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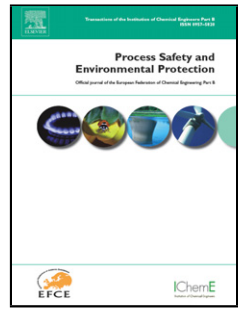

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Authors: Muhammad Athar, Azmi Mohd Shariff, Azizul Buang, Salman Nazir, Heri Hermansyah, Tan Lian See

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Process equipment common attributes for inherently safer process design at preliminary design stage

Muhammad Athar [a], Azmi Mohd Shariff [a, *], Azizul Buang [a], Salman Nazir [b], Heri Hermansyah [c], Tan Lian See [d]

[a] Centre of Advanced Process Safety (CAPS), Chemical Engineering Department, Universiti Teknologi PETRONAS, 32610 Bandar Seri Iskandar, Perak Darul Ridzuan, Malaysia.

[b] Training and Assessment Research Group (TARG), Department of Maritime Operations, University of South-Eastern Norway, Postboks 4, 3199, Borre, Norway.

[c] Industrial Bioprocess Engineering Laboratory, Department of Chemical Engineering, Faculty of Engineering, Universitas Indonesia, Kampus UI Depok, Depok 16424, Indonesia.

[d] Department of Chemical Process Engineering, Malaysia - Japan International Institute of Technology (MJIIT), Universiti Teknologi Malaysia (UTM), Jalan Sultan Yahya Petra, 54100 Kuala Lumpur, Malaysia.

[*] Corresponding author:

Tel: +60 5 3687570

Fax: +60 5 3656176

E-mail Address: azmish@utp.edu.my

Graphical abstract

Highlights

- Inherently safer process design method using process equipment common attributes is developed
- New intra-equipment index identifies critical equipment using equipment features
- Explosion as the equipment failure outcome is focused on using accidents data
- Risk is analyzed against the newly developed explosion risk matrix
- Various ISD options are implied to shift the risk towards acceptability range

Abstract

Hazards associated with chemical processes can lead to accidents and require therefore proper management. An inherent safety strategy is a proactive approach to serve this purpose, one capable of minimizing hazards whilst offering sustainable process design. Current inherent safety techniques are limited to comparing process routes and selecting safer one using the process parameters, whereas process equipment characteristics are rarely scrutinized. Therefore, this paper consolidates a new technique that integrates the mutually shared attributes of process equipment, in order to offer inherently safer process design at the preliminary design stage. Inherent safety assessment for process equipment (ISAPE) consists of an indexing procedure, followed by risk assessment. The indexing procedure can highlight the critical process equipment, which can be further studied through risk assessment. When the risk is beyond acceptable threshold and must be minimized, inherent safety concepts are implemented, leading towards inherently safer process design. The complete ISAPE technique is exhibited through the case study of the acetone production process. In this case study, various ISD options have been applied to the critical process equipment, identified through the proposed indexing; the options have then been compared to select the best one. The method is easy to use, and as such, it is suitable to be put into practice by design engineers at the preliminary design stage.

Keywords: Indexing; Inherent safety; Inventory; Plant layout; Preliminary design engineering; Sustainable process design

Nomenclature

Greek Symbols

f	function of
η	efficiency factor
ρ	density (kg/m^3)
γ	ratio of specific heat capacities, C_P/C_V

Alphabetical Symbols

AIT	auto ignition temperature (˚C)
A	empirical constant
A_h	opening area (m^2)
a	time constant
a_1	constant for overpressure
b_1	constant for overpressure
c_1	constant for overpressure
c	concentration
C_D	discharge coefficient
f	explosion frequency (yr^{-1})
k	strength constant
m	total mass (kg)
MIE	minimum ignition energy (mJ)
p	pressure (kPa)
p_{ovr}	overpressure (kPa)
P	probability
T	operating temperature (˚C)
H	energy of combustion (kJ/kg)
z	distance (m)
$\bar{z}$	scaled distance (m)

Subscript Symbols

amb	ambient
del, ign	delayed ignition

exp	explosion
exp/ g/ ign	probability of delayed ignition leading to explosion
f	flammable
imm, ign	immediate ignition
IL	initial leakage
LFL	lower flammability level of mixture
mix	chemical mixture
V	vapor
TNT	trinitrotoluene

1. INTRODUCTION

The progression of the chemical process industry has escalated the magnitude of unwanted loss-containment scenarios. To forestall such situations, both academia and industry professionals are multiplying their efforts to achieve state-of-the-art hazard-tackling strategies. These strategies, which serve both in creating sustainable process designs for new facilities and in managing the hazards in the existing process plants, can be classified along four dimensions (Hendershot, 2011). These four strategies act as the pillars of modern process safety, as demonstrated in Fig. 1(a). The overall target for all these strategies is to minimize the impact of process hazards up to a tolerable level, as the complete elimination of process hazard is practically unfeasible. Each of these strategies offers a certain contribution to the process life cycle. The inherent safety strategy is typically applied at the preliminary design stage. It aims to deliver a sustainable process design allowing to bring the hazard to a minimum level from the start of the life cycle (Song et al., 2018). The sustainability in the process design can be obtained by taking into consideration the typical metrics related to sustainability, for example, energy, water, wastes and pollutants (Carvalho et al., 2008; de Jesús Guillén-Cuevas et al., 2018; Lu et al., 2011; Sikdar, 2003). The inherent safety strategy is mostly dealing with chemicals, operating conditions and inventory. Although the remaining strategies of process safety contribute to the later stages of process lifecycle by controlling the hazard through add-on measures, these pillars are also essential for a perfect process safety system (PSS) in order to achieve the objective of minimum acceptable risk (MAR), which is imposed by authorities. Furthermore, the inherent safety strategy is primarily introduced to address the safety issues in the complete life cycle of a process plant (CCPS, 2010b; Hendershot, 2010). However, it is considered preferable for earlier stages, due to its capabilities in risk management and capital investments, which ultimately leads to generating sustainable process design. This approach has been established over the years, and strategies of inherent safety have been devised by various sources available in the literature (CCPS, 2010b; Kletz and Amyotte, 2010), which are presented in Fig. 1(b). Among inherent safety strategies, intensification, also known as minimization, has been the most frequently identified strategy in industries with the aim of generating the sustainable process designs. On the other hand, the remaining three strategies, i.e., moderation, simplification, and substitution, receive comparatively less attention in sustainable process designing (Hussin et al., 2015).

Since, as mentioned earlier, the inherent safety strategy is mostly related to earlier stages of the process life cycle. However, the available methods such as Dow FEI index cannot quantify the inherent safety at the earlier design stage. Therefore, there is a need for a dedicated tool to quantify the inherent safety at this stage. Prompted by the unavailability of any such method, researchers have endeavored to develop the required tool. In this context, the very first dedicated method, named proto-type index for inherent safety (PIIS) came into existence (Edwards and Lawrence, 1993; Lawrence, 1996), followed by inherent safety index (ISI) (Heikkilä, 1999; Heikkilä et al., 1996). Both these methods are capable of comparing various process routes to identify the safest option. A consequence-based method, called integrated inherent safety index (I2SI), has merged safety and cost dimensions in a single technique. The method aims to analyze the impact of unwanted chemical releases by incorporating various process and material characteristics as well as the cost analysis for different process routes in order to choose the best possible option (Khan and Amyotte, 2004, 2005). All these methods so far have focused on the safety dimension alone, while the environment and safety features have been coupled for the process routes comparison by concentrating on the material and energy consumption (Andraos, 2016). A few other methods are also available which have considered multiple aspects for sustainable process design, including atmospheric hazard index (Gunasekera and Edwards, 2003), life cycle index (LInX) (Khan et al., 2004), integrative environmental performance index (IEPI) (Frank et al., 2016) and inherent chemical process route index for safety, health and environment prospects (Warnasooriya and Gunasekera, 2017). LInX compares different process routes to identify the best one using environment, health, safety, and cost sub-indices. On the other hand, IEPI has been developed to observe the environmental sustainability performance for oil and gas companies. A few other examples of tools and techniques that address both inherent safety and sustainability metrics can be found in the literature (Ee et al., 2015; Koller et al., 2000). Recently, Song et al., (2018) developed an inherent safety performance index (ISPI) which have engaged the fuzzy set with the indexing to solve the issue of exact boundaries for sub-indices. The method was meant for the evaluation of design alternatives to generate the sustainable process design. This method extends the initially proposed indexing methods - such as PIIS and ISI by the inclusion of weighting factors for reaction and equipment. All of these techniques rely upon the indexing concept, which has certain limitations and drawbacks, such as subjective scaling and weighting factors (Ahmad et al., 2014; Srinivasan and Nhan, 2008). To deal with these shortcomings, other themes have been introduced, such as graphical techniques and optimization for estimation purposes. The graphical techniques include a simple graphical method for the inherently safer route (Gupta and Edwards, 2003), a graphical technique for root-cause analysis in inherent safety assessment (Ahmad et al., 2015) and a graphical method for consideration of inherent safety in process design (Ortiz-Espinoza et al., 2017). The optimization methods are more flexible because of their capability to tackle conflicting objectives, which can be solved through a multi-objective optimization (MOO) scheme. This type of methods include the integration of quantitative risk assessment (QRA) for safety analysis in the optimization approach focusing on economics, safety and environment (Fuentes-Cortés et al., 2016), an optimization technique for analyzing design alternates (Ruiz-Femenia et al., 2017) and a multidimensional optimization method to concentrate on safety, environment, economics, and sustainability aspects (Guillen-Cuevas et al., 2018). Besides these formal methods, a new concept has now been introduced for the identification of hazard prevention strategies (HPS) in process designing via inherent safety assessment and can be used at the earlier design stages. The framework for this concept has been based upon the thematic analysis applied to the accident database screening leading to the formulation of the HPS (Ahmad et al., 2019).

