Failure Potential Assessment of Heat Exchanger Network for

Inherently Safer Design

by

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16710

Dissertation submitted in partial fulfilment of
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CERTIFICATION OF APPROVAL

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Approved by,

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(Dr. Dzulkarnain B. Zaini)

UNIVERSITI TEKNOLOGI PETRONAS

BANDAR SERI ISKANDAR, PERAK

May 2015

CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own with the exception of as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

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SUNITARAJ KAUR

ABSTRACT

The inherent safety indices are the widely accepted tool to assess the inherent safety level at the preliminary design stage. Numerous inherent safety indices have been designed to assess the failure potential of shell and tube heat exchanger but limited indices have a capability to measure the failure potential of the heat exchanger network at the preliminary design stage. Moreover, integration between process design stages with risk and consequence estimation is extremely important in order to design inherently safe process plants. However, the lack of formal integration between process design stages with risk and consequence estimation results in unproductive estimation of risk levels and consequence that occurs during a particular process route until the design is completed. Few studies on the integration of risk estimation with process design are available but a viable framework is yet to be reported. Hence, based on the highlighted issue, application of integrated Risk Estimation Tool (iRET) for explosion scenarios is proposed to study the failure potential assessment of heat exchanger network (HEN) at the preliminary design stage.

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CHAPTER 1

INTRODUCTION

1.1 Background of Study

Shell and tube heat exchanger is an equipment which is utilized widely in chemical process and manufacturing industry, power generation, waste heat recoveries, refrigeration, air-conditioning and space applications. It can be used in the industry as a cooler, radiator, evaporator, condenser and boiler. Shell and tube heat exchanger can be utilized under two important conditions which are high temperature range and high flow rates. These conditions differentiate a shell and tube heat exchanger with the rest of the exchangers available. However, this heat exchanger experiences a high failure rate compared to other process equipment as it is used for variety of applications.

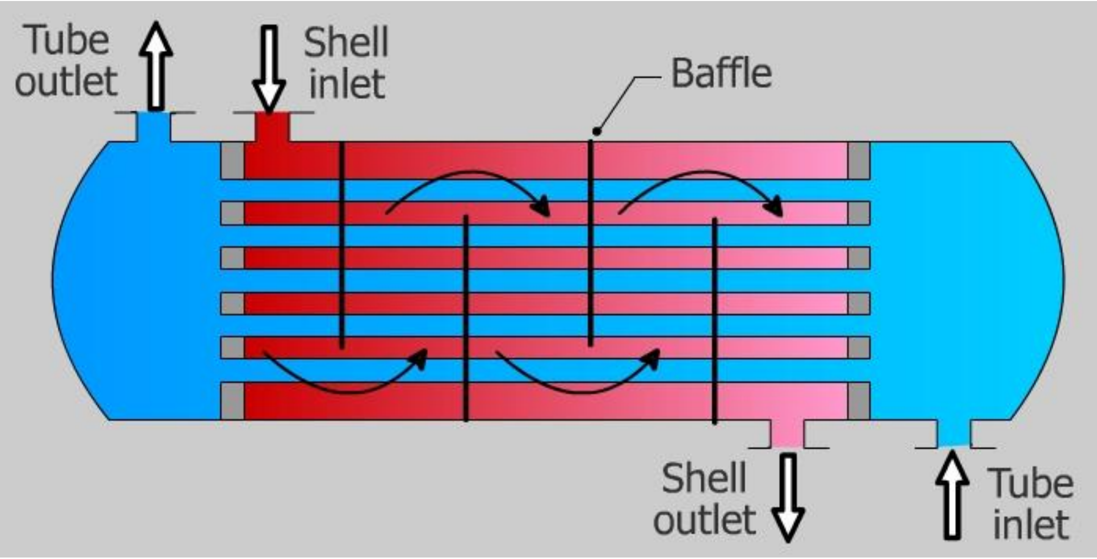

FIGURE 1 Shell and Tube Heat Exchanger

It is observed that shell and tube heat exchanger fails very frequently in chemical process industry whereby this specific equipment is led towards failure without accomplishing its useful life. Various shell and tube heat exchanger failure incidents have been reported as listed in appendix-A and the frequent breakdown of this heat exchanger has caused heavy production loss to the industry. The latest incident took place on the 2nd April, 2010 at Tesoro Anacortes refinery United States. Catastrophic rupture of shell and tube heat exchanger was observed by high temperature hydrogen attack (HTHA).

The failure analysis of various process equipments has been done by few researchers whereby the recent analysis is performed by (Kamarizan, K., & Markku, H., 2013). (Kamarizan, K., & Markku, H., 2013) used failure knowledge database (FKD) system which has been developed in Japan to compare the data obtained to the previously available failure data. The comparison is as follows:

TABLE 1 Type of Equipment Causing the Most Accidents (Kamarizan, K., & Markku, H., 2013)

Equipment Type	Percentage of Accidents					
	Hurme (2013)	Duguid (2001)	Vilchez (1995)	Instone (1989)	Marsch (1987)	Average
Piping System	25	33	16	14	33	24
Reactors	14	9	14	5	10	10
Storage Tanks	14	20	2	14	17	13
Pressure Vessels	10	9	18	3	5	9

Heat Transfer Equipment	8	11	6	19	4	10
Separations Equipment	7	-	-	5	3	5
Other Equipment	22	8	44	40	28	30

From Table 1, it can be seen that the piping system has the largest failure percentage compared to the rest of the process equipment. In order to rank the inherent safety level of process stream at preliminary design stage, (Azmi, M. S., Chan, T. L., & Dzulkarnain, Z., 2012) have developed a process stream index (PSI). Besides that, storage tank is the second highest failure equipment. However, (Kamarizan, K., & Markku, H., 2013) has highlighted that the failure of storage tank is mostly resulted from human and organizational errors. Hence, it is difficult to predict the failure assessment of storage tank by process integration design. (Heikkila, A.M., 1999) considers storage tanks as offsite battery limit (OSBL) equipment which means storage tank are not the part of an integrated design process, so it is difficult to implement process integration methods for the inherently safer design of storage tanks.

