METHODOLOGY FOR INHERENT SAFETY ASSESSMENT OF HEAT EXCHANGERS IN EARLY DESIGN STAGES

A Thesis

by

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ABSTRACT

Shell and tube heat exchangers are used extensively in the process industry; they are a part of every process and make up for significant part of the capital cost investment. Due to high number of heat exchangers present in plant, their frequent failure is also a continuous problem. The heat exchanger failures can lead to lower production, unplanned shutdown and in some cases injury or loss of life. Insufficient safety analysis and lack of risk assessment followed by inherent safety considerations in the initial phases of design are the primary reason behind frequent failure of these exchangers. Investigations of previous incidents indicate that the failures could have been avoided, if appropriate safety assessment of the equipment was carried out in the basic engineering phase.

This study focuses on underlying reason behind frequent failure of shell & tube heat exchangers and develops a methodology for inherent safety quantification of these exchangers in early design stage. The current practice is to use QRA for safety assessment of heat exchangers, which can only be done at later stages in project development, and by this time the opportunity to implement inherent safety principles is minimum and expensive. In this work, an index has been developed which incorporates the safety aspect of metallurgy selected and possible interaction between the material of construction and selected process chemical. Additionally, a framework has been proposed which provides systematic evaluation method for equipment safety along with process safety in the basic design phase of the project.

The developed methodology was applied to a fire and explosion which occurred at Tesoro’s, Anacortes refinery due to catastrophic rupture of heat exchanger and resulted in seven fatalities.

An important conclusion that can be drawn from this work is that inherent safety principles can be applied to metallurgy selection process of equipment, which is typically during pre-design and basic engineering phase of the project. Then utilizing the information available inherent safety level assessment of both process and equipment can be carried out in the early design stage.

DEDICATION

My parents,

My teachers,

My siblings,

My friends,

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NOMENCLATURE

ISD	Inherently Safer Design
PA	Pinch Analysis
HE	Heat Exchanger
HEN	Heat Exchanger Network
STHE	Shell & Tube Heat Exchanger
CMTD	Corrected Mean Temperature Difference
IRA	Inherent Risk Assessment
QRA	Quantitative Risk Assessment
ISI	Inherent Safety Index
PIIS	Prototype Index of Inherent Safety
PRI	Process Route Index
PSI	Process Stream Index
F&EI	Fire & Explosion Index
2TISI	Two Tier Inherent Safety Index
ESI	Equipment Safety Index
HESI	Heat Exchanger Safety Index
HENOSI	Heat Exchanger Network Overall Safety Index
W_{HESI}	Worst Heat Exchanger Index
MISF	Modified Inherent Safety Framework
MOC	Material of Construction

HTHA	High Temperature Hydrogen Attack
NHT	Naphtha Hydrotreater
ALARP	As Low As Reasonably Practicable
I_F	Flammability index
I_{EX}	Explosiveness index
I_P	Pressure index
I_T	Temperature index
I_M	Metallurgy index

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1. INTRODUCTION

1.1 Introduction

Heat exchangers are extensively used in process industries, in the form of cooler, heater, condenser, evaporator, boiler, etc. They contribute towards a big part of capital investment during the initial stages and operating expenditure throughout the life cycle. Heat integration/optimization to reduce energy consumption is the constant focus of the industries. Heat Exchangers Networks (HEN) is well-known concept to achieve this minimization of energy consumption. Pinch Analysis (PA) is a well-researched methodology that provides options maximizing heat recovery and minimizing heating and cooling requirements, while simultaneously minimizing the number of heat exchangers. However, while PA has become a significant tool in achieving energy saving, safety aspect of the final HEN is rarely incorporated during the HEN design phase and additional engineering/managerial controls are later added as required using the conventional method, e.g. through HAZOP, which can increase the overall cost significantly. In addition, heat integration brings different process streams together for energy optimization, and while selecting hot and cold streams during HEN design, if due consideration is not provided to safety aspect of the HEN, in terms of both – compatibility of shell side and tube side process fluid, and metallurgy suitability for different process fluids, it can introduce additional operation hazards which would necessitate further investment on add-on safety features.

The most common safety issues associated with HE design are possibility of contamination, leakage, reaction between shell and tube fluid leading to runaway reaction or in extreme cases explosion, selection of inappropriate material of construction leading to mechanical failure. Including equipment safety along with process safety in the initial phase of design can reduce these risks significantly and thus will reduce the requirement of additional safety features, thus making the design inherently safer.

Inherently Safer Design (ISD) is a concept that provides a way to enhance process safety by introducing fundamentally safer characteristics into the development of a process. This concept can be applied at various stages in the life cycle of a plant – from early process invention and research through development, plant design, operation, to eventual shutdown and demolition. The concept of Inherently Safer Design was first introduced by Dr Trevor Kletz in 1978 in an article "What you don't have, can't leak" in the 19th Loss prevention symposium of the American Institute of Chemical Engineers, based on lessons learnt from Flixborough disaster (Kletz, 1978). The fundamental concept behind ISD is making process safer by "inherent" nature of the process and not by initially accepting hazards and then then relying on added on safety features as followed in conventional methods.

The aim of this study is to propose an index/framework which can be utilized during the design of Shell & Tube Heat Exchanger (STHE) and also for development of HEN which are inherently safer. The literature review which was carried out while developing this framework, application of the methodology to case studies, results obtained and subsequent analysis, followed by conclusions and future work are described in the following chapters.

1.2 Motivation

Shell and Tube Heat Exchangers (STHE) are extensively used in various processes. The continuous demand for cost minimization and energy optimization results presence of significant number of Heat Exchanger Networks (HEN) across the site. Unfortunately the failure of these STHE are also frequently observed, these failures can lead to toxic chemical releases to atmosphere, fire and in extreme cases explosions which then results in unfavorable and sometimes catastrophic consequences like production loss, plant damage, injuries and fatalities.

The largest fatal incident in US petroleum refinery since BP Texas city incident in 2005, was the catastrophic explosion that occurred at the Naphtha Hydrotreater Plant of Tesoro Anacortes refinery in April, 2010. This incident claimed seven lives and the investigation report led by CSB revealed that the incident could have been avoided by modifying the convention design of STHE using ISD principles. (CSB, 2014)

1.3 Inherently Safer Design

Inherent Safety is the design philosophy primarily based on reduction and elimination of hazards. (Mannan, 2002)

The fundamental principles of Inherently Safer Design (ISD) can be described as follows:

- Substitution: substitute a hazardous chemical in the process with a safer alternative. Hazards associated with a chemical can be described by using flammability potential, reactivity, toxicity and explosiveness
- Minimization / Intensification: use the smallest quantity of hazardous materials feasible for the process, reduce the size of equipment operating under hazardous conditions, such as high temperature or pressure

- Attenuation or moderation: reduce hazards by dilution, refrigeration, process alternatives that operate at less-hazardous conditions; reduce the potential impact of an accident by siting hazardous facilities remotely from people and other property
- Simplification: eliminate unnecessary complexity, design simpler plants.

1.4 Application of Inherent Safety Principles – Current Status

The concept of inherent safety principles is deep rooted within various regulations and recommended practice guidelines issued by authoritative bodies like EPA (Environmental Protection Agency), CCPS (Center for Chemical Process Safety), AIChe (American Institute of Chemical Engineers), NSC (National Safety Council) and ACS (American Chemical Society).

There has been significant effort from major corporations as well to promote the development of inherently safer chemical processes and products (Khan & Amyotte, 2003). Dow developed Dow F&EI and Dow Chemical Exposure Index as relative risk ranking tool utilizing inherent safety principles. Exxon Chemical review process is based on life cycle approach (Khan & Amyotte, 2003). Rohm and Haas developed Major Accident Prevention Program which is a four-step process and works on the principal of consequence analysis for credible events and checklists for reduction of hazards. (Renshaw, 1990)

2. PREVIOUS RESEARCH ON INHERENTLY SAFER DESIGN OF HEAT EXCHANGERS

2.1 Indices for Inherently Safer Process Design

Significant research has been done to integrate the inherent safety philosophy in the design phase of a process. Several safety indices and assessment methods have been developed to understand the potential hazards associated with various process alternatives available during the design phase, since the since the conceptualization of ISD by Dr Trevor Kletz in 1978.

Prototype Index of Inherent Safety (PIIS) by (Edwards & Lawrence, 1993) was the first index published for evaluating inherent safety of a process. This index only considers the reactions steps and raw material used. PIIS is calculated as a sum of process score and chemical score. The sub-indices for chemical score are toxicity, flammability, explosiveness and inventory; the sub-indices for process score are temperature, pressure and yield. (Rahman, Heikkilä, & Hurme, 2005).

