.579)	4.6 (0.317)	8.4 (0.579)
38" x 38" (965mm x 965mm)	1332 (0.859)	1.7 (0.117)	8.4 (0.579)	4.6 (0.317)	8.4 (0.579)
25.5" x 57.1" (648mm x 1450mm)	1334 (0.904)	1.7 (0.117)	8.4 (0.579)	4.6 (0.317)	8.4 (0.579)
31.5" x 47.2" (800mm x 1200mm)	1372 (0.885)	1.7 (0.117)	8.3 (0.572)	4.6 (0.317)	8.3 (0.572)
39.4" x 39.4" (1000mm x 1000mm)	1434 (0.925)	1.6 (0.110)	8.3 (0.572)	4.5 (0.310)	8.3 (0.572)
27" x 58" (686mm x 1473mm)	1440 (0.929)	1.6 (0.110)	8.2 (0.565)	4.4 (0.303)	8.2 (0.572)
27.5" x 57.1" (698mm x 1450mm)	1444 (0.931)	1.6 (0.110)	7.9 (0.545)	4.4 (0.303)	7.9 (0.545)
36" x 44" (914mm x 1118mm)	1466 (0.946)	1.6 (0.110)	7.9 (0.545)	4.4 (0.303)	7.9 (0.545)
25" x 66" (635mm x 1676mm)	1515 (0.977)	1.6 (0.110)	7.8 (0.538)	4.3 (0.296)	7.8 (0.538)
27" x 66" (686mm x 1676mm)	1644 (1.061)	1.6 (0.110)	7.4 (0.510)	4.2 (0.290)	7.4 (0.510)
36.2" x 49.4" (920mm x 1254mm)	1662 (1.072)	1.6 (0.110)	7.4 (0.510)	4.2 (0.290)	7.4 (0.510)
30" x 60" (762mm x 1524mm)	1667 (1.075)	1.6 (0.110)	7.4 (0.510)	4.2 (0.290)	7.4 (0.510)
42" x 44" (1067mm x 1118mm)	1721 (1.110)	1.6 (0.110)	7.4 (0.510)	4.1 (0.283)	7.4 (0.510)
44" x 44" (1118mm x 1118mm)	1806 (1.165)	1.5 (0.103)	7.2 (0.496)	4.1 (0.283)	7.2 (0.496)
44.5" x 44.5" (1130mm x 1130mm)	1849 (1.196)	1.5 (0.103)	7.1 (0.490)	4.0 (0.276)	7.1 (0.490)
36" x 60" (914mm x 1524mm)	2018 (1.302)	1.5 (0.103)	7.0 (0.483)	4.0 (0.276)	7.0 (0.483)
39" x 66.5" (990mm x 1690mm)	2437 (1.572)	1.4 (0.097)	6.4 (0.441)	3.8 (0.262)	6.4 (0.441)
44" x 60" (1118mm x 1524mm)	2486 (1.604)	1.4 (0.097)	6.4 (0.441)	3.8 (0.262)	6.4 (0.441)
42" x 65" (1067mm x 1651mm)	2571 (1.659)	1.4 (0.097)	6.3 (0.434)	3.8 (0.262)	6.3 (0.434)
44" x 69" (1118mm x 1753mm)	2868 (1.850)	1.3 (0.080)	6.1 (0.421)	3.7 (0.255)	6.1 (0.421)

For Sizes Or Pressures Not Listed, Please Consult Factory

[illegible] 2/17/04

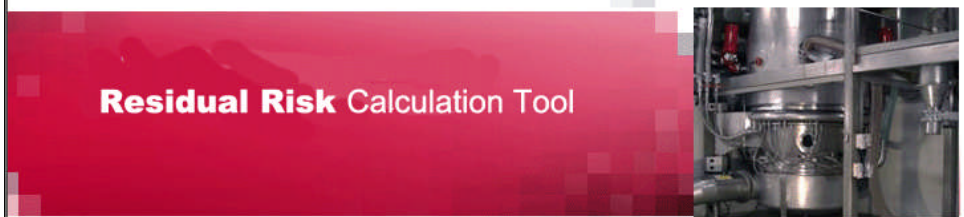
Residual Risk Calculations

BASELINE CALCULATION RESIDUAL RISK SUMMARY



OPTIMIZED DESIGN – OPTION-1 RESIDUAL RISK CALCULATION SUMMARY

File Design Tools Calculation



Summary

Plant Design

- Silo (Ref: V4, Zone: 1, Kmax: 128., Ignition Prob: 0.1 Pred:0.38)
  - Silo_2 ->Baghouse_(1): Duct Diameter0.12, Pred=0.08, ER=0.47, EPF=1.00, Qsf=0.0725164000)
  - Silo_1 ->Cyclone_(2): Duct Diameter0.50, Pred=0.60, ER=3.17, EPF=1.00, Qsf=0.3393262500)
- Hammermill (Ref: V1, Zone: 1, Kmax: 128., Ignition Prob: 0.9 Pred:0.29)
  - Hammermill_(1) ->Cyclone_1: Duct Diameter0.50, Pred=0.60, ER=3.17, EPF=0.10, Qsf=0.0364155000)
- Cyclone (Ref: V2, Zone: 1, Kmax: 128., Ignition Prob: 0.7 Pred:0.19)
  - Cyclone_1 ->Hammermill_(1): Duct Diameter0.50, Pred=1.06, ER=3.64, EPF=0.45, Qsf=0.1633500000)
  - Cyclone_(1) ->Baghouse_1: Duct Diameter0.50, Pred=0.31, ER=1.84, EPF=1.00, Qsf=0.3641550000)
  - Cyclone_(2) ->Silo_1: Duct Diameter0.50, Pred=0.79, ER=2.12, EPF=1.00, Qsf=0.3393262500)
- Baghouse (Ref: V3, Zone: 1, Kmax: 128., Ignition Prob: 0.3 Pred:0.17)
  - Baghouse_1 ->Cyclone_(1): Duct Diameter0.50, Pred=0.60, ER=3.17, EPF=1.00, Qsf=0.3641550000)
  - Baghouse_(1) ->Silo_2: Duct Diameter0.12, Pred=0.49, ER=1.31, EPF=1.00, Qsf=0.0725164000)

Print Close

Residual Risk

Zeta - Per-Vertex Risk

Plant Item	Zeta	One In
Silo	1.24E-03	807
Hammermill	1.61E-04	6227
Cyclone	4.77E-03	210
Baghouse	5.59E-05	17879

Delta - Per-Ignition Risk

Plant Item	Delta	One In
Silo	1.17E-03	855
Hammermill	1.43E-04	7017
Cyclone	2.98E-04	3355
Baghouse	4.61E-03	217

Print Close

OPTIMIZED DESIGN – OPTION-2 RESIDUAL RISK CALCULATION SUMMARY