Chemical reactors and heat transfer equipment are the next process equipments with the highest failure. (Duguid, I. M., 2001) has ranked heat transfer equipment above chemical reactors due to their scope of work in hydrocarbon industry. On the other hand, Kamarizan, K., & Markku, H. (2013) classified chemical reactor above heat transfer equipment due to the work scope which covered chemical process industry. Both of these equipments are part of process design and integration.

Currently, there are limited inherent safety indices available in order to assess inherent safety level of shell and tube heat exchanger. Two examples of inherent safety indices are safety weighted hazardous index (SWeHI) and integrated inherent safety index (I2SI). These indices are able to measure the inherent safety level of shell and tube heat exchanger equipment at the later design stages whereby inherently safer design (ISD) principles are difficult to be implemented.

Key Performance Indicator (KPI) developed by (Alessandro, T., & Valerio, C., 2007) used consequence based approach to analyse failure assessment of shell and tube heat exchanger. However, this methodology only engulfs the single shell and tube heat exchanger. This index has a limited access to assess the failure potential of heat exchanger network. Secondly, this methodology doesn't provide any solution regarding the prioritization of heat exchanger in the given heat exchanger network.

The latest concern in the industry is that most of the current safety indices have inadequacy to assess the failure potential for the single and multiple shell and tube heat exchangers at the preliminary design stage. A user friendly methodology is required to assess the inherent safety level of shell and tube heat exchanger at the preliminary design stage. Hence, this project proposes the consequence based study of heat exchanger network by application of integrated Risk Estimation Tool (iRET). This tool has been developed by (Azmi, M. S., et al, 2006) and is based on the TNT equivalence and TNO correlation methods. Initially, the implementation of this tool was considered for failure potential assessment of piping system only and for design and layout modification. Currently, the implementation of iRET considers the failure potential assessment of heat exchanger network at the preliminary design stage.

1.2 Problem Statement

Various techniques and methods such as hazard and operability analysis (HAZOP), quantitative risk analysis (QRA), failure mode effect analysis (FMEA) and safety indices are used to quantify the safety level of shell and tube heat exchanger. However, following shortcomings are found in order to assess the inherent safety level of heat exchanger network:

- Meagre accessibility to evaluate the failure potential of heat exchanger network at the preliminary design stage.

- Available safety analysis methodologies have limited potential to layout an inherently safe heat exchanger network.

1.3 Objectives

The objectives of this project are as follows:

- Consequence based strategy by applying iRET tool to analyse the failure potential of shell and tube heat exchanger (STHE).

- Prioritization of heat exchanger according to their failure potential.

- Possible opportunities of Inherently Safer Design (ISD) of Heat Exchanger Network (HEN).

1.4 Scope of Study

The scope of this project is limited on the implementation of iRET tool for explosion scenarios to determine the failure potential assessment of shell and tube heat exchanger (STHE) and achieve a safer design at preliminary design stage. Shell and tube heat exchanger has been selected due to various reasons. First and foremost, it is an integral component of most of the equipment such as distillation column, absorber, and stripper and basket type chemical reactors.

Secondly, it is a part of process integration design and plays a vital role in pinch analysis which is used for energy optimization, eventually, it has substantial failure history. Several case studies and failure analysis reports of heat exchanger are given in appendix-A. Besides that, iRET tool is developed based on TNT equivalence and TNO correlation methods for consequence analysis of worst case explosion scenarios. Probabilities of consequence impacts such as structural damage, glass breakage, fatalities and injury are used to prioritize the heat exchangers in the heat exchanger network and Aspen Hysys V (8.0) is used as simulation platform for this project.

CHAPTER 2

LITERATURE REVIEW

2.1 An Overview of Inherently Safer Design

Inherently safer design (ISD) is an alternative philosophy for addressing safety issues in the design and operation of chemical plants. Eliminating or essentially reducing hazards is the main focus of ISD. According to Centre for Chemical Process Safety (CCPS), the customary approach to deal with chemical process safety has acknowledged the existence and magnitude of hazards in a process. In order to reduce risk, efforts have been put into managing risks associated with the hazards. ISD has the potential to make chemical processing technology simpler and more economical. Besides that, it also provides strong and reliable risk management.

(Hendershot, D. C., 1999) states that the main focus of ISD is on the immediate impacts of single events or chemical accidents on people, the environment, property and business. This generally means the immediate impacts of fires, explosions, and the release of toxic materials in a chemical manufacturing plant. However, these types of events will also have the potential for long-term impacts on people, the environment, and possibly property and business. Reducing the magnitude or potential likelihood of accidents will also have benefits from the perspective of the potential long-term impacts. Even though engineers have recognized the potential benefits of ISD in these other areas, the main intent of ISD is to reduce the frequency and potential impact of chemical plant accidents.

ISD can be considered to be a subset of green chemistry and engineering. Green chemistry and engineering have a much broader scope which includes the following (Hendershot, D. C., 1999):

- Health and environmental impacts of emissions from routine plant operations
- Health and environmental effects of all phases of the production and use life cycle of a material, from the basic raw materials through the final product, including all by-products and wastes
- Sustainable development and impact on non-renewable resources

Safety incidents such as fires, explosions, and toxic releases have both immediate and long-term impacts and are clearly part of green chemistry and engineering. (Hendershot, D. C., 1999) further states that a "green" process is often inherently safer whereby it uses less toxic materials. This type of process may reduce safety consequences, immediate injury from exposure to released material as well as offer reduced long-term health and environmental hazards. However, conflicts may occur as well. A more efficient chemistry may reduce consumption of resources and produce less waste but the chemistry may be more energetic, increasing the safety risk of a reactive chemistry explosion.