Dow F&E index and Mond index developed by Dow, and latest revision published in 1994, is one of the many safety indices developed to quantify and assess hazards associated with a process. These indices primarily consider fire and explosion hazards and are widely accepted and used in the industry for hazard identification at plant level. They have undergone subsequent revisions to be a reliable indicator of fire and explosion hazards. But limitation of data availability at conceptual stage and various types of hazards present at a plant, apart from fire and explosion, (e.g. toxicity, runaway reactions, decomposition, equipment health/reliability etc.), indicated and motivated the need for more research towards quantifying safety considerations and thus development of many other safety indices which can be applied at much earlier stages of process design.

Inherent safety index (ISI) developed by (Heikkilä, 1999) provides a much simpler calculation technique for ISL assessment, with information available during predesign stage. This index takes into account much larger scope of process steps – it includes evaluation of 12 parameters, most of which can be estimated by using the physical and chemical properties of the material being used and operating conditions. An additional sub-index allows inclusion of experience based evaluation of process structure. This index also includes equipment safety sub-index apart from the other general process safety sub-indices. The overall ISI is calculated as summation of inherent chemical safety index and inherent process safety index.

i-safe index developed by (Palaniappan, Srinivasan, & Tan, 2002) uses sub-indices from ISI and PIIS and additional NFPA reactivity rating in the Individual Chemical Index (ICI). In this method process route is compared based on Overall Safety Index (OSI) which is calculated by Overall Chemical Index (OCI) and Overall Reaction Index (ORI), in addition it provides three more indices – Worst Chemical Index (WCI), Total Chemical Index (TCI) and Worst Reaction Index (WRI). These additional sub-indices are used for process evaluation when two process routes have similar OSI. An automated tool was also developed as part of this which helps in inherently safer route selection and flowsheet development.

Another significant improvement towards quantification of inherent safety was development of inherent safety index based on fuzzy logic by (Gentile, Rogers, & Mannan, 2003). This method attempts to address the subjective nature of information used in ISI method.

Safety Weighted Hazard Index (SWeHI) was developed by combining process hazards and available safety measures in the final hazard assessment score to provide an overview of the plant safety level. (Khan, Hussain, & Abbasi, 2001)

One of the latest indices for inherent safety assessment is Process Stream Index (PSI) which is utilized once a chemical process route has been selected and considers the process stream as a mixture and not individual component, as is generally the case with other published indices. This method uses relative ranking of a process stream against others streams in the process route and represents the inherent safety level, from the perspective of explosion, of process streams during simulation work. (Chan, Alwi, Hassim, Manan, & Klemeš, 2014)

Table 1 describes the information which is utilized in all the different indices that was explained in this section.

Safety Index	Requirements
Dow F&EI	reactivity, flammability or combustion potential, heat of combustion, reaction – endothermic/exothermic, toxicity, corrosion/erosion, plot plan, process flow sheet, pressure, temperature, leakage around joints & packing, inventory, emergency equipment access
Mond Index	reactivity, ignition sensitivity, spontaneous heating & polymerization, toxicity, explosiveness, physical changes, material transfer & handling, process conditions, layout spacing
PIIS	flammability, explosiveness, toxicity, inventory, temperature, pressure, yield
ISI	Heat of reaction – main reaction/side reaction, chemical interaction – with air/water/other process chemical/MOC, flammability, explosiveness, toxicity, corrosivity, inventory, temperature, pressure, equipment, process structure
i-safe	Flammability, toxicity, explosiveness, NFPA reactivity rating, temperature, pressure, yield, heat of reaction
PRI	stream composition – not pure component, density, pressure, energy, combustibility

Table 1 Input requirement for various safety indices calculation. Adapted from Rahman, Heikkilä, & Hurme, 2005

In 2005, a benchmarking study was carried out by (Rahman, Heikkilä, & Hurme, 2005) for the three most accepted inherent safety indices – PIIS, i-safe and ISI, using the methyl methacrylate process and the outcome was compared against expert judgment. Depending on the index used there was a difference of 10-15% in subprocess evaluations from expert values. For process route evaluation, the difference was about 4 to 10% from expert opinion, based on the safety index used. The comparison study noted that ISI is the most elaborate and accurate when compared with expert values. These results also help in concluding that safety indices can be successfully utilized for process route evaluations with reasonable accuracy. (Leong & Shariff, 2009). A review of development of various safety indices with emphasis on application in design phase of a project is well covered by (Roy, et al., 2016).

2.2 Safety Analysis at Different Design Phases

Dr Kletz had stated that implementing inherent safety principles becomes more and more difficult as the design of the process plant progresses. This is due to the absence of information at preliminary design stages which complicates safety considerations. As the process design develops, more information is made available however carrying out safety analysis in later phases of design, limits the option of making process inherently safer and safety is achieved by means of add-on engineering protection and management protection techniques. This phenomenon referred to as design paradox by (Hurme & Rahman, 2005) is illustrated in figure 1.

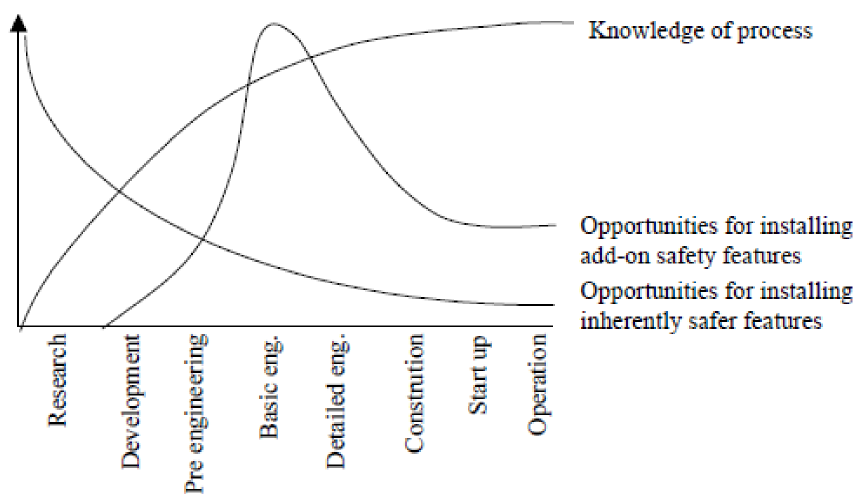

Figure 1 Design paradox and ISD. Adapted from Hurme & Rahman, 2005

Different evaluation techniques appropriate during different stages of process design, based on the information available is described in table 2.

Design stage	Information available	Evaluation technique
Process R&D	physical and chemical properties of the selected raw material, chemical reactions and interactions, thermodynamics, preliminary process concept	Laboratory screening and testing
Predesign	mass balance, energy balance, process concept, operating conditions, preliminary layout sketch	ISI, PIIS, i-safe, PRI
Basic engineering	process data on equipment, piping and instruments, procedures – normal operation/start-up/shut-down, preliminary layout	Dow F&E Index, Mond index, PSI, ESI
Detailed engineering	detailed engineering data for equipment, piping, instruments, controls, electricals, constructions, structure, layout of the plant	Dow F&EI, Mond index
Construction	vendor data for equipments, as built data	What-if, Checklist
Start-up	process performance data, commissioning data	What-if, Checklist
Operation	operation data and experience	Dow F&EI, Mond index,

Table 2 Inherent safety evaluation techniques at different process design stages. Adapted from Heikkilä, 1999

Though the idea of ISD had been initiated a long time ago, it was not successfully integrated with basic engineering stage due to lack of systematic methodology (Jha, Pasha, & Zaini, 2016). A significant contribution towards integration of ISD principle in basic engineering/simulation stage was the development of systematic approach framework – Two-tier inherent safety index (2TISI) shown in figure 2. (Leong & Shariff, 2009). The aim of this framework is to first select inherently safer process chemical route by using Process Route Index (PRI), the second step is to carryout Inherent Risk Assessment (IRA) to analyze consequent impact and frequency of credible event (explosion) for the selected process route. If the risk is found unacceptable, further inherent safety level (ISL) assessment of process streams is carried out by using Process Stream Index (PSI) (Shariff, Leong, & Zaini, 2012) at the preliminary design stage. Inherent Risk Assessment (IRA) concept introduced by (Shariff & Leong, 2009) and utilized in this framework takes the structured approach of QRA and implements them at early phases of design to enable the assessment, control and reduction of risk using the principles of inherent safety and safety indices.