Although all the above-mentioned approaches have been aimed at the earlier design stages, there are some common limitations, such as manual data extraction, dealing with individual chemical instead of the chemical mixture and the failure to use inherent safety guide words for safety improvement. The gaps in manual data extraction and chemical mixture have been covered through integrated risk estimation tool (iRET) by linking HYSYS and MS Excel to analyze the information (Shariff et al., 2006). The use of inherent safety guide words has been demonstrated in the extended techniques of iRET, which include process stream index (PSI) (Shariff et al., 2012), toxic release consequence analysis tool (TORCAT) (Shariff and Zaini, 2010), toxic release inherent risk assessment (TRIRA) (Shariff and Zaini, 2013) and inherent fire consequence estimation tool (IFCET) (Shariff and Wahab, 2013; Shariff et al., 2016). The ISPI method, described above, has also considered the chemical mixture for the comparison of various process routes, integrating the process simulation software ASPEN PLUS for this purpose. These techniques are capable of using the inherent safety strategies in order to prepare sustainable process designs.

Inherent assessment methods for sustainable process design are classified into two approaches: an overall approach and an individual equipment approach. All of the above-discussed methods rely upon the overall approach, whereas the individual equipment approach has not been explored extensively (Hendershot, 2010). Furthermore, individual equipment has also been indicated in the "onion model" of sustainable process design as an important level for the recent trends of sustainable process design (Martinez-Hernandez, 2017). A few methods have indeed incorporated the individual process equipment such as safety weighted hazard index (Khan et al., 2001), inherent safety index calculation (Abedi and Shahriari, 2005) and comprehensive inherent safety index (Gangadharan et al., 2013). However, these methods are not capable of accommodating the individual characteristics of process equipment (Kidam et al., 2016). Recently, a few methods have been developed to produce the inherently safer process design for heat exchanger network, process piping, and the mechanical material selection via the incorporation of dedicated characteristics (Athar et al., 2018; Athar et al., 2019b; Pasha et al., 2017). Nonetheless, the consideration of mutual characteristics shared by each individual process equipment for the inherently safer process designing has been the missing factor in the currently available methods. Therefore, a new method is presented in this paper, to deal with the mutual attributes of each process equipment for the inherently safer process design. The method is proficient in investigating the mutually shared attributes of process equipment, such as the distance between process units, inventory and other characteristics. The complete technique consists of two sections, namely 1) indexing method and 2) risk assessment and analysis, followed by incorporation of inherent safety strategies to provide a safer process design. While the indexing section helps in identifying the most critical process equipment in the chemical process, the second section ascertains the risk acceptability of the critical process equipment. For the unacceptable risk, the inherent safety principles are employed to reduce the potential and the consequences of explosions. The process simulation software, ASPEN HYSYS, and the spreadsheet tool, MS Excel are integrated into this method in order to extract and handle the process information, with the purpose of achieving inherently safer process designing.

2. ETHODOLOGY

2.1. Inherently safer process design via mutual attributes of process equipment

It is highlighted in the literature that not all equipment share the same hazard in the process and that some equipment are more hazardous than others (Heikkilä, 1999). Each process equipment contributes to a different level of hazard in the process due to the variance in the characteristics of each individual process equipment. However, there are certain attributes which are shared by each process equipment, even though their magnitude can be different depending upon the type of equipment. In the ISI method, the distance between various process equipment is scrutinized, and a generic scale is proposed regarding the safety of process equipment using the subjective indexing (Heikkilä, 1999). The score value here is not dependent upon the adjacent process equipment, which may vary in individual processes. The distance between process equipment is an equipment attribute, which does not only depend upon the considered equipment itself but also the adjacent, connected process equipment. Furthermore, there are other characteristics of process equipment mutually shared by each equipment which have not been considered in earlier works.

A chemical process becomes functional when numerous process equipment are connected with each other. This interaction is dependent upon the chemicals, the process conditions and the kind of the process equipment. The previous research on inherent safety presented in section 1 has mostly focused on the first two aspects, whereas the later prospect has been left wanting for consideration. It is precisely this aspect that is scrutinized in this paper to generate the inherently safer process designs for the chemical processes. Certain attributes of process equipment can define the individual nature of such process equipment and are classified in terms of physical and non-physical characteristics. Physical features are defined in terms of working parts, while non-physical features can be defined via inventory and ignition sources. All these attributes are shared by each process equipment, albeit the magnitude can vary for each process equipment. Inventory is initially defined as the capacity or the production scale of the process and has been termed as the process attribute (Heikkilä, 1999). However, the inventory is basically the equipment characteristic rather than the process characteristic, on the grounds that for a specific production scale of the chemical process, each process equipment has a different inventory level than the others. Since there may be other potential common attributes that have been missed, the failure rate can be employed for the definition of process equipment attributes in order to bridge the gap and cover for the missing factors. Furthermore, this failure rate is based upon the real behavior of each process equipment in chemical processes in addition to process conditions and chemical characteristics. Therefore, these parameters are jointly considered to avoid duplication. However, the failure rate considered for this present study is independent of different operating scenarios. Usually, the failure rate of process equipment operating under severe corrosive environment would be different from the process equipment dealing with mild chemicals. For this purpose, a modifier can be included in the future works to account for various operating scenarios in calculating the failure rate. Conclusively, the mutually shared equipment characteristics relevant to safety can be defined as follows:

1. The distance between process equipment
2. The nature of individual process equipment
    a) type of parts, i.e., static or moving
    b) type of ignition source
    c) equipment size using the inventory level
3. The equipment failure rates

All process equipment in a chemical process can be analyzed through the mentioned factors. The relative ranking-based indexing would help to highlight the critical process equipment, as well as reduce the efforts of process engineers during the preliminary design stage. For the purpose of identifying the critical process equipment, a new indexing method is developed in this paper, namely inherently safer intra-equipment index (I_aEI). The most critical process equipment can be further scrutinized through the risk assessment for its acceptability. For the unacceptable risk, the inherent safety themes are applied to suggest how the design can be changed so that it generates an inherently safer process design. The whole concept is named as an inherent safety assessment for process equipment (ISAPE) and will be explained in the next subsections, while the relevant framework is provided in Fig. 2.

2.2. Inherent safety intra-equipment index (I_aEI)

The subjective scale indexing method adopted in the ISI method has certain drawbacks, as mentionesd in Section 1. Therefore, in this paper, it is replaced with the relative indexing method, which has been described in the PSI method (Shariff et al., 2012). The inherently safer intra-equipment index based on the factors mentioned above can be estimated by:

$$I_aEI = A \times (I_D \times I_N \times I_{FR}) \quad (1)$$

For the smaller magnitude of I_aEI, a magnifying factor, A, can be engaged to amplify the numerical value and significantly differentiate the level for individual process equipment. This index would be estimated for all process equipment in the process; the higher the value of I_aEI, the more critical is the equipment in the chemical process and vice versa. I_aEI identifies the most critical process equipment based on the mentioned factors through the sub-indices for each aspect. In the following subsections, the calculation of these sub-indices is explained.

2.2.1. The distance between process equipment

Hazard avoidance or minimization are usually the two viable options to achieve acceptable risk. This is in accordance to the theme presented by Kidam et al., (2016) for inherent safety. In order to avoid the hazard, the distance between units is of vital importance as it would be related to domino effects to neighboring process equipment. Increasing the distance between different process equipment would minimize the effect of consequences from unwanted events on nearby process equipment in the process, without reducing the hazard magnitude. Furthermore, the distance is a key factor in defining the density of the process equipment in a chemical process. This is because with shorter distance between the process equipment, numerous equipment are closely packed either on one floor or on multi-floor. It leads to a compact layout in which the piping needs to be whirled in a small space, thus, increasing the density of process equipment in a chemical process. Guidelines are available in the literature for process equipment placements to create the best possible plant layout (CCPS, 2010a; Mannan, 2013; Moran, 2016; Vazquez-Roman and Mannan, 2010). It should be noted that these guidelines only provide the information regarding the minimum distance requirements; the actual distance required for process plant layout of any specific process is not available in the open literature. Typically, the plant layout analysis is performed at later design stages of process lifecycle (Brunoro Ahumada et al., 2018). There are a number of research works available in the literature to propose the plant layout, for example, cost and safety aspects for plant layout by genetic algorithm (Caputo et al., 2015), mixed integer nonlinear programming (MINLP) (Latifi et al., 2017), mixed integer linear programming (MILP) (Guirardello and Swaney, 2005; Patsiatzis et al., 2004; Rahman et al., 2014) and mixed integer optimization (MIO) (Xu and Papageorgiou, 2009). Although the MILP approach has been recently applied for the layout of storage vessels at the design stage (de Lira-Flores et al., 2018), it is not clearly defined whether the design stage is preliminary, basic or detailed. Indeed, a general limitation shared by all these methods is that detailed information regarding the process is seen as a pre-requisite, but it might not be available at the preliminary design stage. Considering the plant layout at the preliminary design stage, on the other hand, would be vital in creating the inherently safer process designs for chemical processing facilities. For the preliminary design stage, only process equipment inside battery limit are taken into account, since information regarding the process equipment outside the battery limit is not available from process simulation. For this reason, the process equipment outside the battery limit is excluded from this paper. The area where the raw material is converted to the product is termed as 'area inside battery limit', whereas the remaining area is known as 'outside the battery limit' (Heikkilä, 1999).

Minimum distance guidelines, available in the literature (CCPS, 2010a; Mannan, 2013; Moran, 2016; Vazquez-Roman and Mannan, 2010), have been studied and are summarized in Table 1, which can serve as the starting-point for studying the distance between different process equipment. In any chemical process, various kinds of process equipment are available; these, in turn, are connected to other process equipment. For any process equipment in the process, each connected equipment has individual characteristics; therefore, that particular process equipment needs varying distancing requirements need to be applied for each connected process equipment. For any process equipment, the distance required with each connected process equipment can be established using Table 1. Among these distances, we have selected the maximum distance value, defining the criticality level of the considered process equipment in the process. For example, in the case of a heat exchanger, the connected process equipment are two other heat exchangers: a compressor and a high hazard reactor. In this case, the distances required for this heat exchanger are 2, 2, 9 and 8 meters from the connected process equipment respectively. Among these distance values, the maximum distance for the subject heat exchanger to define the critical level in the process is 9. In a similar fashion, all the process equipment in the chemical process are studied to identify the distance required w.r.t the connected process equipment and determine its criticality. The maximum distance required for all process equipment is then converted into index value through relative ranking, followed by a mutual comparison to identify the most critical process equipment in the chemical process. The distance sub-index can be estimated by:

$$I_D = \frac{\text{Max. distance needed for an individual equipment}}{\text{Average of max. distance needed of all equipment}} \tag{2}$$

The higher the magnitude of I_D, the higher the chances for this equipment to create a severe domino effect and vice versa.