Furthermore, inherent safety strives to enhance process safety by introducing fundamentally safer characteristics into process design. Inherently safer plants have less built-in hazard potential than plants with a conventional process concept. (Kletz, T. A., 1991) has formalized six key principles of inherent safety which are intensification, substitution, attenuation, limits effect, simplification and error tolerance. The brief description of each element is given in Table 2.

TABLE 2 General Principles of Inherent Safety (Kletz, T. A., 1991)

PRINCIPLES	EXPLANATION
Intensification	• Reduction of the inventories of hazardous materials
Substitution	• Change of hazardous chemicals substances by less hazardous chemicals
Attenuation	• Reduction of the volumes of hazardous materials required in the process. • Reduction of operation hazards by changing the processing conditions to lower temperatures, pressures or flows.
Limitation of Effect	• The facilities must be designed in order to minimize effects of hazardous chemicals or energies releases
Simplification	• Avoidance of complexities such as multi-product or multi-unit operations, or congested pipe or unit settings.
Error Tolerance	• Making equipment robust, processes that can bear upsets, reactors able to withstand unwanted reactions, etc.

Implementation of ISD at early phase of research and development yields the best opportunities as the changes in process design are cheaper and easier to be done. Inherent safety is difficult to be implemented at a later stage as it is tougher to change the basic technology and probability. (Hurme, M., & Rahman, M., 2005) have highlighted that inherent safety should be implemented at early design stage. However, most of the information is not available at early design stage. Besides that, the importance of inherent safety tool is also emphasized at the preliminary design stage. Figure 2 illustrates the best time to apply ISD principle at research and development phase and preliminary design stage.

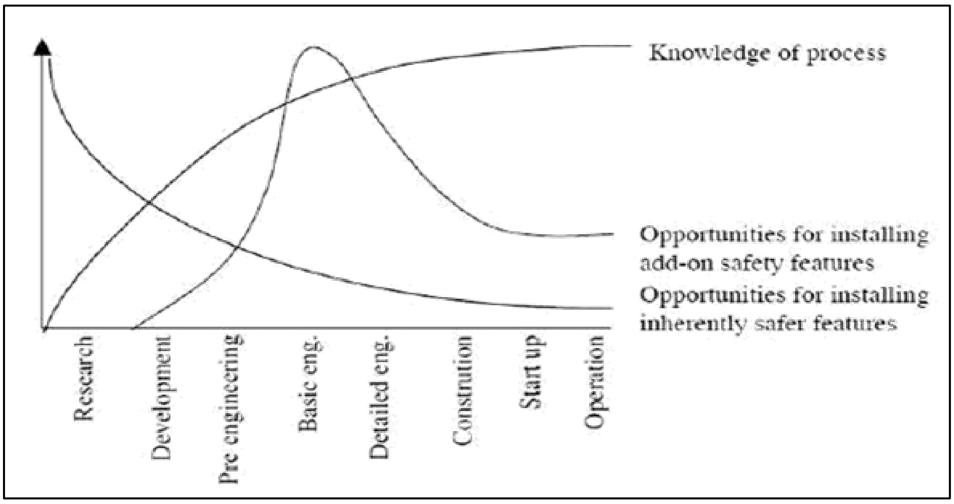

FIGURE 2 The Design Impossibility and Inherently Safer Design (Hurme, M., & Rahman, M., 2005)

Zwetsloot, G. I. J. M., & Askounes Ashford, N., 1999 highlighted that the "primary prevention" tool for chemical accidents is inherent safety. The difference between inherent safety from secondary accident prevention and accident mitigation techniques is that inherent safety can reduce the probability of occurrence of chemical accidents, seriousness of injuries, property as well as environmental damage. The protective measures often require regular preventive maintenance and calibration at later stage of process design which helps to enhance plant operation cost and keeping documentation up-to-date for auditing.

Various books and research publications such as (CCPS, 2000), (Edwards, D. W., Lawrence, D., & Rushton, A. G., 1996), (Hendershot, D. C., 2000) and (Faisal, I. K., & Amyotte, P. R., 2002) concluded that inherent safety is a cost optimal approach when considering the lifetime cost of a process and its operation. Besides that, (Warwick, A.R., 1998) and (Crawley, F., & Tyler, B., 2003) have emphasized that largest pay offs are achieved by implementing inherent safety principles in the early design stage of a plant design.

2.2 Inherent Safety Indices for Consequence Based Analysis

The inherent safety index is a quantitative tool to measure ISD options. It is usually proposed to represent the process hazards that allow the identification of inherently safer design alternatives. Inherent safety indices are given single values that assess inherent safety level of the system. Numerous quantitative tools were developed for inherent safety assessment in process design.

(Gentile, M., et al, 2003) state that inherent safety quantification methodologies can be classified in three groups:

- Collection of various well-known indices used to evaluate various safety aspects but results cannot be aggregated as an overall index.
- Single overall index that evaluate the aspect relevant to inherent safety and results aggregated as an overall index.
- Risk based approach to quantify inherent safety.

The integrated Risk Estimation Tool (iRET) was developed by (Azmi, M. S., et al, 2006) for consequence analysis of worst case explosion scenario. (Azmi, M. S., et al, 2006) used TNT equivalence and TNO correlation methods to develop this tool. This prototype tool was integrated with process design simulator through MS-Excel which reduces time and error. This prototype tool (iRET) can estimate flammable properties, flammable mass, explosion parameters and potential damage. It is applicable at preliminary engineering design stage and requires chemicals and theirs properties, simulated PFD, material balance and operating conditions data for its evaluation.

Besides that, Key Performance Index (KPI) was developed by (Alessandro, T., & Valerio, C., 2007). Its basic objective is to evaluate potential and hazardous index by failure modes and consequence analysis of credible scenarios of single units and the whole process. A specific equipment classification and related failure modes were identified in order to define the potential accidental scenarios associated to each process unit. It can be implemented at preliminary design stage. Process flow diagram (PFD), material balance, operating conditions, equipment technical detail data and substance inventories are required to evaluate the unit potential index (UPI) and unit hazardous index (UHI). The summation of UPI and UHI will provide the potential index (PI) and hazardous index (HI) of the whole process.