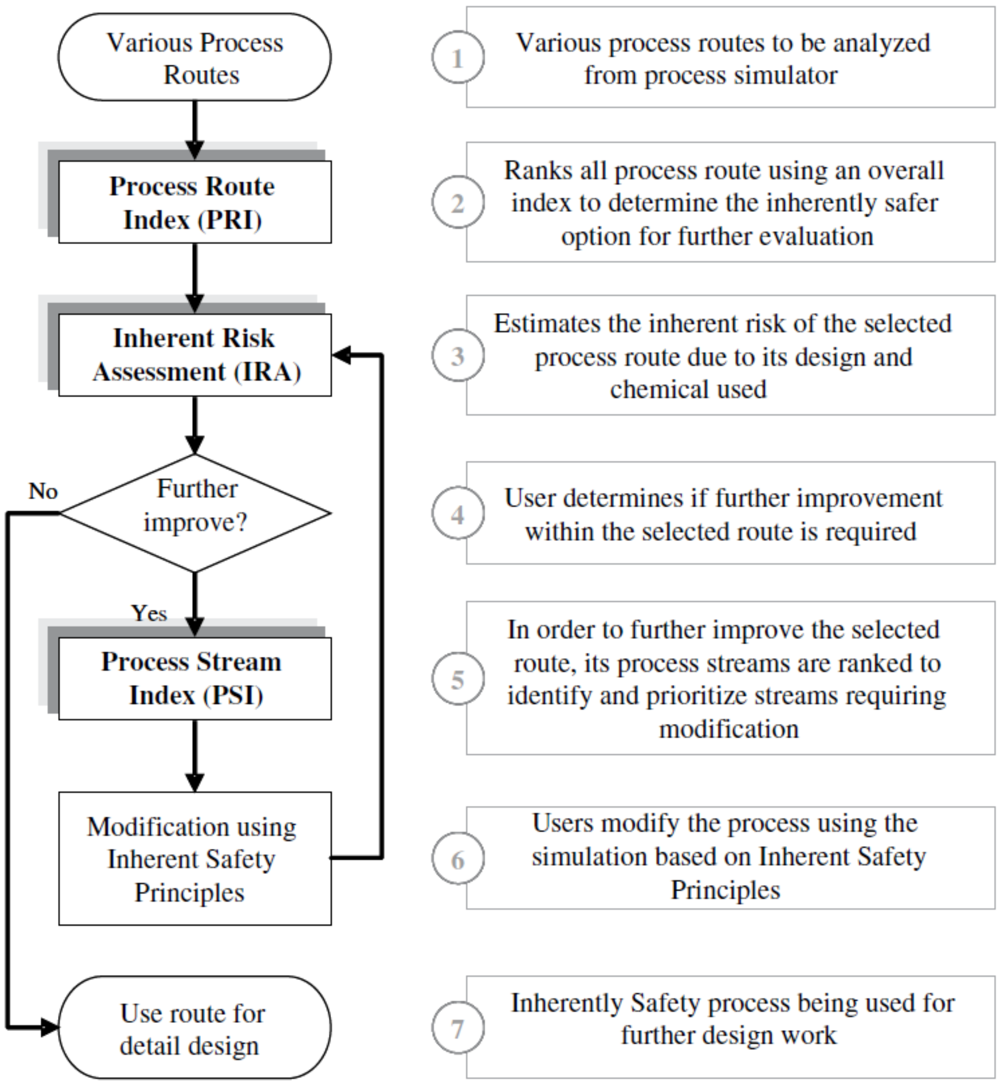

Figure 2 Framework of two-tier inherent safety index (2TISI). Adapted from Leong & Shariff, 2009

2.3 Common Causes of Heat Exchanger Failure

Lack of inherently safer design considerations combined with inadequate mechanical safety analysis are one of the most common causes associated with failure of STHE. Failure of STHE can be caused by several factors e.g., corrosion, mechanical vibrations, design faults, fabrication issues, inappropriate material of construction, flow and heat transfer related issues. As the failure can be combined outcome of several variables, it becomes difficult to carry out assessment of some of these failures during early design stage. Flow, thermal instabilities and inappropriate metallurgy are frequently reported as causal factor of STHE failures. Flow and thermal instabilities are generally associated with process variables like – pressure, temperature and flow velocity. To estimate flow velocity during preliminary design stage is difficult as it requires actual configuration of STHE, however pressure, temperature and combustibility potential values are available during preliminary design stage. Corrected mean temperature difference can be easily estimated using simulation. Using this information explosion potential and consequence of such explosion can be estimated during preliminary design stage. (Pasha, 2017).

A database study carried out in 2013, to identify and categorize the reasons behind chemical process industry accidents based on equipments. 364 incidents from the Japanese Failure Knowledge Database were studied. This study showed heat transfer equipments are the third most common cause behind incidents. Overall heat transfer equipments account for on average 8% of the chemical process incidents, 4% of these incidents are due to heat exchanger failures. Further detailed analysis behind the failure of heat transfer equipments indicates process contamination, wrong metallurgy selection and corrosion as few of the common causes. (Kidam & Hurme, 2013)

2.4 Heat Exchanger Network Design and Inherent Safety

Typically safety aspect for a HEN is considered when HAZOP is conducted on a completed HEN design, which results in add-on safety measures and increase in investment. There has been growing interest and some significant progress done in HEN design considering inherent safety. A methodology integrating the ISI with the STEP graphical approach to achieve an inherently safer HEN design is discussed in (Chan, Alwi, Hassim, Manan, & Klemeš, 2014). This method constructs the STEP by matching the hot and cold stream based on the ISI value of the stream, in place of utilizing conventional FC_p values. High ISI value hot stream is matched with high ISI cold stream, thus reducing the distribution of hazard throughout HEN and thus reduces the total area of hazard.

Another Pinch Analysis based safety assessment technique for optimal HEN design was proposed by (Hafizan, Alwi, Mannan, & Klemes, 2016), this method modifies the ISI based stream matching proposed by (Chan, Alwi, Hassim, Manan, & Klemeš, 2014) and extends it further to include operability of HEN – in terms of flexibility and controllability of HEN, while considering inherent safety.

Most recent study on inherently safer design of HEN was proposed by (Pasha, 2017), in this technique new safety indices are developed for assessing inherent safety level of individual heat exchangers and an overall safety index for HEN. This index considers pressure, CMTD, heating value and combustibility potential of the stream. Risk assessment in the event of explosion is then carried out, using IRA method and if found outside the defined acceptable range, inherent safety principles are applied to modify the network design. The developed methodology is also linked with process simulation tool HYSYS for ease of data transfer and assessment.

2.5 QRA vs IRA

A conventional QRA is typically carried out during later phase of design, when plant design has been completed. At this late stage in design opportunities to include inherent safety design principles is very low and could increase the cost considerably. Also important to consider is that QRA requires an estimated duration of anywhere between 40 and 1500 manhours, depending on the level of detail. Owing to this extensive nature QRA generally covers few selected cases and very specific elements of the aspects involved in the overall safety of the plant. (Shariff & Leong, 2009). The output of both QRA and IRA is judged based on the FN curve. In case of QRA a 3-region FN curve is used, while for IRA this has been modified to 2-region FN curve. The “tolerable if ALARP” and “tolerable” division of QRA is merged to one division of “tolerable”. This is to account for unavailability of safety measures and control mechanisms data during pre-design stage of a project, which are required to estimate and reduce risk to ALARP. (Shariff & Leong, 2009). Typical FN curves which are used for QRA and IRA methods are shown in figure 3 & 4.