2.2.2. Nature of process equipment

Since each unit operation is based on a unique principle, the contributing hazard of each process equipment is different from the others. The attributes associated with the nature of process equipment can be defined in terms of parts type involved, related ignition source and inventory, which are all relevant for their contribution towards the hazardous scenario.

On the basis of the parts, process equipment can be categorized into two groups: static and rotary. The domino effect would be more severe for equipment containing rotating parts compared to static process equipment. For example, should there be an explosion in a compressor, the rotor of the compressor can be flung at longer distances because of its high rotation speed. Conversely, the domino effect of an explosion in a separation vessel is comparatively weaker, as no rotating parts are involved. Since this parts-based categorization of process equipment is qualitative in nature, no numerical magnitudes derived from the process simulation software are available for these scenarios. This limitation, however, can be eluded by assigning the subjective values for these characteristics, as given in Table 2. In this table, the higher numeric value indicates the more hazardous parts type, while the lower value is associated with the less hazardous ones. Each process equipment in the chemical process is assigned with a numeric value based on Table 2. It is then converted into the index value which is estimated through relative ranking with the following equation:

$$I_{PS} = \frac{\text{Part hazard level of individual equipment}}{\text{Average part hazard level of all equipment}} \tag{3}$$

The ignition source is an essential factor for explosions and other fire accident scenarios. In the process industries, ignition sources can be of several types. Nevertheless, it can be grouped into two general classes, i.e., direct and indirect. The ignition sources relevant to process equipment available in each class are:

1. Direct ignition source
    a) Flame
2. Indirect ignition source
    a) Friction
    b) Hot surface

Irrespectively of the equipment part type, there is no quantitative grading for these ignition sources. Therefore, the subjective values outlined in Table 2 are proposed to quantify the hazardous level of individual ignition sources. The higher the numeric rating, the higher the hazardous ignition source involved in the particular process equipment. For any chemical process, each process equipment is assigned a specific hazard value based on the associated ignition source. Meanwhile, the relative ranking is used to estimate the index value for this parameter. The number of process equipment in each ignition class would affect the index value of individual process equipment and it is independent of subjective indexing.

$$I_{IS} = \frac{\text{Ignition source hazard of individual equipment}}{\text{Average ignition source hazard of all equipment}} \quad (4)$$

In the case of an accident scenario, inventory leakage is the initiating event. The more the contributing inventory is involved, the more severe the consequences of the accident would be. Although the plant capacity is specified for any chemical process, each process equipment has a different level of chemicals termed as the inventory of process equipment. Therefore, the hazard contribution by individual equipment based on the inventory is different in magnitude. The actual inventory information is available if the dimensions of the equipment are known. However, this is not usually the case at the preliminary design stage. Nevertheless, this limitation can be countered by a recently presented concept byWarnasooriya and Gunasekera (2017), and the inventory information can be estimated by:

$$\text{Inventory} = \text{Flow rate} \times \text{Residence Time} \quad (5)$$

Flow rate related to each process equipment is accessible from the process simulation software ASPEN HYSYS, which is transferred to MS Excel through VBA coding. The residence time is also a function of equipment dimensions, which is missing at the preliminary design stage. For this reason, at the preliminary design stage, heuristic values of residence time for different equipment - available in the literature - can be used (Coker, 2014; Couper et al., 2012; Warnasooriya and Gunasekera, 2017), as shown in Table 3.

Once the inventory of each process equipment in the chemical process is estimated through equation (5), it can be converted into the index value using the concept of relative ranking as follows:

$$I_{INV} = \frac{\text{Inventory level of individual equipment}}{\text{Average inventory level of all equipment}} \tag{6}$$

The higher the value of I_{INV}, the more critical the process equipment, as the higher inventory translates into the higher amount of chemical involved in an unwanted scenario, with more severe consequences.

All three aforementioned characteristics namely the part type, ignition source, and inventory are combined to estimate the overall effect of equipment nature towards inherently safer process design, which can be estimated by:

$$I_N = I_{PS} \times I_{IS} \times I_{INV} \tag{7}$$

2.2.3. Equipment Failure Rates

There could be many attributes associated with process equipment, which can contribute to an accident scenario. A few of these have been identified in the previous subsections. However, there could be other equipment attributes which may have been missed and could be of vital importance for accidents. The failure rate of each process equipment can cater for all these missing factors, as each process equipment has an individual failure rate based on the associated equipment attributes as well as on the interaction of chemicals and process conditions, which depicts the actual behavior of process equipment in the process industry. An equipment is considered as failed if a leak has occurred. The leak rate of each equipment is available in the literature for different hole sizes. The failure rate of individual process equipment is shown in Table 4. The failure rate for each process equipment in the chemical process can be converted into index value using the following relationship

$$I_{FR} = \frac{\text{Failure rate of individual equipment}}{\text{Average failure rate of all equipment}} \qquad (8)$$

High I_{FR} means a high probability of failure for any equipment in the specific process. The value is dimensionless and is dependent upon the number of process equipment in the chemical process.

2.3. Risk Estimation

Risk assessment is a four-step procedure, where hazard identification is the first and most critical step (Arendt and Lorenzo, 2010; CCPS, 2000). Hazard can be identified through various methods. Past accidents analysis (PAA) is one of the tools for this purpose. In this work, the scope of risk estimation is restricted to a single accident scenario, and PAA is used to identify the possible hazard scenario from the process industry. An explosion has been identified as the most frequent accident event in the process industry, as reported in numerous accidents analysis available in the literature (Khan and Abbasi, 1999; Koteswara and Kiran, 2016; Mihailidou et al., 2012). Although for certain process equipment, such as piping and storage vessels, fire is the most frequent accident scenario (Chang and Lin, 2006; Jang et al., 2012), for an overall process plant, the explosion is the most frequent accident event (Athar et al., 2019a). Additionally, for the domino effect, it is identified by Darbra et al. (2010) that the most frequent primary scenario is an explosion which is more damaging than fire. Furthermore, among the several types of explosion scenarios, vapor cloud explosion (VCE) is identified as the most frequent scenario (based on PAA). Hence, this particular accident is used in this work, only for demonstration purposes. However, the ISAPE tool can be extended to include fire and toxicity in future works.

The explosion risk can be estimated through:

$$\text{VCE Risk} = \text{Consequences of VCE} \times \text{Probability of VCE} \tag{9}$$

2.3.1. Consequences of VCE

VCE consequences are dependent upon the blast wave overpressure, which is the function of initially released amount, as well as process and environmental conditions. The released chemical quantity depends on process conditions and type of the equipment, as each process equipment contains specific inventory level. For a small leak, a release rate needs to be estimated, while in case of full rupture of process equipment complete inventory can be assumed as the total quantity, estimated using equation (5). For a specified hole size in the process equipment, the release rate can be estimated as follows (CCPS, 2000; CEPPO, 1999; Cowley and Johnson, 1992; Tweeddale, 2003):

$$m = C_D A_h \sqrt{\gamma \rho p \left(\frac{2}{\gamma+1}\right)^{\frac{\gamma+1}{\gamma-1}}} \tag{10}$$

The next step is the estimation of flammable mass, which is the contributing factor involved in generating the overpressure. This can be computed by:

$$\frac{m_f}{m} = \exp\left[\sqrt{\ln\left(\frac{c_o}{c_{LFL}}\right)}\right] - \frac{2c_{LFL}}{c_o\sqrt{\pi}}\sqrt{\ln\left(\frac{c_o}{c_{LFL}}\right)} \tag{11}$$

Explosion overpressure is also dependent upon the energy released by the explosion and the scaled distance. Trinitrotoluene (TNT) equivalent method is used to estimate these parameters and the necessary equations are provided as follows:

$$m_{TNT} = \frac{\eta_{ex} m_f H_V}{H_{V,TNT}} \tag{12}$$

$$\bar{z} = \frac{z}{(m_{TNT})^{1/3}}$$

(13)

Overpressure can be estimated by the non-linear regression of experimental data (Shariff et al., 2006), which is demonstrated in Fig. 3. The overpressure regression equation is given as follows:

$$p_{ovr} = a_1 (b_1)^{(1/z)} z^{c_1} \qquad (14)$$

2.3.2. Frequency estimation for VCE

VCE frequency can be estimated through event tree analysis (ETA) methodology, for which the detailed procedure is available in the literature (CCPS, 2014; Moosemiller, 2011), and is discussed below. This framework of ETA is reliable to use because the intermediate probabilities are dependent upon the process conditions and the chemicals in the process rather than on the expert judgment. The frequency of VCE scenario can be estimated using:

$$f = f_{IL} \times (1 - P_{imm,ign}) \times P_{del,ign} \times P_{exp/g/ign} \qquad (15)$$

The explosion frequency is a function of the initial release frequency from process equipment and the probabilities of intermediate events. There are three intermediate events involved, namely immediate ignition, delayed ignition, as well as delayed ignition which leads to an explosion. If the ignition leads to the fire event before the accumulation of chemical mixture, then it is termed as immediate ignition (Javidi et al., 2015). The probability of immediate ignition, $P_{imm,ign}$, is dependent upon autoignition temperature and minimum ignition energy. The later aspect would identify the static discharge potential of the chemical mixture being released (Moosemiller, 2011). The probability of immediate ignition can be estimated by combining the potential of autoignition and static discharge (Moosemiller, 2011), using the following equation:

$$P_{imm.ign} = \left[1 - 5000e^{-9.5\left(\frac{T}{AIT}\right)}\right] + \left[0.0024 \times \left(\frac{(p)^{1/3}}{(MIE)^{2/3}}\right)\right] \quad (16)$$