Furthermore, Inherent Fire Consequence Estimation Tool (IFCET) was developed by Centre for Chemical Process Safety (CCPS). Its basic objective is to eliminate or minimize the consequence of fire accidents during preliminary design stage. The tool is developed in MS Excel for pool fire model and linked with process design simulator, iCON. It can be implemented at preliminary design stage. However, practical application of inherent safety is still limited due to non-availability of easy to use tool for direct application in a process plant.

CHAPTER 3

METHODOLOGY OR PROJECT WORK

3.1 iRET Algorithm

An iRET algorithm uses an established explosion model in order to estimate consequences resulting from explosions. Determination of mass released in an accident, which is a function of hole diameter, pressure and leak duration is required in an explosion risk and consequence estimation. The iRET algorithm to determine the failure potential of shell and tube heat exchanger is provided in Figure 3:

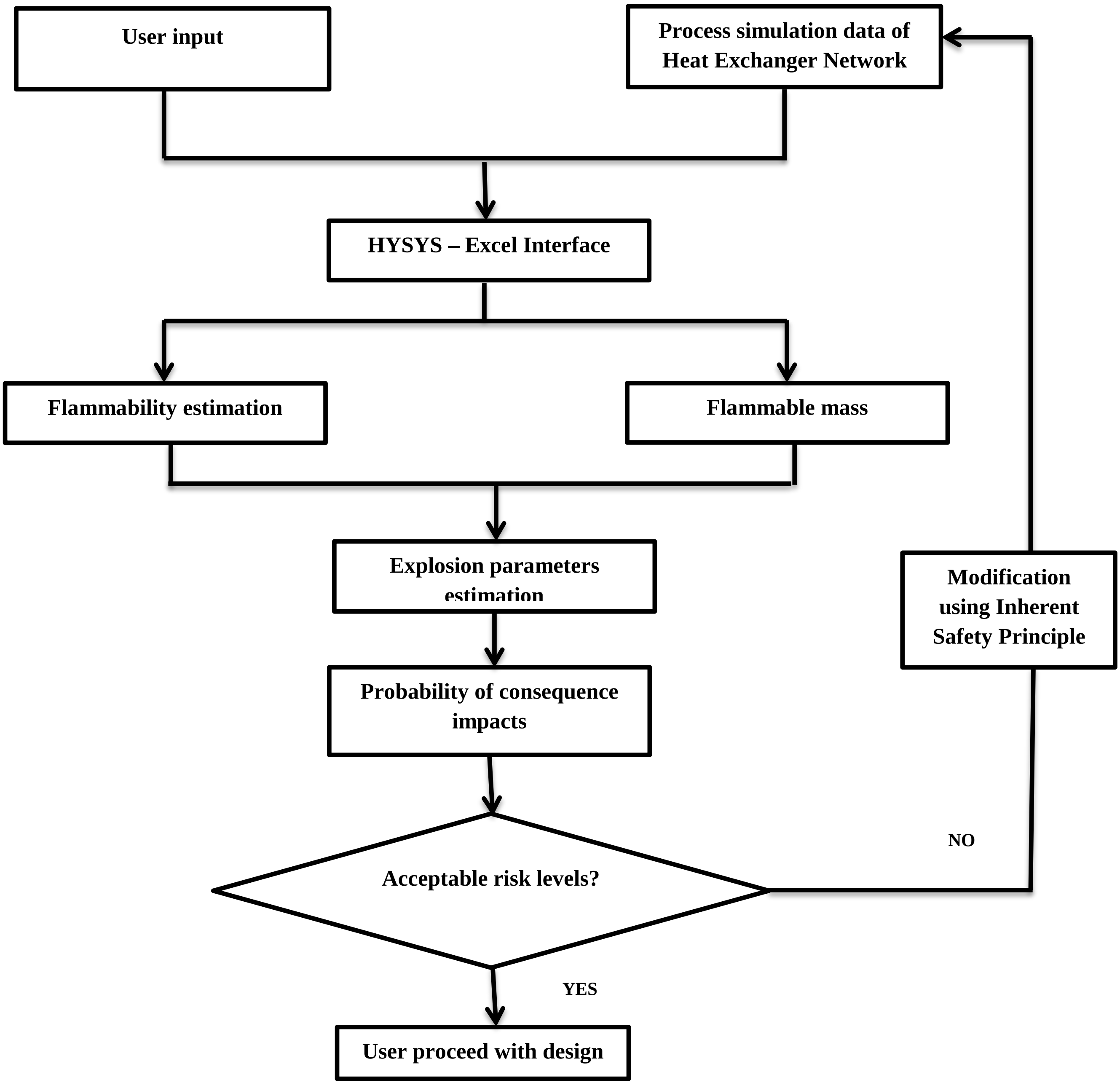

FIGURE 3 iRET Algorithm

(CCPS, 1992) recommends that relative ranking is an appropriate method to compare hazardous attributes, process conditions and operating parameters at the preliminary design stage. However, in this project the values of consequence impacts such as structural damage, glass breakage, fatalities and injury are used to prioritize the heat exchangers in the heat exchanger network. Besides that, in order to define possible options of Inherently Safer Design (ISD) for HEN, various inherent safety principles can be considered. However, it is not recommended to implement all inherently safer design principles to one process system.

Moreover, the selection of possible inherent safety principle that needs to be implemented depends upon the process system design and required outcome. There might be a trade-off among inherently safer design level and magnitude of heat recovery required for a specific heat exchanger. The inherent safety principles such as intensification, substitution, attenuation, simplification and limitation effect can be implemented at preliminary design stage. For this study, attenuation is applied which means reduction of the volumes of hazardous materials required in the process or reduction of operation hazards by changing the processing conditions to lower temperatures, pressures or flows.