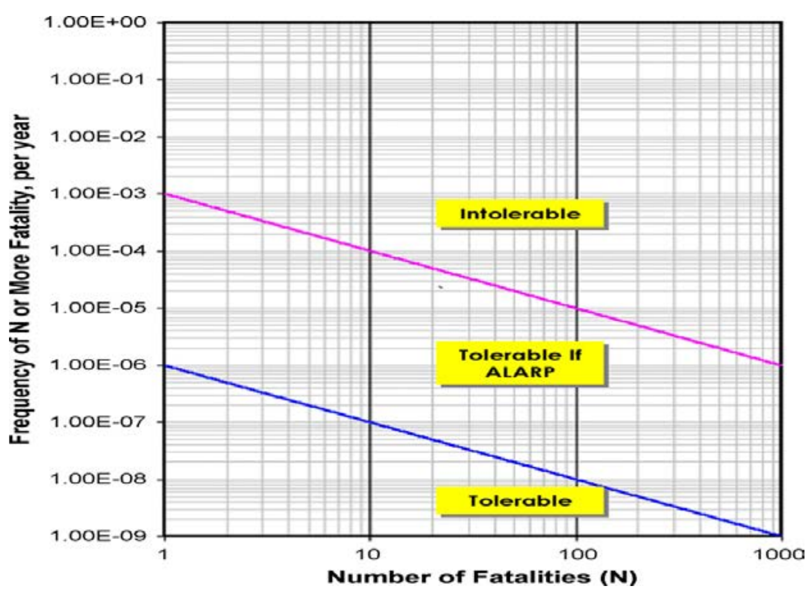

Figure 3 3-region FN curve – QRA. Adapted from Shariff & Leong, 2009

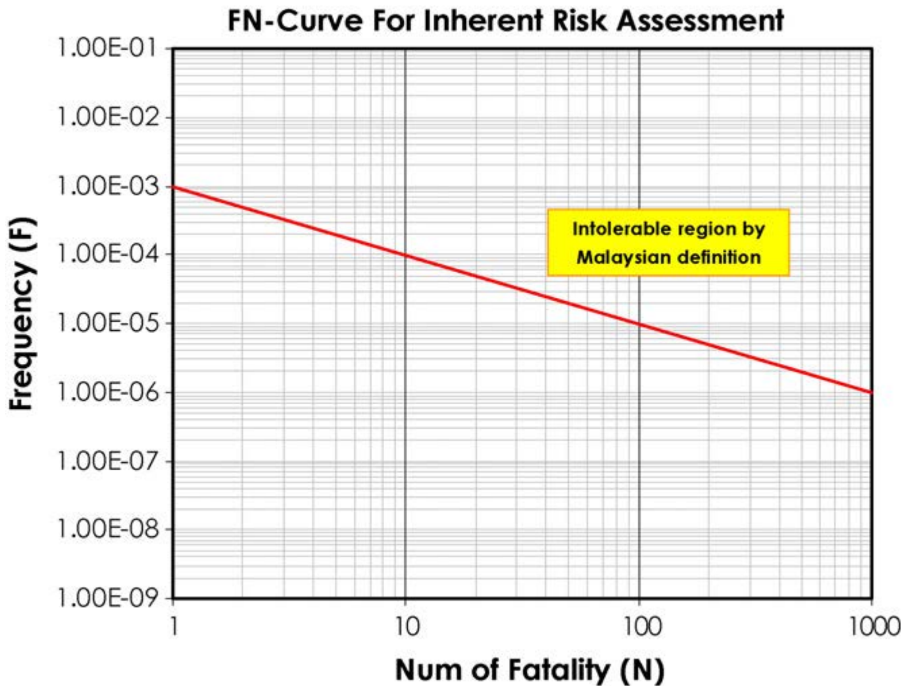

Figure 4 2-region FN curve – IRA. Adapted from Shariff & Leong, 2009

Table 3 describes the comparison between QRA and IRA.

Criteria	QRA	IRA
Implementation stage	after completion of detailed design	preliminary design / simulation stage
Purpose	to demonstrate or prove "safety case" as required by regulatory agencies	to proactively identify risk inherent to the design and guide its reduction by implementing inherent safety principles
Regulatory requirements	required by regulatory agencies	no regulatory requirement
Information required	Process & Instrumentation diagram (P&ID), detailed historical weather data	simulation data and predicted piping and equipment sizing
Scenario	few credible scenario studied in detail	basic scenario, such as equipment leak
Duration of analysis	ranging from 40 to 1500 manhours	relatively quick, carried out parallel to simulation work
Representation of result	3-region FN curve	2-region FN curve

Table 3 Comparison between QRA & IRA. Adapted from Shariff & Leong, 2009

2.6 Limitations of Current Research

Literature review of various ISL techniques which are available shows that there are multiple safety assessment indices developed which accounts for various process factors like toxicity, flammability, reactivity, explosiveness, temperature, pressure to name a few. However, there are hardly any indices available and attention given on equipment safety – equipments like heat exchangers can also be made inherently safer mechanically, by using ISD principles, during the basic engineering phase of design. ISI is the only pioneered safety index which places focus on equipment safety, in initial design phase of a project. ISI defines equipment by using two parameters – chemical interaction with other chemical & MOC, and type of equipment. Chemical interaction is a sub-index for reaction hazards and type of equipment is sub-index for process hazards. In this method of quantifying equipment safety, using the technique of type of equipment – all the heat exchangers or reactors or compressors receive the same level of inherent safety, irrespective of the MOC or operating conditions. Thus, this method is not very helpful in determining safety level of standalone equipment, once the process route has been finalized. Also, the chemical interaction parameter is a subindex of reaction hazards, along with six more parameters quantifying reaction hazards. As a result, the negative effect of one parameter can cancel the positive of another and can lower the value of overall subindex. Thus, it is needed to evaluate equipment safety as a separate subindex in process and taking into consideration the operating conditions of the equipment, MOC and not only relying on type of equipment.

The research area of inherently safer HEN design has mostly been concentrated on improvement of HEN design by analyzing sub-indices which are defined by process chemical properties and operating conditions. The possible interaction of selected chemical or the operating conditions with MOC of the heat exchanger which can result in mechanical failure of the equipment is not considered. There are numerous methods available to perform hazard assessment of STHE in the later phases of project design, but there is very limited research done on application and validation of safety indices on heat exchangers and heat exchanger networks which considers both process and mechanical safety aspects of the equipment and which can be utilized in early phases of design.

This study aims to include learnings from historical incidents/performances into equipment safety at basic engineering/simulation stage of project design. The intent is that once the inherently safer process route has been selected; ask the question ‘is there an alternative to make the equipment inherently safe? Are there inherently safer metallurgy available for the subject chemical?’ – And include the outcome as part of safety assessment carried out during initial stages of design. Information regarding material of construction (MOC) of major equipments is generally available during basic engineering phase and if ISL assessment of suitability of MOC for given chemical and operating conditions is incorporated in initial safety considerations along with analysis of PSI, it would help in avoiding incidents due to inappropriate metallurgy selection, which is a significant cause behind the failure of a number of heat exchanger.

3. PROPOSED METHODOLOGY

This section describes the developed method for safety assessment of heat exchanger, in the design/simulation stage of any project, which is divided in three parts. First accumulate data on past heat exchanger incidents and analyze causal factors of these incidents to determine if sufficient information about the defining variable is available during the preliminary design stage to carry out safety assessment and implementing ISD alternatives to avoid such incidents. The second part is development of a Heat Exchanger Safety Index (HESI) to analyze the risk associated with the heat exchangers and then use them to assess the overall risk of heat exchanger network using Heat Exchanger Network Overall Safety Index (HENOSI) during initial phase of design of a process. The third section of the study utilizes the developed safety indices in Two-tier Inherent Safety Index (2TISI) and provides a Modified Inherent Safety Framework (MISF), utilizing both PSI and ESI.

3.1 Past Heat Exchanger Failure Data

Table 4 summarizes few of many heat exchangers failures that happened in the industry and the root cause of the failure identified during the investigation, as can be seen for quite a few incidents that sufficient information regarding the root cause variable was available during the basic engineering phase of process design and was found during the investigation that some of the incidents could have been avoided if inherently safer design principle was applied early-on in the design phase.

S. No.	Exchanger Service	Failure Causes	Contributing variables	Variable defining stage	Future direction	Reference
1	Preheat feed gas of reactor by outlet gas of the same reactor in naphtha hydrotreating unit	High temperature hydrogen attack (HTHA)	High concentration of reactive component	Preliminary design stage	Use compatible and inherently safe material	(CSB, 2014)
			Inappropriate material	Basic engineering stage		
2	Industrial water at shell side and cooling water at tube side.	Erosion Corrosion	High concentration of reactive component	Preliminary design stage	Select optimized flow velocity and appropriate tube material	(Kuznicka, 2009)
			High flow velocity	Basic engineering stage		
			Inappropriate material of tubes	Basic engineering stage		
3	Cooling water in tubes and steam is on the shell side	Flow-induced Erosion	Low velocity	Preliminary design stage		(Ranjbar, 2010)
			Inappropriate tube material	Basic engineering stage		
4	Flue gas at shell side and Boiler Feed Water (BFW) at tube side.	Creep attack due to corrosion in the whole system	Tubes overheating	Preliminary design stage	Improved design of heat exchangers	(Jahromi, AliPour, & Beirami, 2003)
			Poor water treatment	Operations		
5	Four gas coolers, gas is inside of tube and seawater is on the shell side.	Crevice Corrosion	Inappropriate tube Material	Basic engineering stage	Use compatible and inherently safer material	(Allahkaram, Zakersafaee, & Haghgoo, 2011)
6	Process gas in tube side while cooling water in the shell.	Stress corrosion Cracking	Inappropriate material of tubes	Basic engineering stage	Use of appropriate tubes material	(Esaklul, 1992)
7	Process gas at shell and BFW at tube side	Thermal fatigue	Excessive heating	Preliminary design stage	Timely inspection	(Usman & Khan, 2008)
8	Ammonia in the shell side and process chemical in the tube side.	Over Pressurization	Pressure	Preliminary design stage	Emphasize on workers safety training	(CSB, 2011)
9	Condensate at Tube side and Heavy Gas Oil (HGO) at the shell.	Intergranular stress corrosion cracking	Poor fabrication (welding)	Not applicable	Improved welding process	(Guo, Han, Tang, Zuo, & Lin, 2011)

Table 4 Failure Analysis of STHE through various accidents. Adapted from Pasha, 2017

3.2 Index Calculation

The Heat Exchanger Safety Index (HESI) is developed using concept similar to calculation of Inherent Safety Index (ISI) & Process Stream Index (PSI). The mathematical formulation of HESI will take into consideration various factors as mentioned below:

HESI = f(Pressure, Temperature, flammability, flammable range, metallurgy interaction)

HESI = f(P, T, FL, FR, M)

Pressure, temperature values is with units bar, degC respectively. The conversion of individual parameters to dimensionless index and scoring is based on ISI methodology (Heikkilä, 1999) and is scoring is explained through table 5 to table 8.