The autoignition temperature and the minimum ignition energy for the chemical mixture can be estimated using the Le-Chatelier's mixing rule (Crowl and Louvar, 2011), through the following equations:

$$AIT_{mix} = \frac{1}{\sum_{i=1}^{n} \frac{y_i}{AIT_i}} \quad (17)$$

$$MIE_{mix} = \frac{1}{\sum_{i=1}^{n} \frac{y_i}{MIE_i}} \quad (18)$$

The absence of immediate ignition would lead to the event of the delayed ignition and is the function of ignition strength (Bob et al., 2014). It can be calculated using the following equation:

$$P_{del,ign} = 1 - \left[\left(1 - S^2\right) \times e^{-St}\right] \quad (19)$$

The ignition source and the size of the flammable cloud are the parameters to define the ignition strength. Values for different ignition conditions are provided in the literature (Bob et al., 2014; Moosemiller, 2011). There are two different probable outcomes for delayed ignition event: either explosion or the fire. For an explosion to occur, a substantial amount of flammable mass is needed to convert to a flammable cloud (Pasha et al., 2017). If an adequate quantity of flammable mass accumulates, then the probability of explosion scenario is higher. On the other hand, the probability of explosion diminishes if the quantity of flammable mass is lower. The probability of an explosion, therefore, is dependent upon the flammable mass, and can be computed using:

$$P_{exp/g/ign} = 0.024\, m_f^{0.435} \quad (20)$$

2.3.3. Risk Analysis

After the risk estimation, the analysis must proceed along with some acceptance criteria. For VCE risk, a matrix is proposed in this paper, as shown in Fig. 4. This matrix is based on the overpressure consequences, placed on the horizontal axis, and the frequency of VCE scenario reported on the vertical axis. The description of the consequences in terms of overpressure and the frequency for VCE scenario is provided in Table 5. The risk is divided into three levels: acceptable, tolerable and unacceptable. The tolerable risk is usually defined as the remaining risk after implementing all risk minimization strategies. This matrix is used to analyze the VCE risk for the critical process equipment identified in section 2.2. The identified critical process equipment is analyzed to identify whether the risk is within the acceptability range or not. For this process equipment, if the risk is beyond acceptability, the inherent safety strategies are implemented to reduce the risk to an acceptable range, as demonstrated in the ISAPE framework given in Fig. 2. Conversely, if the explosion risk is unacceptable, the inherent safety strategies mentioned in section 2.4 are employed until the risk has been reduced to be within an acceptable range before the designer can proceed.

2.4. Available ISD options

For the unacceptable risk, modifications in the design through inherent safety strategies can lead to the creation of inherently safer process designs. There are different ISD options available to modify the design by exploiting the parameters of equipment attributes. These options are as follows:

1. Increasing the distance of critical process equipment from the remaining process equipment.

This option would reduce the domino effect in case of an unwanted event, i.e. hazard avoidance. Shifting of the critical process equipment to a far distance, however, would not reduce the magnitude of the risk.

2. Reduce the size of critical process equipment.

Reducing the equipment size would minimize the inventory so that the impact of consequences can be reduced, i.e. hazard minimization. This idea can be applied in two ways:

a) reduce the capacity of the entire process plant

Not only the inventory of critical equipment is reduced, but also the capacity for the entire process is affected, which might limit the production capacity of the process and influence the economics directly.

b) divide the flow rate between parallel equipment

The inventory of the critical equipment is divided into several small-sized parallel process equipment. For any unwanted scenario in the small-sized process equipment, the amount of contributing chemical is considerably reduced. In this scenario, the production capacity of the process would not be affected, although the capital cost would escalate.

3. The combined option of reducing equipment size and increasing the distance.

The combined option would reduce the quantity of released material as well as eliminate the domino effect, without affecting the production capacity of the chemical process.

These ISD options can be studied by the design engineers for any specific process. The comparison of all these options would aid the inherently safer process design procedure, so that process engineers could pick the best design option in the greater interest of the processing facility based upon the risk criteria only. Since the parameters employed in this paper for indexing are not engaging the process conditions or the chemical properties directly; therefore, the ISAPE methodology is limited to minimization and simplification strategies only. Inventory reduction can be considered as being classified as a form of inherent safety minimization strategy, whereas the increasing the separation distance can be considered as a form of simplification strategy since it reduces the density of process equipment in a chemical process.

3. RESULTS AND DISCUSSION

3.1. Indexing through I_aEI

The complete ISAPE methodology is demonstrated through the acetone production process. Acetone can be produced using different raw materials including cumene, isopropyl alcohol, propene, carbohydrate, ethanol, and calcium acetate. For this work, isopropyl alcohol is selected as the raw material. Acetone produced from isopropyl alcohol is free from trace aromatic compounds, particularly benzene (Akram et al., 2009). In this process, an azeotropic mixture of isopropyl alcohol (IPA) and water (88 wt% IPA), stream 1, is mixed with the recycled unreacted IPA/water mixture, stream 2. This material is then pumped to heat exchanger E-401, where it is vaporized prior to entering the reactor. Heat is provided for the endothermic reaction using the circulating stream of molten salt. The reactor effluent, stream 6, containing acetone, hydrogen, water, and unreacted IPA, is cooled in two exchangers, E-402 and E-403, prior to entering the phase separator V-402. The vapor leaving the separator, stream 10, is scrubbed with water to recover additional acetone. Then the liquid, stream 14, is combined with liquid from the separator V-402, stream 11, and is sent to the separations section as stream 15. The non-condensable gases leaving the acetone scrubber, T-100, are sent off-site to the boiler plant where these are burned to recover the fuel value. Stream 15 is sent for further processing to two distillation columns, which are used to separate the acetone product in T-402 and the excess water from the unused IPA in T-403, as stream 19. The unused reactant is then recycled back to the front end of the process as a near-azeotropic mixture, after a purification process which is not considered in this study. The process simulation diagram of this process is presented in Fig. 5.

In the first step, all process equipment in the acetone process are identified. There are nine process equipment involved in this process. The next step is to recognize all the connected process equipment to ascertain the respective distance requirements for each connected process equipment using Table 1. Among the distance values of all the connected process equipment, the maximum value is picked for the process equipment which depicts the criticality level. In the case of the heat exchanger, E-401, the connected process equipment are P-401 and R-401, while the respective required distances are 5m and 8m respectively. Among these, the maximum value is selected to define the criticality level of E-401 in the process, i.e., 8m. This procedure is repeated for each process equipment in the acetone process, followed by I_D calculation for each process equipment using equation (2) which is provided in Table 6. Next, nature and failure rate sub-indices are estimated through the method explained in subsections 2.2.2 and 2.2.3. The values of these sub-indices along with sub-sub-indices of equipment nature are displayed in Table 6. Finally, the overall index value is estimated for each process equipment in the acetone process using equation (1) and presented in Table 6. In this case study, magnifying factor is not necessary as the index value is considerable in magnitude, i.e., A value is considered as 1. The prioritization of these I_aEI values for the process equipment in the acetone process has identified the distillation column T-402 as the most critical process equipment. This process equipment would be more damaging than the remaining process equipment in the acetone process because of the attributes associated and shared by the process equipment involved. This process equipment is further studied through risk estimation and analysis as presented in section 3.2, while insofar as the unacceptable risk is concerned, aforementioned ISD options would be investigated in section 3.3 to improve the safety level of the acetone process at the preliminary design stage.

3.2. Risk estimation of critical process equipment

The VCE risk for the identified process equipment, T-402, is estimated using equation (9). A full rupture scenario is assumed to fairly estimate the risk because the consequences for this scenario would be more severe. The mass released would be the total inventory of this equipment, calculated using equation (5). The rest of the calculations for consequences estimation of the explosion are performed using equations (11) to (14). The maximum distance of interest is assumed to be 100 m. The prerequisite data for the estimation of the explosion consequences for T-402 is provided in Table 7. The explosion consequences are determined by calculating the overpressure value at every 5m using equation (14). This explosion overpressure profile is displayed in Table 8. The explosion frequency for T-402 is calculated using the event tree analysis method illustrated in sub-section 2.3.2, through equation (15). The mandatory intermediate probabilities for frequency are calculated using equations (16) to (20) and are presented in Table 7. For probabilities computation, the process information available in HYSYS is transferred to MS Excel through VBA coding, while chemicals information is collected from the literature and nested in MS Excel though data entry. The estimated probability of explosion is 8.57×10^{-11} yr^{-1} using the parameters provided in Table 7, and labeled at level 1, as explained in Table 5.

To visualize how the explosion risk affects the acetone process, a tentative plot plan is prepared using the guidelines for distances between process equipment displayed in Table 1. The plot plan for the acetone process is provided in Fig. 6. The top two consequence levels are marked to portray the effect of the explosion, which is estimated by the regression of the overpressure profile. It is illustrated in Fig. 6 that most of the process equipment of the acetone process are in the highly hazardous consequence zone, while a few process equipment and equipment at outside battery limit (OSBL) are beyond the highly hazardous consequence area. This plot plan with risk mapping portrays how the explosion would affect the complete processing facility. Although the overall estimated risk for most of the process equipment lies in the acceptable/tolerable region, however, the consequences of overpressure are very high; therefore, an early stage study is necessary in order to create inherently safer process design, considering the highlighted ISD strategies in section 2.4.

3.3. Improvement through ISD

All highlighted options of ISD, as described in section 2.4, are investigated to create the inherently safer process design for an acetone production facility. All these options are applied to the process, and the tentative mapping of risk is illustrated from Fig. 7 until Fig. 10. In the first option, the identified critical process unit is placed far from the rest of the process equipment, as depicted in Fig. 7. Although the critical process equipment is distant from the rest of the process equipment, the magnitude of the risk and consequences are not reduced. Nevertheless, the domino effect is reduced to minimize the impact to nearby process equipment. This option would lead to an increase of the capital investment because of additional land and piping needed to bridge the long distance of T-402 from the other process equipment. The qualitative evaluation for this ISD opportunity is available in Table 9.