3.2 Integrated Risk Estimation Tool

3.2.1 User Input and Process Simulation Data

Users are required to key in a number of inputs in order to estimate consequences resulting from explosions by using integrated Risk Estimation Tool. In order to estimate the explosion risk and consequence, the mass released in an incident is extremely important. Mass released is a function of hole diameter, leak duration and pressure. Data such as hole diameter, atmospheric pressure and maximum distance of interest are required from users. Other process data such as composition, pressure, mass flow rates and heating value that are required by explosion calculation can be transferred from HYSYS to MS Excel (Azmi, M. S., et al, 2006).

TABLE 3 Nomenclature & Subscripts

Nomenclature		Subscripts	
A_h	: Opening area (m^2)	amb	: ambient
C_D	: Discharge coefficient	ex	: explosion
HC	: Heat of combustion (kJ/kg)	mix	: mixture
HC_{TNT}	: Heat of combustion of TNT (4680 kJ/kg)	f	: flammable
LFL	: Lower flammability limit	i	: component i
m	: Total mass (kg)	m	: mechanical
p	: Pressure (kPa)	x	: component x
P_{ovr}	: Peak overpressure (kPa)	y	: component y
Pr	: Probit unit	c	: critical
P	: Probability	o	: system
r	: Actual distance (m)		
UFL	: Upper flammability limit		
y	: Mole fraction of component in mixture		
z	: Scaled distance ($m/kg^{1/3}$)		
η	: Efficiency factor		
γ	: Ratio of specific heat capacities, Cp/Cv		
ρ	: Density (kg/m^3)		

3.2.2 Estimation of Flammability Properties

The basis to determine the Upper Flammability Limit (UFL) and the Lower Flammability Limit (LFL) is the data on stream compositions. Equations (1) and (2) are used in order to estimate the flammability limits for mixture (Wentz, C.A., 1999).

$$\mathrm{LFL_{mix}} = \frac{1}{\sum_{i=1}^{n} \frac{y_i}{\mathrm{LFL}_i}} \tag{1}$$

$$\mathrm{UFL_{mix}} = \frac{1}{\sum_{i=1}^{n} \frac{y_i}{\mathrm{UFL}_i}} \tag{2}$$

3.2.3 Estimation of Flammable Mass

In order to determine whether the flow is choked or non-choked flow, Equation (3) uses operating pressure from the process stream in conjunction with ambient pressure. When the ratio of downstream to upstream pressure is in range of 0.5–0.9 (CEPPO, 1999), it is considered choked flow. Flow rate is insensitive to pressure variation downstream of the choked point under the choked flow conditions. Discharge rates depend on the pressure upstream of the choked point and choking conditions only occur if:

$$\xi_{amb} = p_{amb}/p < \xi_c, \quad \xi_c = \frac{p_c}{p} = \left[\frac{2}{\gamma+1}\right]^{\gamma/(\gamma-1)}. \tag{3}$$

Equation (4) determines the discharge rate for choked flow conditions, (CEPPO, 1999):

$$m = C_D A_h \sqrt{\gamma p \rho \left(\frac{2}{\gamma+1}\right)^{\gamma+1/\gamma-1}} \quad (4)$$

According to (Woodward, J.L., 1999), the discharge coefficient, C_D in Equation (4) is almost close to 0.975 for gases and vapour. C_D increases from 0.61 to 0.975 as vapour fraction increases from 0 to 1 for two-phase flows and C_D for liquids discharging from smooth orifice is around 0.61. Equation (5) is used to calculate the flammable mass when $C_O > C_{UFL}$:

$$\frac{m_f}{m} = \mathrm{erf}\left[\sqrt{\ln\left(\frac{C_o}{C_{LFL}}\right)}\right] - \mathrm{erf}\left[\sqrt{\ln\left(\frac{C_o}{C_{UFL}}\right)}\right] - \frac{2C_{LFL}}{C_o\sqrt{\pi}}\sqrt{\ln\left(\frac{C_o}{C_{LFL}}\right)} + \frac{2C_{UFL}}{C_o\sqrt{\pi}}\sqrt{\ln\left(\frac{C_o}{C_{UFL}}\right)} \quad (5)$$

Explosion is not possible if initial Co is greater than C_{UFL} as the mixture would be out of the flammability limits. However, it is assumed that Co will reduce and fall within flammability limits, thus allowing a possible explosive condition. The value of C_{LFL} should be lower than C_O for the calculations to be valid. In practice, the calculations of release rates cannot be carried out at great precision as neither the prevailing physical conditions nor the failure location are accurately defined. Therefore, it is acceptable that some simplifying assumptions are made allowing the necessary calculations to be performed manually (Woodward, J.L., 1999).

3.2.4 Estimation of Explosion Parameters

Equation (6) is used to calculate the energy released from explosion which is determined by the TNT equivalent method:

$$m_{TNT} = \frac{n_{ex} m HC}{HC_{TNT}} \tag{6}$$

Scale distance, a parameter to determine the explosion overpressure was calculated by using Equation (7):

$$z = \frac{r}{(m_{TNT})^{1/3}} \tag{7}$$

Equation (8) is used to calculate the overpressure:

$$\log p^0 = \sum c_i \ (a + b \log_{10+} z)^i \tag{8}$$

Whereby a, b and c are constants and their values are as follows:

TABLE 4 Values of constants a, b and c

Constants	**Values**
a	-0.2143
b	+1.3503
c_0	+2.7810
c_1	- 1.6960
c_2	- 0.1542
c_3	+0.5141
c_4	+0.0988
c_5	- 0.2939
c_6	- 0.0268
c_7	+ 0.1091
c_8	+ 0.0016
c_9	- 0.0215
c_{10}	+ 0.0001
c_{11}	+ 0.0017

3.2.5 Estimation of Potential Damages

The corresponding damages illustrated in Figure 4 are based on the observation by (Clancey, V., 1972) and are used to predict the potential damages and injuries for TNT equivalent method.