Flammability sub index

Flash Point (°C)	Flammability	Score (I_F)
Undefined	Nonflammable	0
Flash point > 55 °C	Combustible	1
Flash point ≤ 55 °C	Flammable	2
Flash point < 21 °C	Easily flammable	3
Flash point < 0 °C	Very flammable	4

Table 5 Flash point conversion

Flammable range for the process fluid can be calculated as follows:

$FR = UFL_{mix} – LFL_{mix}$

$UFL_{mix} = \frac{1}{\sum_{i=1}^{n} \frac{y_i}{UFL_i}}$ $LFL_{mix} = \frac{1}{\sum_{i=1}^{n} \frac{y_i}{LFL_i}}$

where, y_i = mole fraction of an individual component in the mixture.

Flammable range sub index

Flammable range (UFL-LFL) vol%	Score (I_{FR})
Non explosive	0
0 – 20	1
20 – 45	2
45 – 70	3
70 – 100	4

Table 6 Flammable range conversion

Temperature sub index

Process temperature (°C)	Score (I_T)
< 0 °C	1
0 – 70 °C	0
70 – 150 °C	1
150 – 300 °C	2
300 – 600 °C	3
> 600 °C	4

Table 7 Temperature conversion

Pressure sub index

Process pressure (bar)	Score (I_P)
0.5 – 5 bar	0
0 – 0.5 or 5 – 25 bar	1
25 – 50 bar	2
50 – 200 bar	3
200 – 1000 bar	4

Table 8 Pressure conversion

Metallurgy conversion sub index is developed based on guidelines applicable to a chemical process, which takes into consideration learnings from previous incident investigations and root cause of failures identified and is shown in table 9. A set of questions are used to estimate metallurgy interaction as a dimensionless score, where lower score implies safer selection and higher is relatively unsafe. This sub-index gives the opportunity to analyze and incorporate the learnings from previous incidents which are not a part of "mandatory regulations", but good engineering practice.

Metallurgy interaction sub index

Process fluid-metallurgy		MI
previous history of exchangers in similar service checked	Yes	0
	No	1
design improvements and learnings from previous incidents included, if applicable	Yes	0
	No	1
Corrosion/erosion at selected process conditions - velocity/Pressure/Temperature	acceptable	0
	higher	1
is there better/inherently safer metallurgy available for selected conditions - Pressure/Temperature/process fluid	Yes	1
	No	0
If yes, for inherently safer metallurgy availability - Is the metallurgy selected inherently safer, after considering consequence analysis	risk acceptable	0
	risk unacceptable	1
additional layer/cladding required	Yes	1
	No	0
Post-weld heat treatment done	Yes	0
	No	1
process fluid can react with MOC resulting in pressure/temperature build up/gas generation	Yes	1
	No	0
MOC susceptible to HTHA/Sulfidatin corrosion/embrittlement/stress corrosion cracking	at normal operating conditions (include safety margin of 55 deg F and 50 psi)	2
	at higher/lower operating conditions ±100 degF and 50 psi	1
	not susceptible	0
Possibility of contaminants in the process fluid which can lead to metal embrittlement/ pitting corrosion	Yes	1
	No	0

Table 9 Metallurgy interaction conversion

The metallurgy sub-index, I_M is estimated as per the equation below:

$$I_M = \sum MI$$

Combining these dimensionless numbers gives the inherent safety level of the heat exchanger. Higher the value of HESI, indicates lower safety level of the heat exchanger.

$$HESI = I_F + I_{FR} + I_P + I_T + I_M$$

In case of heat exchanger network design, the calculated HESI will be used to estimate HENOSI

$$HENOSI = \sum HESI$$

The calculated HESI, can be used to relatively rank all the heat exchangers in the network and worst heat exchanger can be identified as

$$W_{HESI} = \max(HESI)$$

The worst heat exchanger is then analyzed further for frequency of a credible event and consequence of that event. Loss of containment can lead to fire, explosion or catastrophic rupture of the heat exchanger, severity of these events can be estimated using the systematic approach defined in (Crowl & Louvar, 1996). Estimation of consequence for the event of process stream explosion can be carried out by using iRET tool (Shariff, Rusli, Leong, & V.R. Radhakrishnan, 2006). iRET tool can also be utilized for estimation of explosion consequences in the event of heat exchanger failure, as shown in the study by (Zaini, Pasha, & Kaura, 2016). For estimation of frequency of explosion, event tree analysis (ETA) method can be used. A simple framework for estimating release outcome frequencies is shown in figure 5.

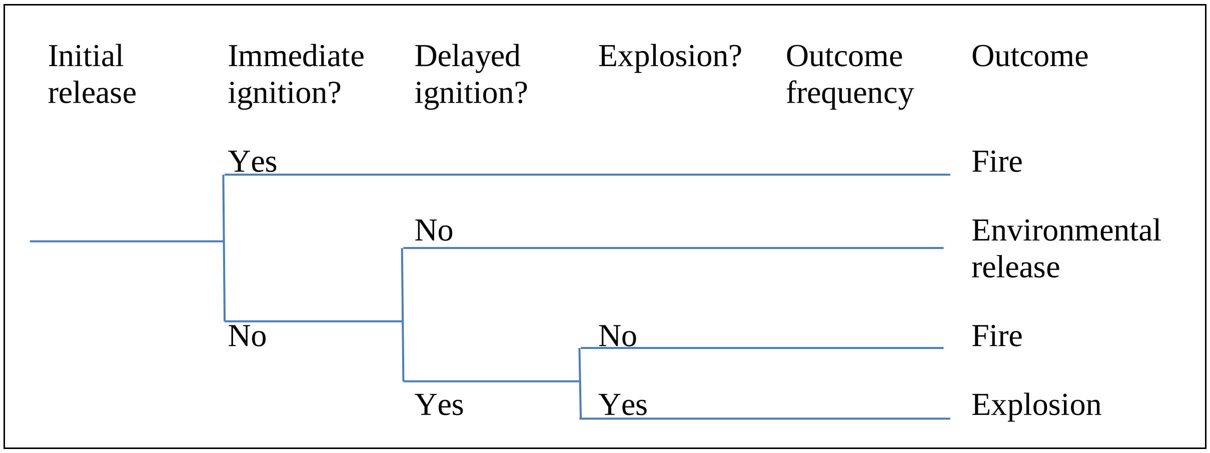

Figure 5 Event tree for calculating release outcome frequencies. Adapted from Moosemiller, 2011

Default frequency of leak from a shell and tube heat exchanger resulting in a rupture can be referred from (Moosemiller, 2011). If the risk estimated falls within the acceptable range previously defined in IRA step, the heat exchanger and HEN design can be recommended for detailed design phase. If the estimated risk for worst case scenario lies outside the tolerable range, further modification should be carried out using inherent safety principles. This can be achieved by changing HEN flow arrangements, varying operating condition or by changing the MOC.