With regards to the reduced equipment size, as described earlier, there are two possible scenarios. The first option is to reduce the flow rate of the overall process through a hit-and-trial method aimed at decreasing the risk level by moderating the flow rate while maintaining the process conditions unchanged. By reducing the process throughput, the risk contours are narrowed due to the reduced amount of the mass released. The adjacent process equipment, however, are still in the highly hazardous consequence zone, as demonstrated in Fig. 8. A positive aspect regarding this ISD scenario for the acetone process is that the quality of enriched acetone is undisturbed, whereas the amount of lean acetone is disturbed sharply. The change in mass balance have affected the amount of acetone due to the new thermodynamic stability. This resulted in the maximum amount of acetone being shifted to the enriched stream. Furthermore, no additional capital investments are required to implement this option.

A variation of reduced equipment size is possible by employing small and identical sized parallel process equipment so that the output of the production facility is not compromised. In this option, the critical process equipment T-402 is replaced with two small-sized equipment connected in parallel. The inventory of original T-402 is equally divided among the two small parallel columns. These small-sized process equipment are placed in the original plot plan using the guidelines mentioned in Table 1 and the amended plot plan with the revised risk is shown in Fig. 9. In this scenario, the risk contours shrunk because of little inventory size, but a few of the nearby process equipment are still inside the highly hazardous consequence zone. With the option of small parallel equipment, the total amount and the quality of the final product is not affected, neither in enriched nor in lean acetone streams; However, the capital investment has increased significantly because of extra piping, civil structure, process equipment, and land. The data for both cases of reduced equipment size is provided in Table 9.

The last ISD option is to place the small-sized parallel columns far away from the remaining process equipment. In this option, the two small-sized columns are placed not only far from each other, but also from the other neighboring process equipment so that the domino effect is reduced. The revised plot plan, along with the risk contours for this option are presented in Fig. 10. For this ISD option, the capital investment has increased excessively, because of additional land, piping, civil structures, and extra process equipment. On the other hand, the product quality of enriched acetone and lean acetone streams remain unchanged, exactly like in the second option of reduced equipment size, because the process conditions have not been altered. Finally, a qualitative comparison of all ISD options is presented in Table 9. The most prominent feature of the risk aspect is that by incorporating any of the above-mentioned ISD strategies, only the consequences of the explosion have been reduced. On the other hand, the frequency remains the same for all the options, as highlighted in Table 9, due to the unchanged process conditions. Process designers can compare all the ISD options and make a decision based on the constraints implemented by various authorities, which may vary by location, company policies or government regulations, and may also include land availability or capital investment. In this scenario, a multi-objective optimization would be required for a better decision making by considering a number of factors such as risk, cost, material residence time, increased risks are associated with longer pipe runs, plot size availability and land costs. This optimization can be incorporated in the future works. Furthermore, the highlighted metrics of sustainability in Section 1 can be coupled with this method in the future for inherently safer sustainable process designing. Nonetheless, the current proposed method has equipped the designer for the creation of different ISD scenarios. The method also helps the designer to visualize the residual risk after implementation of ISD by considering the mutually shared attributes of individual process equipment. The well-known indices used in process industry such as Dow's Fire and Explosion Index and Chemical Exposure Index require detailed information about the process, and this data is not available at the preliminary design stage. A few examples of the detailed prerequisite information include material handling and transfer, drainage and spill control, leakage for joints and packing and hot oil heat exchange system. However, ISAPE is applicable to the preliminary design stage by considering the equipment aspects, available at the preliminary design stage and does not require the detailed information like Dow's Fire and Explosion Index, highlighted above. The chemical and process information in this method are indirectly incorporated using the failure rate data. For the presented case study of the acetone production process, taking only risk criteria into consideration, option 1 can be considered as the better option for inherently safer process design.

4. CONCLUSION

This paper has presented a new technique for generating inherently safer process design at the preliminary design stage; we have called it inherently safety assessment for process equipment (ISAPE). The method concentrates on the commonly shared attributes of process equipment, rather than on the process parameters, in the pursue of inherently safer process design. ISAPE consists of two sections: first, the indexing part, which recognizes the critical process equipment; second, further investigation of the identified critical equipment through risk assessment and analysis. Finally, for process equipment with unacceptable risk, the ISD themes are applied in order to obtain an inherently safer process design. The complete ISAPE methodology is demonstrated through a case study of the acetone production process. Although all the individual parameters have pointed to the high criticality of the individual process equipment, for the overall scenario, the most critical equipment is identified by combining all attributes of process equipment. Next, the explosion risk is assessed and analyzed against the newly developed explosion matrix for the identified critical process equipment. A tentative plot plan is used to visualize the explosion risk effect on the neighboring process equipment. Several ISD options have been applied to minimize the risk, and a qualitative comparison is conducted to select the better option. Conclusively, we argue that the ISAPE method could define the new process design norms in the industry because of its simplicity and the wide range of equipment attributes available to provide inherently safer process designs.

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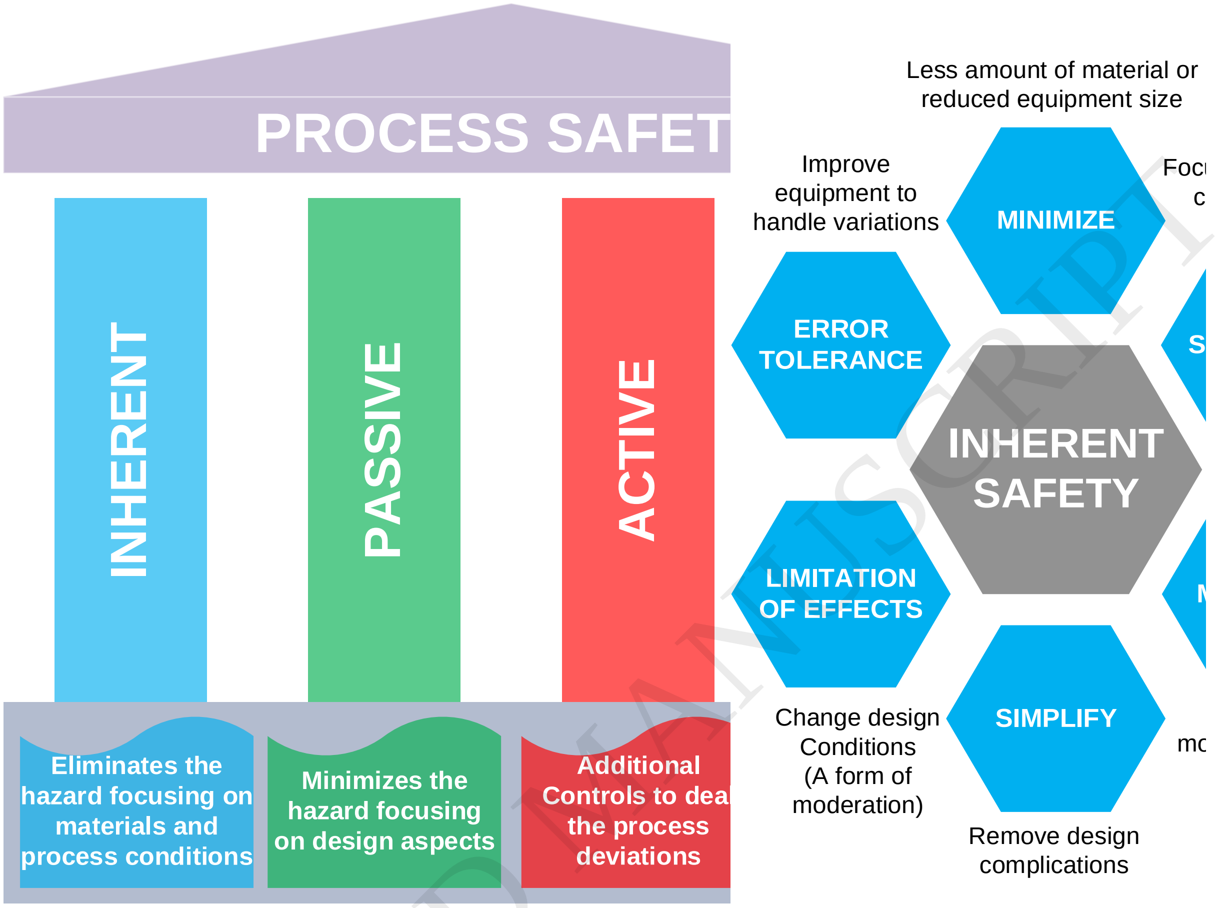

(a) (b)