Overpressure (psi)	Damage observed
0.02	Annoying noise (137 dB) if of low frequency (10–15 Hz)
0.03	Occasional breaking of large glass windows already under strain
0.04	Loud noise (143 dB), sonic boom, glass failure
0.10	Breakage of small windows under strain
0.15	Typical pressure for glass breakage
0.30	"Safe distance" probability 0.95 no serious damage beyond this value; projectile limit; some damage to house ceiling; 10% window broken
0.40	Limited minor structural damage
0.50	Large and small windows usually shattered; occasional damage to window frames
0.70	Minor damage to house structures. Large and small windows usually shattered; occasional damage to window frames
1.00	Partial demolition of houses, made uninhabitable. Large and small windows usually shattered; occasional damage to window frames
1–2	Corrugated asbestos shattered; corrugated steel or aluminum panels, fastenings fail, followed by buckling; wood panels (standard housing) fastening fail, panels blown in
1.30	Steel frame of clad building slightly distorted
2.00	Partial collapse of walls and roofs of houses
2–3	Concrete or cinder block walls, not reinforced, shattered. Destruction of cement walls 20 to 30cm width
2.30	Lower limit of serious structural damage
2.50	50% destruction of brickwork of houses. Distortion of steel frame buildings
3.00	Heavy machines (3000 lb) in industrial building suffered little damage; steel frame building distorted and pulled away from foundations
3–4	Frameless, self-framing steel panel building demolished; rupture of oil storage tanks
4.00	Cladding of light industrial building ruptured
5.00	Wooden utility poles snapped; tall hydraulic press (40,000 lb) in building slightly damaged
5–7	Nearly complete destruction of houses
7.00	Loaded train wagon overturned
7–8	Brick panels, 8–12 in thick, not reinforced, fail by shearing of flexure
9.00	Loaded train box cards completely demolished
10.00	Probable total destruction of buildings; heavy machines tools (7000 lb) moved and badly damaged, very heavy machines tools (12,000 lb) survived
300.00	Limit of crater lip

FIGURE 4 Explosion effects and damage (Clancey, V., 1972)

Probit equations are used to calculate the effects of explosions on humans in order to estimate the probability of damage and injuries as a function of overpressure determined in the TNT equivalent method. These equations are derived from experimental studies and actual observations based on (The Netherlands Organization of Applied Scientific Research, 1989).

Probit equation relating structural damage to overpressure:

$$Pr = -23.8 + 2.92 \ln p_{ovr} \tag{9}$$

Probit equation relating glass breakage to overpressure:

$$Pr = -18.1 + 2.79 \ln p_{ovr} \tag{10}$$

Probit equation relating fertility to overpressure:

$$Pr = -77.1 + 6.91 \ln p_{ovr} \tag{11}$$

Probit equation relating injury to overpressure:

$$Pr = -15.6 + 1.93 \ln p_{ovr} \tag{12}$$

The probability of damages and injuries was calculated from probits by using Equation (13):

$$P = 50\left[1 + \frac{P_r - 5}{|P_r - 5|}\operatorname{erf}\left(\frac{|P_r - 5|}{\sqrt{2}}\right)\right] \tag{13}$$

3.3 Gantt Chart & Key Milestones

TABLE 5 Gantt Chart & Key Milestone for FYP 1

No.	**Detail/Week**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**
1.	Selection of Project Topic														
2.	Preliminary Research Work														
3.	Submission of Extended Proposal														
4.	Proposal Defence														
5.	Project Work Continues														
6.	Submission of Interim Draft Report														
7.	Submission of Interim Report														

Key Milestone

Process

TABLE 6 Gantt Chart & Key Milestone for FYP 2

No.	**Detail/Week**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
1.	Project Work Continues															
2.	Submission of Progress Report															
3.	Project Work Continues															
4.	Pre – EDX															
5.	Submission of Draft Report															
6.	Submission of Dissertation (Soft Bound)															
7.	Submission of Technical Paper															
8.	Oral Presentation															
9.	Submission of Project Dissertation (Hard Bound)															

Key Milestone Process

CHAPTER 4

RESULTS AND DISCUSSION

Two case studies were carried out to partly demonstrate the application of the proposed framework by the implementation of iRET tool to determine the failure potential of shell and tube heat exchanger at preliminary design stage. Case studies 1 and 2 demonstrated the capabilities of iRET tool to assess the failure potential of heat exchangers in a steam reforming plant as well as in a methanol plant. Results of case studies 1 and 2 are provided in Section 4.1 and Section 4.2 respectively.

4.1 Case Study 1: Steam Reforming Plant

4.1.1 Steam Reforming Plant before Applying Inherent Safety Principles

The functionality of iRET tool can be demonstrated by conducting a case study on the failure potential of heat exchangers in a steam reforming plant. Figure 5 illustrates the HYSY simulation of a steam reforming plant:

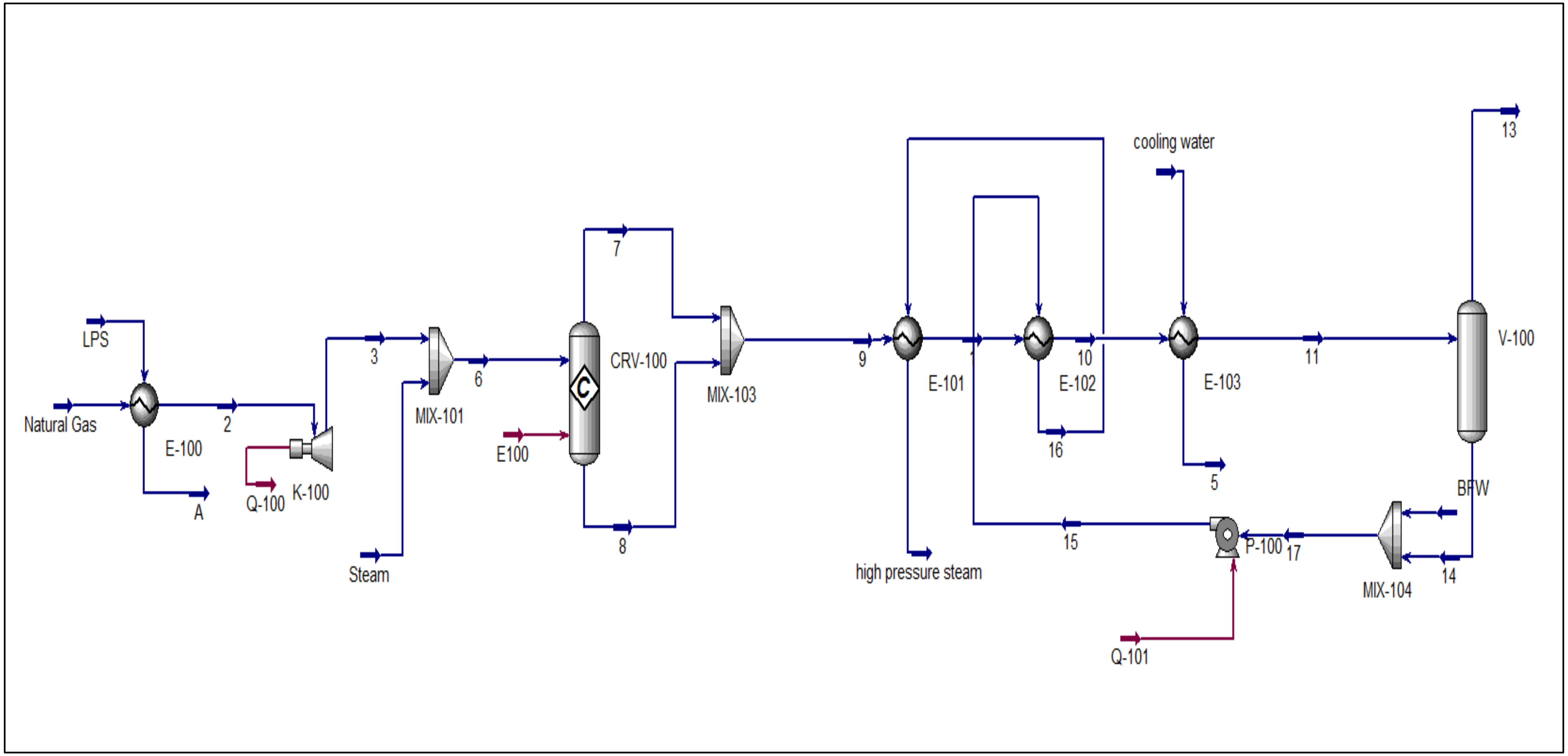

FIGURE 5 Case Study 1 – HYSYS Simulation of Steam Reforming Plant before Applying Inherent Safety Principles

In case study 1, the heat exchanger with the highest failure potential was investigated from a total of four shell and tube heat exchangers. The loss of containment was assumed to be a small leak which originated from 0.0124 meter diameter holes on the heat exchangers. During the initial period, these leaks displayed choked flow conditions due to high pressure in the heat exchangers. Due to the limitation of HYSYS steady-state version, the transition from choked flow to non-choked flow was not considered in this project.

The initial release rate, m was kept constant at 50,000 kg/s. The selected streams resulted in different flammability limits (C_{LFL} and C_{UFL}) as well as flammable mass fraction (m_f/m) due to different compositions. Comparisons were made for the possible destructions observed from a hypothetical explosion from each of the streams. TNO correlation method was used to estimate the consequence from explosion and the results are tabulated in Table 7. The results indicated that any explosion due to leakage from these streams could cause damage up to 100 m.

TABLE 7 Calculations on the Probabilities of Consequence Impacts as a Function of Overpressure in the TNT Equivalent Method

Heat Exchanger	Loss of Containment (LOC)	Diameter of Leak (m)	Flammable Release Rate (kg/s)	Heat of Combustion (kJ/kg)	Actual Distance, r (m)	Overpressure (Pa)	Probability			
							Structural Damage	Glass Breakage	Fatalities	Injury
E-100	Small Leak	0.0124	47606.06	49598.96	100	21590.18	0.63	0.99	0	0.09
E-101	Small Leak	0.0124	41694.54	17027.79	100	18042.72	0.42	0.99	0	0.05
E-102	Small Leak	0.0124	41694.54	17027.79	100	18042.72	0.42	0.99	0	0.05
E-103	Small Leak	0.0124	41694.54	17027.79	100	18042.72	0.42	0.99	0	0.05

The iRET tool prediction of potential structural damage, glass breakage, fatalities and injuries to humans are shown in Table 7. These results were obtained by using the probit functions, based the overpressure calculated by the TNT equivalent method. Table 7 shows that the probability of fatalities is zero at a distance of 100m from the centre of explosion for all streams. However, the probabilities of structural damage, glass breakage and injury are high and need to be reduced. The probabilities of consequence impacts are affected by the heat of combustion whereby the higher the heat of combustion, the higher the probabilities of consequence impacts.

Heat exchanger E-100 has the highest heat of combustion with a value of 49,598.96 kJ/kg compared to other heat exchangers as shown in Table 7. Moreover, this heat exchanger has the highest probability of injury as well at 0.09 due to its high heat of combustion. Hence, the first heat exchanger could be considered as the most critical heat exchanger in this analysis. By adopting inherent safety principles, the process conditions can be modified to achieve a safer design at preliminary design stage and lower the risk and consequence levels of the first heat exchanger which has the highest failure potential.