3.3 Modified Inherent Safety Framework

The developed indices are then used as input to MISF in the framework shown in figure 6 – in the place of ESI

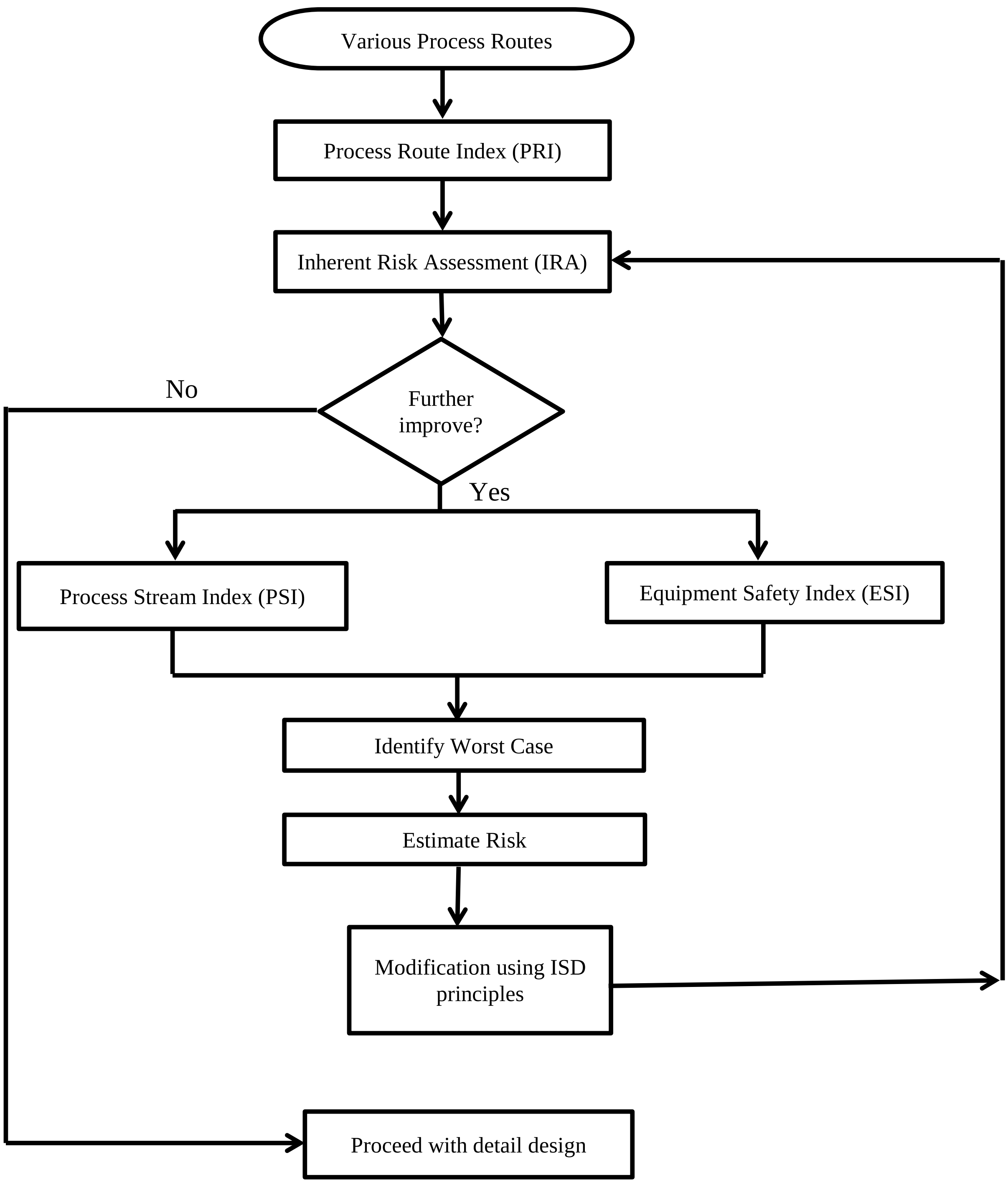

Figure 6 Modified inherent safety framework (MISF)

The inherent safety principles which affect the developed equipment safety sub-index (HESI & HENOSI) are explained in figure 7.

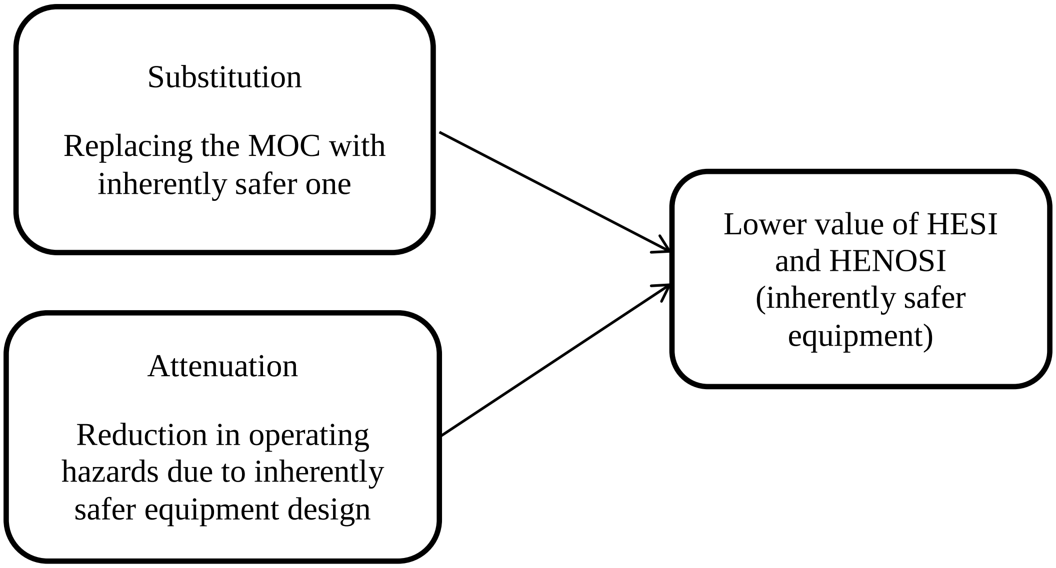

Figure 7 Inherent safety principle and HESI

4. RESULTS AND DISCUSSION

In this chapter, the proposed safety index and framework for determining inherent safety level of STHE and HEN was applied to a case study. The main objective is to show how the methodology could be applied in industry.

4.1 Case Study

For the purpose of this research, catastrophic rupture of heat exchanger at Tesoro, Anacortes refinery in April, 2010 is taken as the case study. This incident is considered the largest fatal incident at a US petroleum refinery, since the BP Texas City incident in March, 2005. (CSB, 2014)

4.2 Catastrophic Rupture of Heat Exchanger at Tesoro, Anacortes refinery

On April 2, 2010, the Tesoro Refining and Marketing Company LLC petroleum refinery in Anacortes, Washington, experienced a catastrophic rupture of a heat exchanger in the catalytic reformer/Naphtha Hydrotreater unit. (CSB, 2014). The heat exchanger was in service for handling highly flammable mixture of hydrogen and naphtha, at temperature higher than 500 degF. The rupture of exchanger resulted in release of the flammable mixture and ignited, causing an explosion and fire that continued burning for three hours. This incident fatally injured seven employees of Tesoro who were working in the nearby area. (CSB, 2014) Figure 8 shows the schematic of exchanger set-up in the unit.