Fig. 1- Pillars of (a) process safety (b) inherent safety

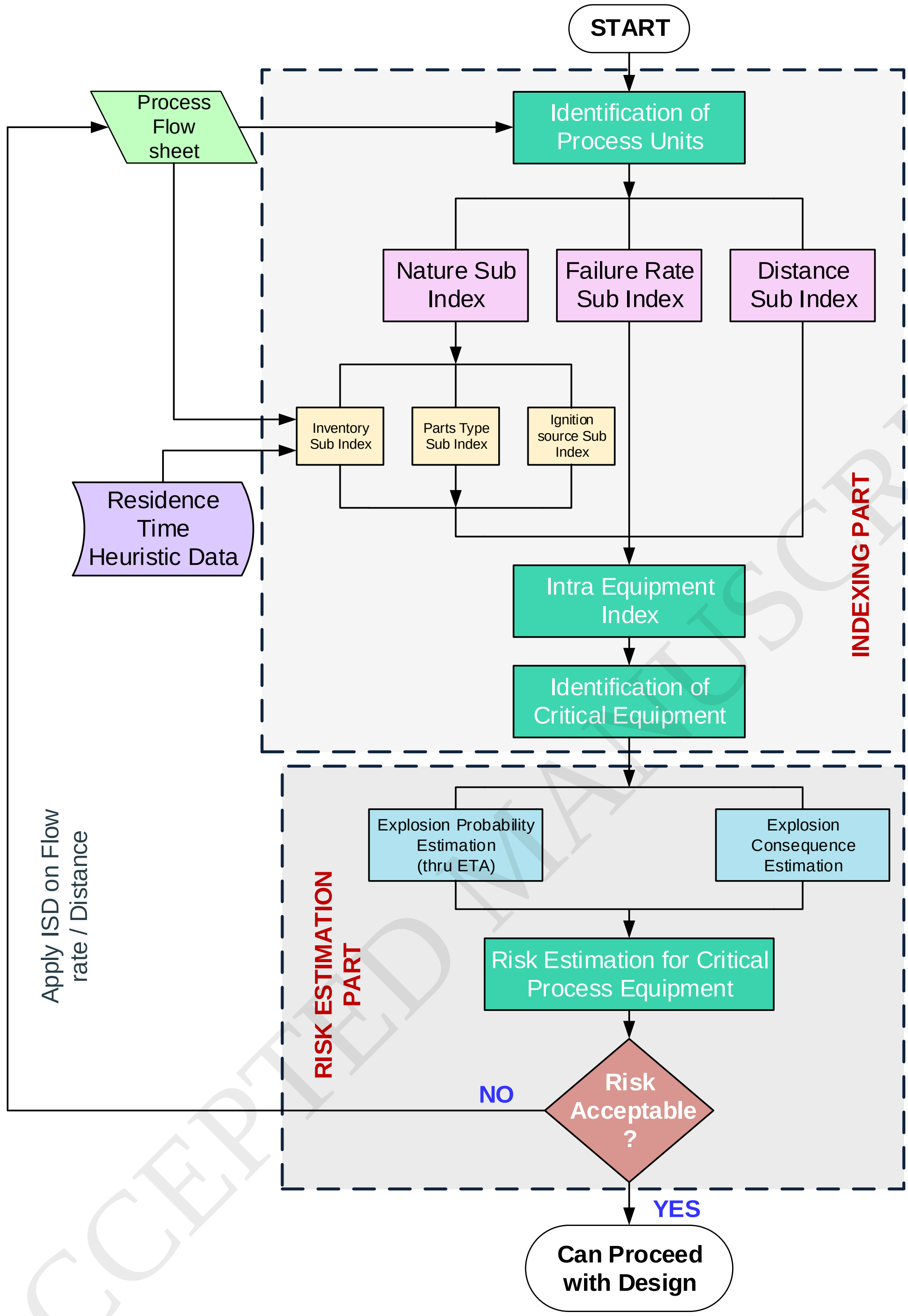

Fig. 2 - Framework for inherent safety assessment for process equipment (ISAPE)

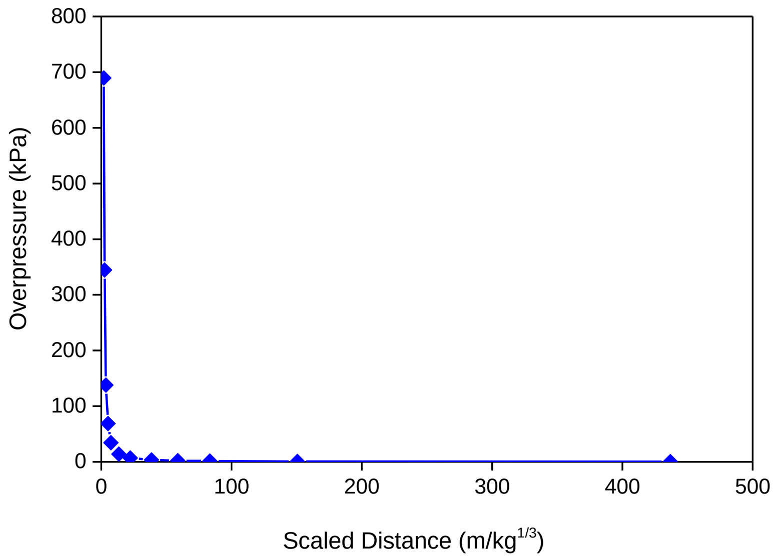

Fig. 3 - Experimental data for overpressure and scaled distance (Shariff et al., 2006)

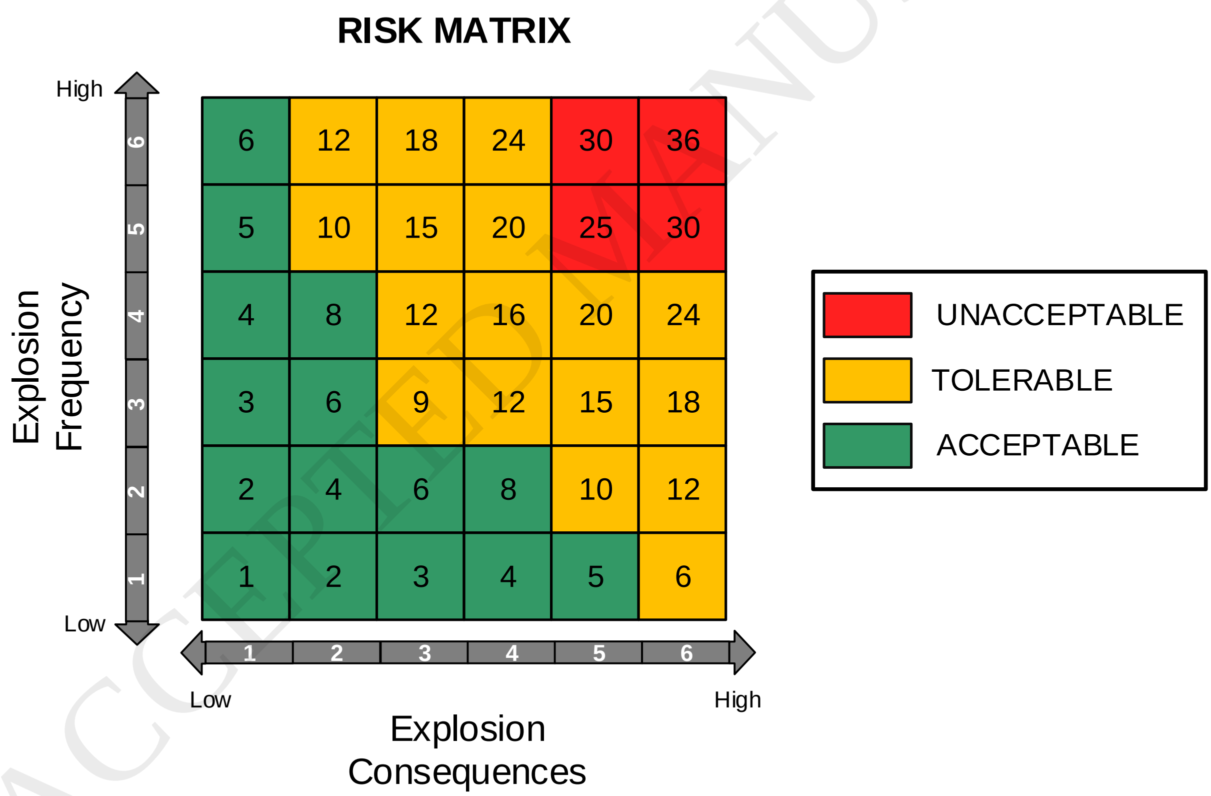

Fig. 4 - Risk Matrix for VCE

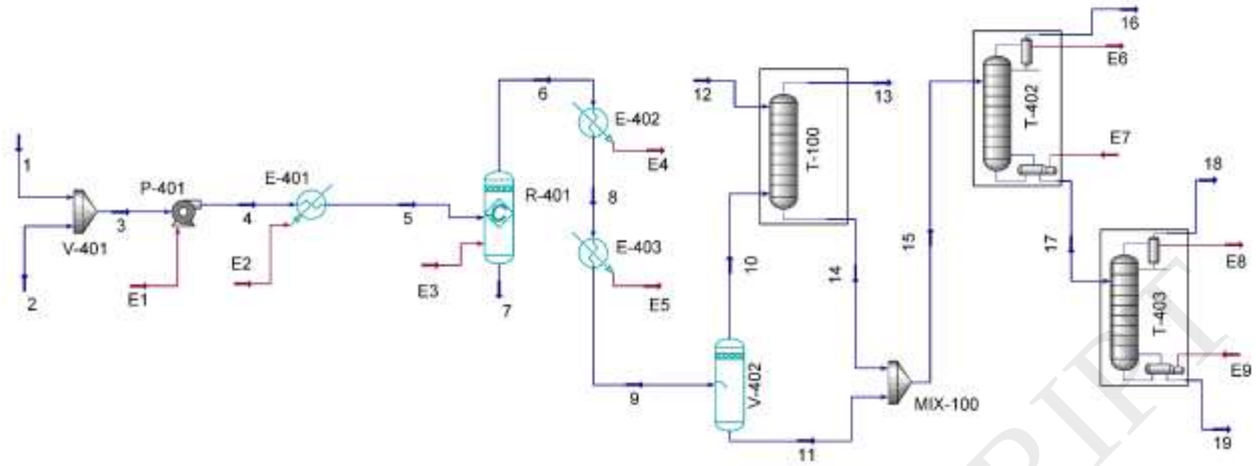

Fig. 5 - Hysys process simulation of the acetone process (Akram et al., 2009)