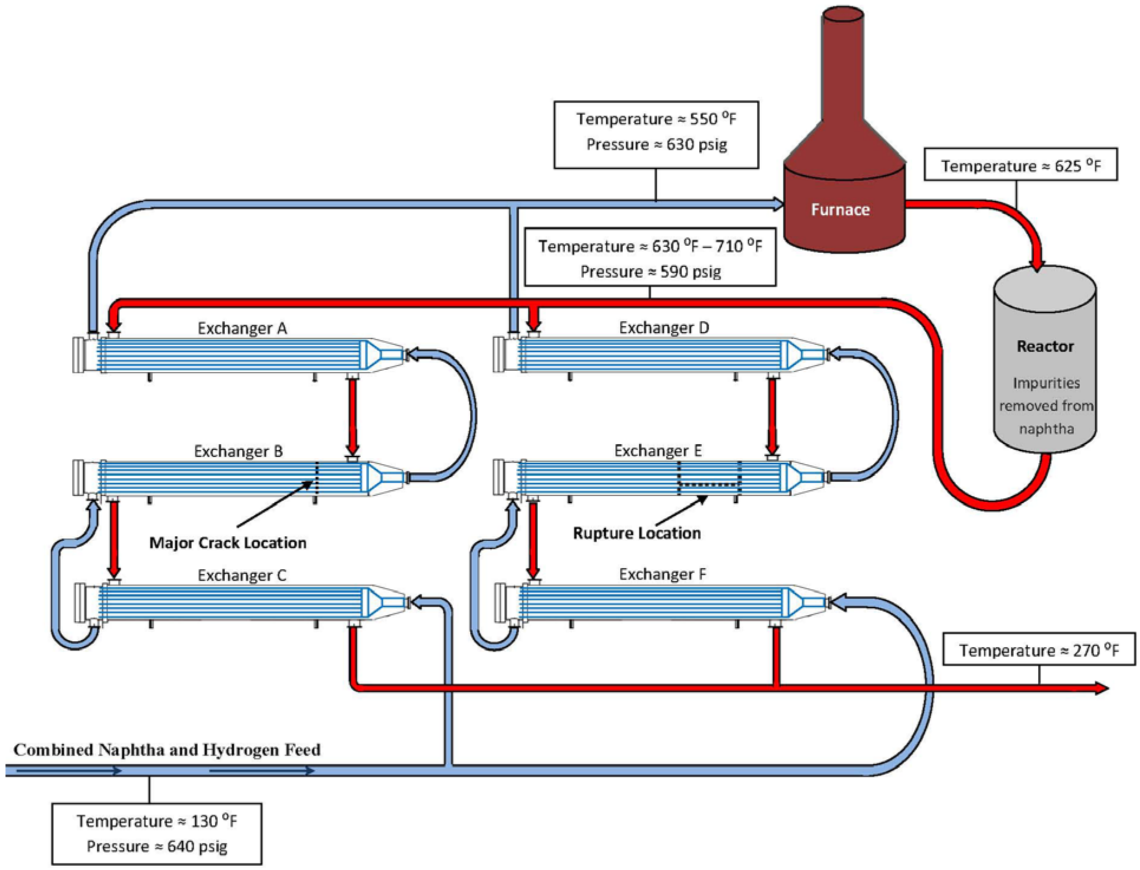

Figure 8 Schematic of the Tesoro Anacortes Refinery NHT heat exchanger bank. Adapted from CSB, 2014

The NHT unit had two parallel banks of heat exchangers (A/B/C & D/E/F), these exchangers were used to pre-heat the reactor feed with reactor effluent. The banks were frequently required to be taken out of service and cleaned due to fouling issues. At the time of the incident, workers were in the process of putting the A/B/C bank of exchanger back in service, after maintenance work was completed, the D/E/F exchanger bank remained in operation during this time. While the start-up operation was being performed, the exchanger E catastrophically ruptured.

4.3 Root Cause for the Exchanger Failure

The primary cause for the exchanger failure was determined to be High Temperature Hydrogen Attack (HTHA). HTHA is a damage mechanism that results in fissures and cracking and it occurs when carbon steel equipment is exposed to hydrogen at high temperature and pressure. (CSB, 2014) There were several other contributing factors like – failure to identify HTHA as a credible even in periodically performed hazard reviews; failure to learn from previous near-misses – the heat exchangers had a history of leaks during start-up, for which the recommendation was to use steam to mitigate leaks. Possibility of HTHA was identified during various PHA that the refinery carried out, however ineffective judgement-based, qualitative safeguards were recommended for equipment protection against HTHA and the adequacy of these safeguards were never evaluated.

4.4 Incident Investigation and HTHA Mechanism

Detailed investigation of the incident was carried out by Chemical Safety Board (CSB). The explosion occurred due to weakening of carbon steel metallurgy of the exchanger due to HTHA. The refinery was purchased by Tesoro in 1998, it was previously owned by Shell Oil. PHA conducted in 1996 by Shell Oil, cited ineffective, qualitative safeguards for protection of equipment against HTHA. The PHA revalidations done in 2001 and 2006 by Tesoro, did not modify the previous recommendations. PHA carried out in 2010, failed to identify HTHA as a credible event for the subject heat exchangers.

HTHA occurs when atomic hydrogen diffuses into the steel walls of process equipment, as shown in figure 9. The hydrogen then reacts with carbon in steel to form methane, this reaction is called decarburization. Methane being larger molecule than atomic hydrogen cannot diffuse out of the steel. Loss of carbides weakens the steel and accumulation of methane exerts pressure in the vessel wall, creating cavities and fissures which then combine to form microcracks and over time go on to form large cracks. HTHA damage is extremely difficult to inspect owing to the microscopic and localized nature of the damage, and thus routine inspection is not sufficient or reliable enough to ensure mechanical integrity.

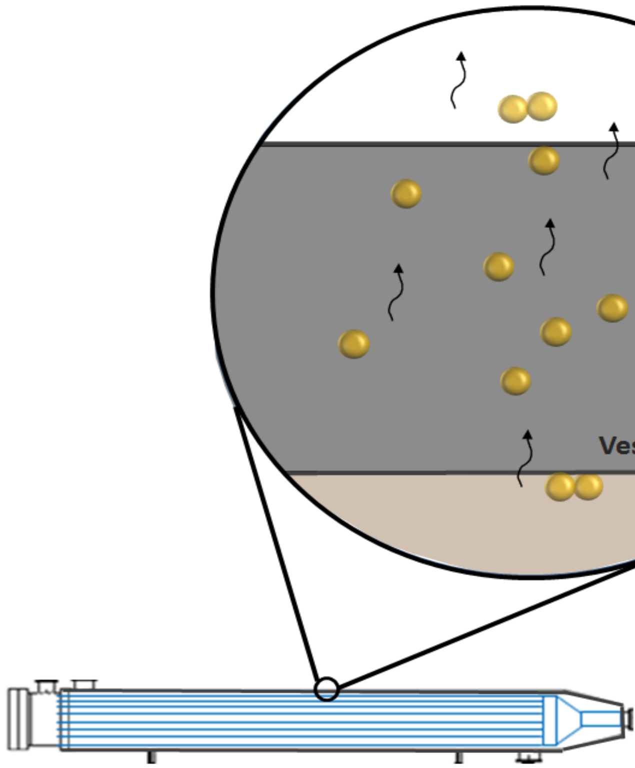

Figure 9 Diffusion of atomic Hydrogen through Carbon Steel. Adapted from CSB, 2014

Industry generally relies on API RP 941 *Steels for Hydrogen Service at Elevated Temperatures and Pressures in Petroleum Refineries and Petrochemical Plants* to predict the occurrence of HTHA. This document uses Nelson curves for HTHA prediction. Nelson curves, developed in 1949 by George Nelson, are empirical and based on data from actual industry experience.

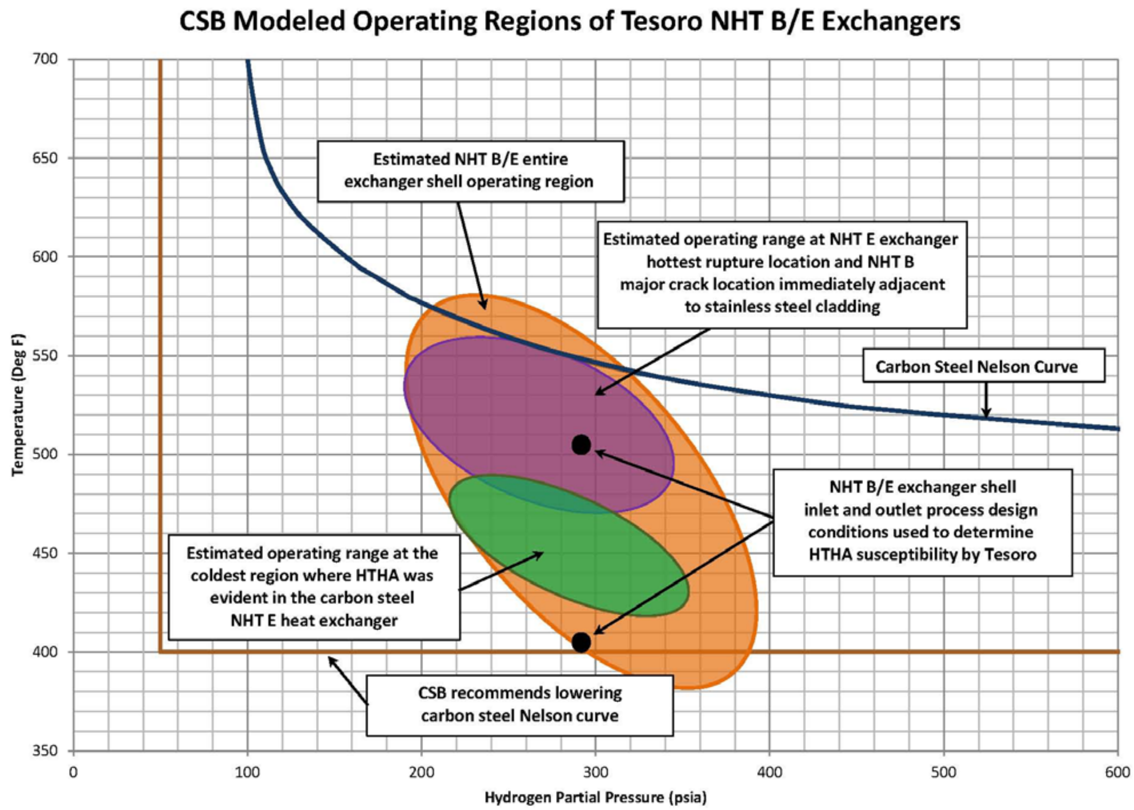

Figure 10 Carbon Steel Nelson Curve and operating conditions for heat exchanger B/E, Tesoro, Anacortes refinery. Adapted from CSB, 2014