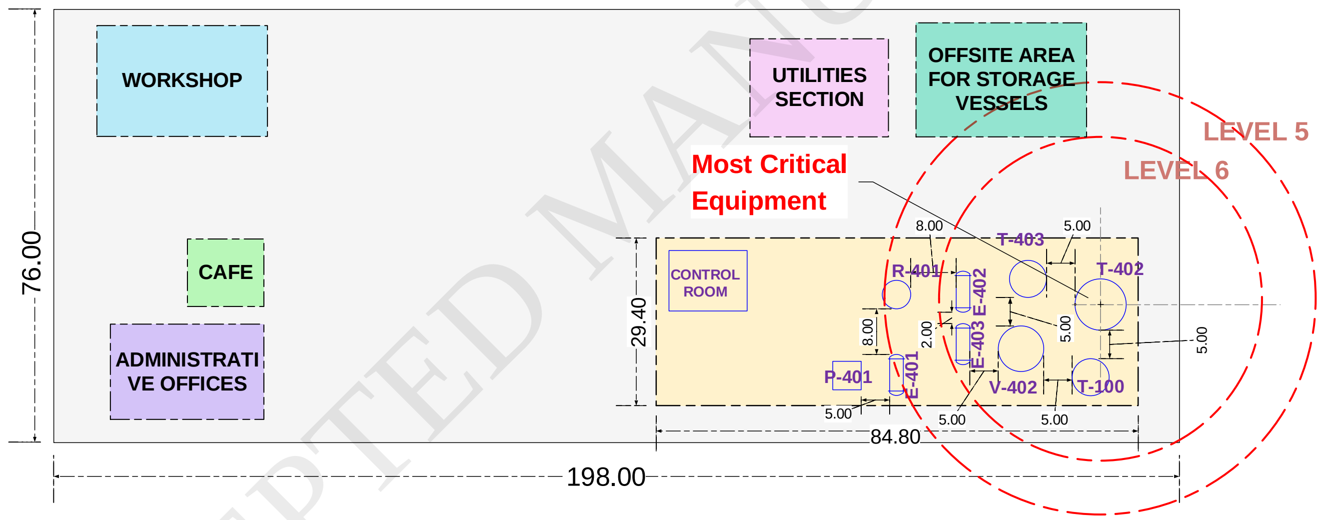

Fig. 6 - Tentative layout with unacceptable risk mapping

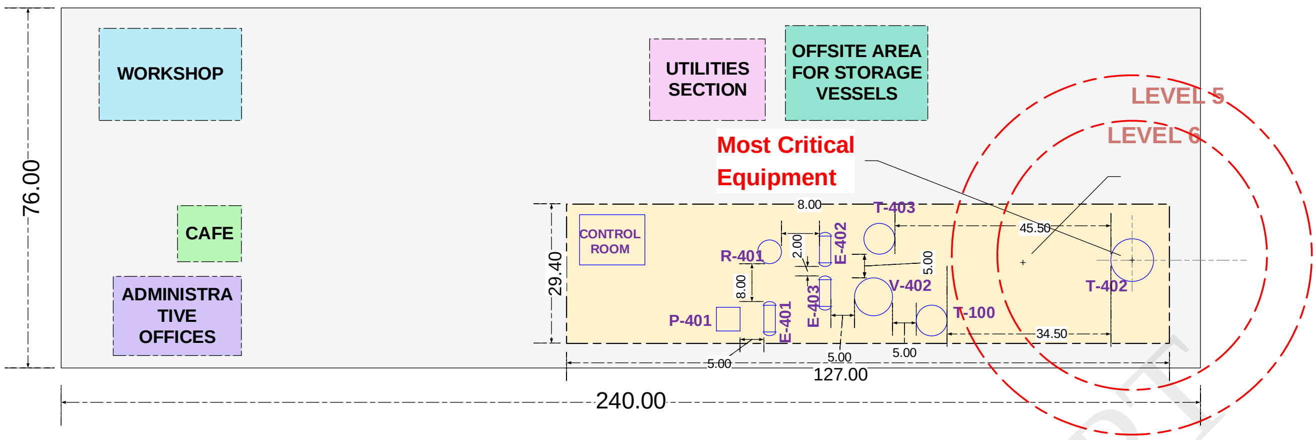

Fig. 7 - Risk mapping for acetone process after ISD option of increasing the distance between units

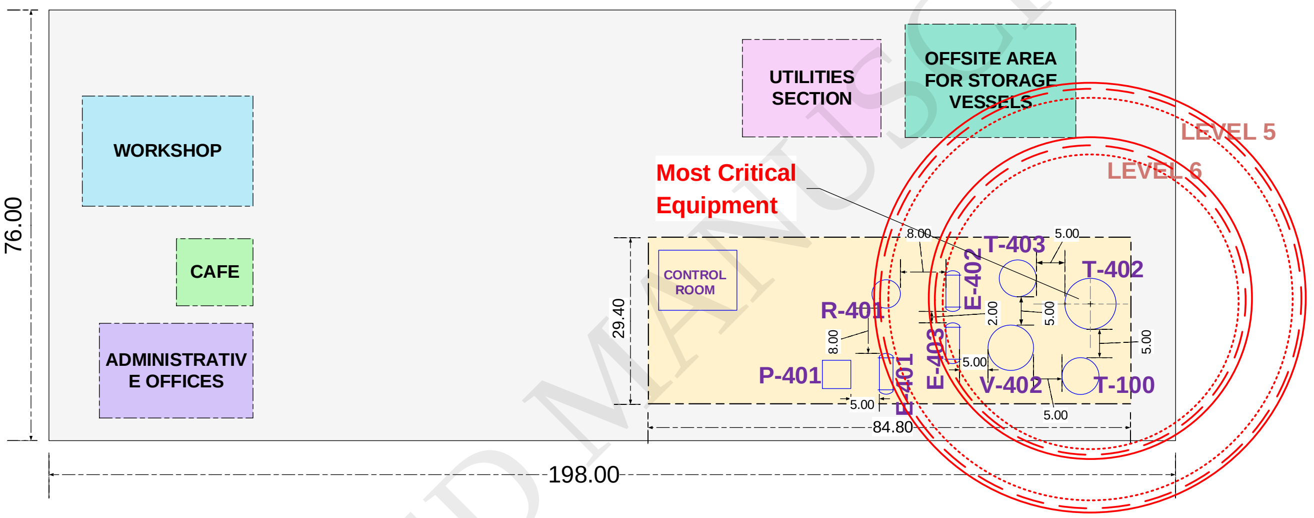

Fig. 8 - Risk mapping for acetone process after ISD option of reduced flow rate

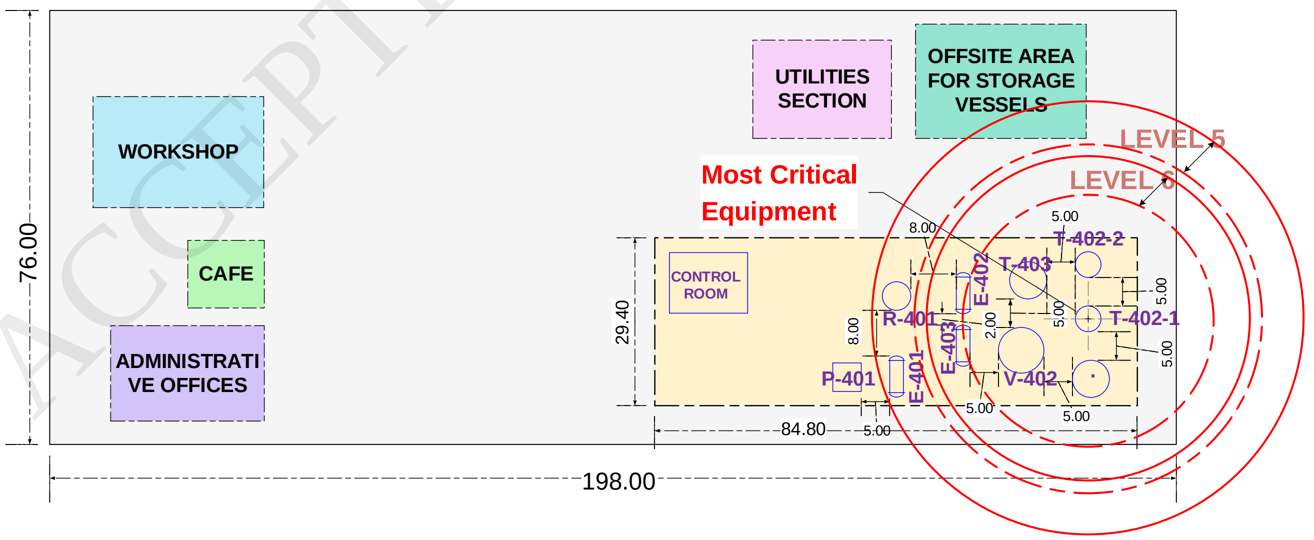

Fig. 9 - Risk mapping for acetone process after ISD option of parallel small size equipment

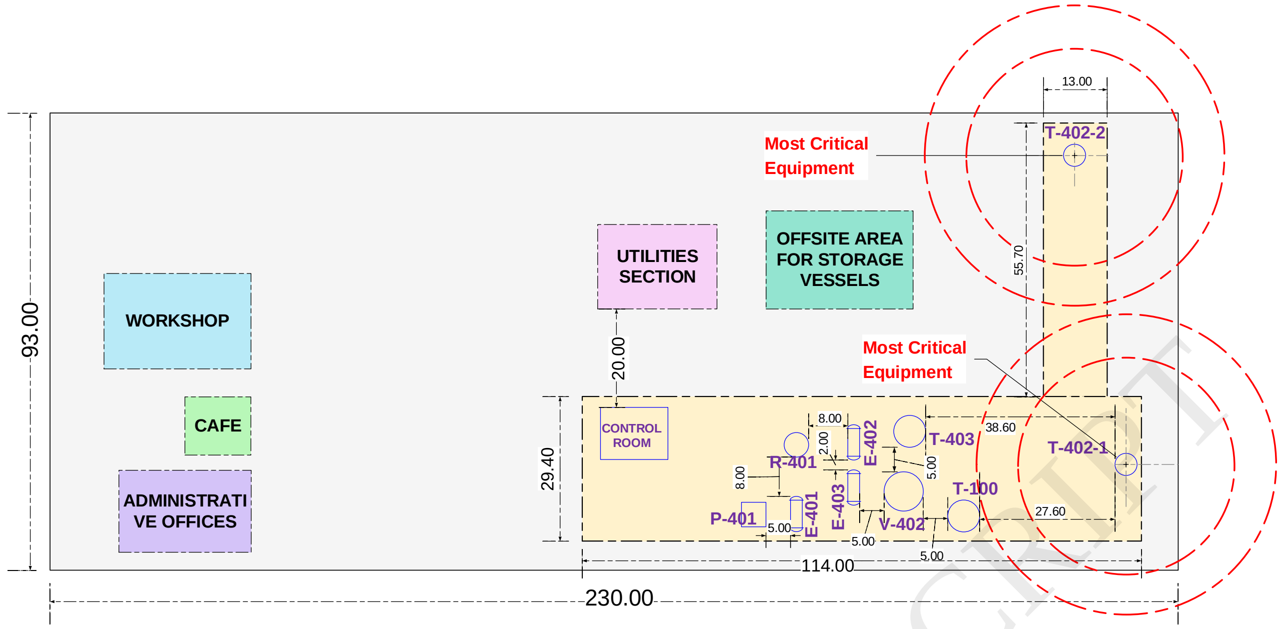

Fig. 10 - Risk mapping for acetone process after ISD option of combined small units with long distances