Figure 10 shows operating conditions of heat exchanger B/E in comparison with Nelson curve. CSB investigation indicated that HTHA damage was present in exchanger B as well, though for both the exchangers B & E actual operating condition were modeled to be lower than Carbon Steel Nelson curve. Thus, indicating that the industry developed Nelson curve is inaccurate and cannot be relied on to prevent HTHA. CSB investigation report mentions that they have learned of at least eight other recent refinery incidents where HTHA occurred below the carbon steel Nelson curve. (CSB, 2014)

Nelson curves predict HTHA based on three parameters – process temperature, hydrogen partial pressure and MOC. Carbon steel is the lowest curve indicating that it is most susceptible to HTHA. Higher alloyed steel is the inherently safer metallurgy to protect equipment against HTHA.

4.5 Application of the Developed Methodology

The developed methodology is applied to the NHT feed-effluent heat exchangers of Tesoro, Anacortes refinery.

For flammability and explosiveness calculations stream composition for the inlet and outlet of a typical NHT heat exchanger is required, based on the shell side operating pressure of ~590 psig and CSB estimated hydrogen partial pressure of ~290 psig, it can be easily estimated that hydrogen mol% in the stream flowing through heat exchanger B/E was ~50 mol%. For the remaining of the heat exchangers in service this composition can be varied directionally with temperature variation, the assumed stream composition for shell side (hot fluid) is shown in table 10.

	Shell (hot fluid)	
	H2 (mol%)	Naphtha (mol%)
A/C	80	20
B/E	50	50
D/F	45	55

Table 10 Vapor composition for typical NHT feed – effluent streams

The naphtha MSDS from Tesoro specifies flash point as '-21.7 degC' (Tesoro), based on this information it can be inferred that naphtha-hydrogen mixture will be in 'very flammable' range. Naphtha LEL and UEL values of 1.2% and 6.9% respectively were taken from Naphtha MSDS (Tesoro) and flammability limits of H2 (4 to 76%) were taken from SDS by (Airgas). Flammable range for the mixture is estimated in table 11.

	LFL (%)	UFL (%)	Flammable range (%)
A/C	2.7	25.3	22.6
B/E	1.8	12.6	10.8
D/F	1.7	11.7	9.9

Table 11 Flammable range estimation

The operating conditions and MOC for the series of heat exchangers are provided in table 12 and the layout with post weld heat treated sections are shown in figure 11.

	Pressure Bar	Temperature degC	MOC
A/C	44	132	Mn-0.5Mo steel, clad with 304 stainless steel
B/E	44	263	Carbon steel, section clad with 316 stainless steel (fig.10)
D/F	44	354	Carbon steel

Table 12 Operating parameters for NHT heat exchangers

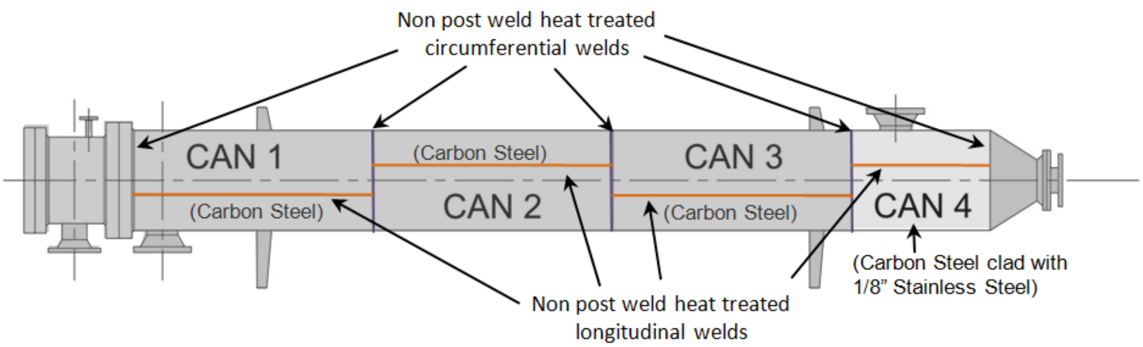

Figure 11 layout of B/E heat exchanger, Tesoro, Anacortes. Adapted from CSB, 2014

Table 13 below shows index corresponding to individual parameters

	I_P	I_T	I_F	I_{EX}	I_M	HESI
A/C	2	1	4	2	4	13
B/E	2	2	4	1	8	17
D/F	2	3	4	1	7	17

Table 13 HESI estimation for the original design

Overall safety index for the network, HENOSI = 47

Worst heat exchanger, W_{HESI} = B/E and D//F

4.6 Analysis of Results

After identifying the worst exchanger of the network, consequence analysis and risk estimation should be carried out. If the risk is found unacceptable, individual components of the HESI is analyzed to check the potential for improving the inherent safety level of the subject exchanger.

For the selected case study, the biggest contributing factor towards higher HESI is seen as I_M followed by I_T. The inherent safety level of the exchanger can be improved in the selected case by reducing the metallurgy sub-index and by optimizing the process conditions. Further analyzing the metallurgy sub-index, it can be seen that not implementing learnings from previous incidents/design modifications are the primary reason behind higher safety level score. Though B/E exchanger operating condition is less severe than D/F, the HESI score for both are same due to design issues which can be modified to reduce the score and improve inherent safety level of the equipment. B/E heat exchangers were provided with additional 316 stainless steel layer for protection against another damage mechanism called sulfidation corrosion (CSB, 2014). The internal layer which was welded to the exchanger resulted in large heat affected zones (HAZ), which was not followed up with post-weld heat treatment (PWHT). The created HAZ and non-PWHT weakened the selected metallurgy further against HTHA. Table 14 shows estimated HESI scores with modified metallurgy selection.

	I_P	I_T	I_F	I_{EX}	I_M	HESI
A/C	2	1	4	2	3	12
B/E	2	2	4	1	0	9
D/F	2	3	4	1	0	10

Table 14 HESI estimation for modified design

By utilizing higher alloy steel metallurgy, which has been proven to be inherently safer against HTHA and sulfidation corrosion, and not relying on cladding followed by manual operation of PWHT, the heat exchanger could be made inherently safer.

5. CONCLUSIONS & FUTURE WORK

5.1 Conclusions

Shell and tube heat exchanger failure is a chronic problem in the process industries. Lack of risk assessment and inherently safer design considerations during the early stages of design is one of the primary causes behind the frequent and repetitive failure of heat exchangers. While there is few pioneered safety indices developed for inherent safety assessment of process streams based on the operating conditions and physical/chemical properties of the process chemical, there is hardly any focus on the safety assessment of equipment which incorporates operating conditions and the interaction between process chemical and selected MOC for the equipment.

The study included the development of a heat exchanger safety index (HESI) and overall safety index for heat exchanger network (HENOSI). HESI includes the mechanical safety factor of the equipment by analyzing how appropriate the material is for the selected process chemical and operating conditions. The developed index can then be utilized in the proposed framework (MISF) which provides systematic approach for evaluating equipment safety in conjunction with safety assessment of process streams, once the process route has been selected.

The developed methodology was applied to the case study of catastrophic rupture of heat exchanger at Tesoro, Anacortes refinery and the obtained results indicate that if inherently safer principles are applied to metallurgy selection process and necessary focus is provided to possible interaction between selected process chemical and material of construction, the safety level of heat exchanger can be improved by using inherent safety principles in the early stages of project. The proposed framework utilizing the developed safety index and inherent risk assessment methods allows for systematic evaluation of safety level of major equipments early on in the project development stage, thereby providing opportunity for implementation of inherent safety principles and could help in reduction of mechanical failures of equipment to a large extent.

5.2 Future Work

- Validate the proposed methodology with another case study implementing the developed indices and framework in the predesign phase of a project e.g. methyl methacrylate process, ammonia synthesis process, methanol production routes
- Automate the whole calculation process by creating spreadsheet in MS Excel and linking with simulation software like HYSYS, to minimize manual inputs/intervention
- Increase the database list of heat exchanger failure incident and their root cause and include more information in the metallurgy interaction sub index, to make it more robust

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