Table 1 - Distance between process equipment inside battery limit (in meters)

Process Equipment	**Compressors**	**Immediate Hazard Pumps**	**High Hazard Pumps**	**High Hazard Reactors**	**Intermediate Hazard Reactors**	**Moderate Hazard Reactors**	**Columns, Accumulators, Drums**	**Rundown Tanks**	**Fired Heaters, Incinerators, Oxidizers**	**Air Cooled Heat Exchanger**	**Heat Exchangers**	**Emergency Exchangers**
Compressors	9											
Immediate Hazard Pumps	9	2										
High Hazard Pumps	15	2	2									
High Hazard Reactors	15	3	5	8								
Intermediate Hazard Reactors	15	3	5	8	5							
Moderate Hazard Reactors	15	3	5	8	5	5						
Columns, Accumulators, Drums	15	3	5	15	8	8	5					
Rundown Tanks	31	31	31	31	31	31	31	31				
Fired Heaters, Incinerators, Oxidizers	15	15	15	15	15	15	15	31	8			
Air Cooled Heat Exchanger	9	5	5	8	5	5	5	31	15	-		
Heat Exchangers	9	3	5	8	5	3	3	31	15	5	2	
Emergency Exchangers	15	15	15	31	15	15	15	31	15	15	15	-

Table 2 - Subjective hazard scaling for equipment nature characteristics

Part Type	**Ignition source**	**Potential Hazard Level**
Static Parts / No Moving Parts	Friction	1
Moving Parts	Hot Surface	2
	Flame available	3

Table 3 - Residence time values of process equipment

Equipment	**Residence time**
Reactor	2 min
Vapor-Liquid Separator	5 min
Other Vapor Liquid Separation (e.g. Distillation Absorption etc.)	15 min
Pumps and Compressors*	10 sec
Heat Exchangers	13 sec (for shell) 3 sec (for Tubes)

* Holdup time for the casing of rotating process units is considered

Table 4 - Failure rates of process equipment (Moosemiller, 2011)

Equipment type	**Leak frequency (yr^{-1})**			
	1/8″–1/2″ hole	**1/2″–2″ hole**	**2″–8″ hole**	**Rupture**
Pressure vessel*	2×10^{-4}	1×10^{-4}	1×10^{-5}	1×10^{-5}
Tank	5×10^{-3}	1×10^{-3}	1×10^{-4}	2×10^{-5}
Centrifugal Pump	2×10^{-2}	4×10^{-4}	–	1×10^{-4}
Reciprocating Pump	7×10^{-2}	2×10^{-3}	–	1×10^{-3}
Centrifugal Compressor	5×10^{-3}	1×10^{-3}	–	3×10^{-5}
Reciprocating Compressor	5×10^{-2}	3×10^{-3}	–	5×10^{-4}
Heat Exchanger	1×10^{-3}	2×10^{-4}	4×10^{-5}	2×10^{-5}

* Reactors and separation columns can be considered as pressure vessels

Table 5 - Grading for explosion risk matrix

Axis Value		**Consequence levels** (CCPS, 2011; Crowl, 2010; Eckhoff, 2016; Hutchinson et al., 2012; Lobato et al., 2009; Zipf and Cashdollar, 2017)	**Likelihood levels** (Shariff and Zaini, 2013)	
	Overpressure (kPa)	**Effects**	**Occurrence Likelihood**	**Description of Occurrence**
6	≥ 138	Fatalities 100%, whole facility destroyed (buildings & process units)	$10^{-0} > P > 10^{-1}$	Very high
5	≥ 69	High fatalities, most process units destroyed	$10^{-1} > P > 10^{-2}$	High
4	≥ 34.5	High injuries with deaths, process units & buildings badly damaged	10-2 > P > 10-3	Moderate
3	≥ 21	Serious injuries with rare deaths, buildings destroyed, process units slightly damaged	$10^{-3} > P > 10^{-4}$	Low
2	≥ 14	Injuries due to secondary effects, buildings damaged but repairable, very mild effects on process units	$10^{-4} > P > 10^{-5}$	Very low
1	≥ 7	Light injuries due to secondary effects, windows damaged, no processing unit is affected	$10^{-5} > P > 10^{-6}$	Unlikely

Table 6 - I_aEI values for acetone process

Tag No	**I_D**	**Subfactors for nature**			**I_N**	**I_{FR}**	**I_aEI**
		I_{PS}	**I_{IS}**	**I_{INV}**			
P-401	0.8333	1.80	0.90	0.0467	0.0757	7.5000	0.4729
E-401	1.3333	0.90	1.80	0.0140	0.0227	0.2727	0.0083
R-401	1.3333	0.90	1.80	0.5604	0.9079	0.1364	0.1651
E-402	1.3333	0.90	1.80	0.0140	0.0227	0.2727	0.0083
E-403	0.8333	0.90	1.80	0.0140	0.0227	0.2727	0.0052
V-402	0.8333	0.90	1.80	1.4011	2.2697	0.1364	0.2579
T-100	0.8333	0.90	1.80	1.2568	2.0361	0.1364	0.2314
T-402	0.8333	0.90	1.80	4.2418	6.8717	0.1364	0.7809
T-403	0.8333	0.90	1.80	1.4511	0.0757	7.5000	0.4729

Table 7 - Parameters for T-402 VCE risk for initial design

<table>
<tr><td rowspan="2">For Consequence</td><td colspan="2">m (kg)</td><td colspan="3">m_f / m</td><td colspan="2">m_{TNT} (kg)</td></tr>
<tr><td colspan="2">750</td><td colspan="3">0.4716</td><td colspan="2">276</td></tr>
<tr><td rowspan="2">For Frequency</td><td>f_{IL}</td><td colspan="2">$P_{imm, ign}$</td><td colspan="2">$P_{del, ign}$</td><td colspan="2">$P_{exp/g/ign}$</td></tr>
<tr><td>0.00001</td><td colspan="2">0.00888</td><td colspan="2">0.00013</td><td colspan="2">0.06694</td></tr>
</table>

Table 8 - Explosion consequence profile for T-402

Actual Distance (m)	**Scaled Distance**	**Overpressure (kPa)**	**(psi)**
5.00	0.77	18,200.26	2,639.73
10.00	1.54	1,254.21	181.91
15.00	2.30	401.73	58.27
20.00	3.07	201.61	29.24
25.00	3.84	124.11	18.00
30.00	4.61	85.64	12.42
35.00	5.38	63.52	9.21
40.00	6.14	49.50	7.18
45.00	6.91	39.97	5.80
50.00	7.68	33.16	4.81
55.00	8.45	28.10	4.08
60.00	9.22	24.21	3.51
65.00	9.99	21.16	3.07
70.00	10.75	18.70	2.71
75.00	11.52	16.69	2.42
80.00	12.29	15.02	2.18
85.00	13.06	13.61	1.97
90.00	13.83	12.42	1.80
95.00	14.59	11.39	1.65
100.00	15.36	10.50	1.52

Table 9 - Qualitative evaluation of all ISD options for acetone process

Parameter	**Unit**	**Base Case**	**ISD Options**				
			Option 1	**Option 2 (a)**		**Option 2 (b)**	**Option 3**
			Increase the distance between units	**Reduce the flow rate**		**Divide the Flow into Parallel Equipment**	**Parallel Small Equipment with Increase in distance**
				10% Reduced	**20% Reduced**		
Product Amount							
Feed Flow Rate (Stream 1)	kg/hr	2,630.00	2,630.00	2,367.00	2,104.00	2,630.00	2,630.00
Feed Flow Rate (Stream 2)	kg/hr	344.30	344.30	304.87	275.44	344.30	344.30
Product Flow Rate (Stream 16)	kg/hr	1,974.00	1,974.00	1,974.00	1,975.00	1,974.00	1,974.00
Acetone Fraction (Stream 16)	-	0.91	0.91	0.91	0.91	0.91	0.91
Product Flow Rate (Stream 18)	kg/hr	408.80	408.80	375.90	202.90	408.80	408.80
Acetone Fraction (Stream 18)	-	0.89	0.89	0.63	0.08	0.89	0.89
Consequence Part for Risk							
Inventory of T-402	kg	750.00	750.00	674.25	600.25	375.27	375.27
Zone 6 Distance	m	28.24	28.24	27.25	26.13	22.10	22.10
Zone 5 Distance	m	38.04	38.04	36.87	35.53	30.59	30.59
Other Process Equipment in consequence zone 5/6	-	Yes	No	Yes	Yes	Yes	No
Frequency Part for Risk							
$P_{imm, ign}$	-	0.00888	0.00888	0.00888	0.00888	0.00888	0.00888
$P_{del, ign}$	-	0.00013	0.00013	0.00013	0.00013	0.00013	0.00013
$P_{exp/g/ign}$	-	0.06694	0.06694	0.06694	0.06694	0.06694	0.06694

Frequency	yr^{-1}	8.57×10^{-11}	8.57×10^{-11}	8.57×10^{-11}	8.57×10^{-11}	8.57×10^{-11}	8.57×10^{-11}
Qualitative Evaluation for Capital Expenditure							
Total Area Required	m^2	15,048.00	18,240.00	15,048.00	15,048.00	15,048.00	21,390.00
Operating Area Required	m^2	2,493.12	3,733.80	2,493.12	2,493.12	2,493.12	3,351.60
Additional Structure Cost Applicable	-	NA	Yes (Only Piping)	No	No	Yes (Piping, Civil Structure, Fabrication of Vessel)	Yes (Piping, Civil Structure, Fabrication of Vessel)