4.1.2 Steam Reforming Plant after Applying Inherent Safety Principles

The modified HYSYS simulation and results after applying inherent safety principles are shown in Figure 6 and Table 8 respectively:

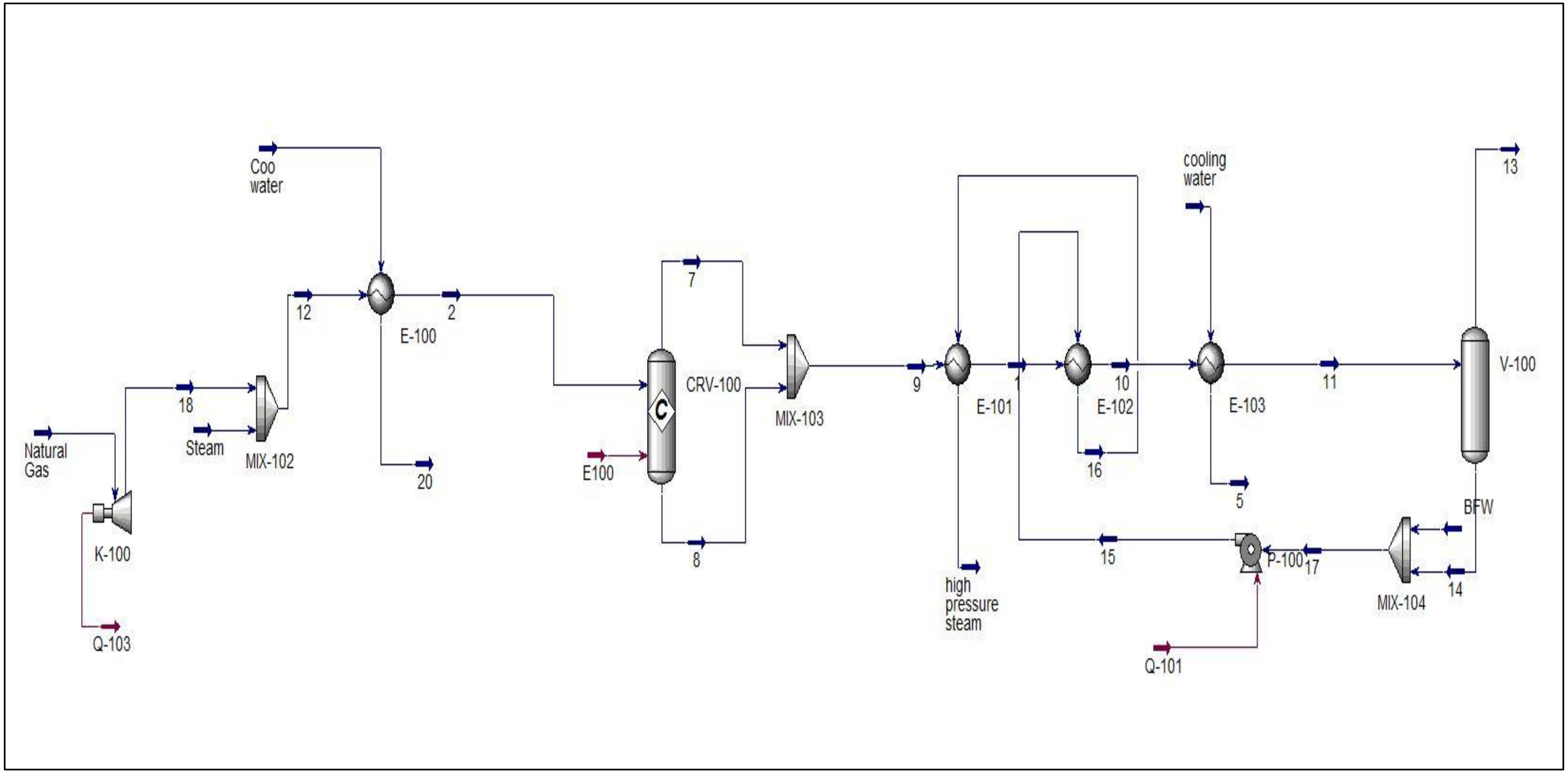

FIGURE 6 Case Study 1 – Modified HYSYS Simulation of Steam Reforming Plant after Applying Inherent Safety Principles

TABLE 8 Modified Calculations on the Probabilities of Consequence Impacts as a Function of Overpressure in the TNT Equivalent Method after Applying Inherent Safety Principles

Heat Exchanger	Loss of Containment (LOC)	Diameter of Leak (m)	Flammable Release Rate (kg/s)	Heat of Combustion (kJ/kg)	Actual Distance, r (m)	Overpressure (Pa)	Probability			
							Structural Damage	Glass Breakage	Fatalities	Injury
E-100	Small Leak	0.0124	47606.06	14133.46	100	17895.56	0.41	0.99	0	0.04
E-101	Small Leak	0.0124	41694.54	17027.79	100	18042.72	0.42	0.99	0	0.05
E-102	Small Leak	0.0124	41694.54	17027.79	100	18042.72	0.42	0.99	0	0.05
E-103	Small Leak	0.0124	41694.54	17027.79	100	18042.72	0.42	0.99	0	0.05

The modification of HYSYS simulation of the steam reforming plant is done by adopting inherent safety principle known as attenuation whereby the mixer (MIX-102) is relocated to the front of the first heat exchanger as illustrated in Figure 6. This modification was necessary in order to reduce the heat of combustion and subsequently reduce the probability of injury. As shown in Table 8, the heat of combustion has reduced by 71% from 49598.46 kJ/kg to 14376.89 kJ/kg and this reduction has managed to reduce the probability of injury from 0.09 to 0.04. Therefore, this case study has successfully demonstrated the potential application of iRET tool in process design to estimate the failure potential of heat exchanger network at preliminary design stages.

4.2 Case Study 2: Methanol Plant

4.2.1 Methanol Plant before Applying Inherent Safety Principles

The functionality of iRET tool was demonstrated by conducting another case study on the failure potential of heat exchangers in a methanol plant. HYSY simulation of a methanol plant is illustrated in Figure 7. In case study 2, the heat exchanger with the highest failure potential was investigated from a total of six shell and tube heat exchangers. The loss of containment in this case study was assumed to be a small leak which originated from 0.0124 meter diameter holes on the heat exchangers. During the initial period, these leaks displayed choked flow conditions due to high pressure in the heat exchangers. Due to the limitation of HYSYS steady-state version, the transition from choked flow to non-choked flow was not considered in this project.

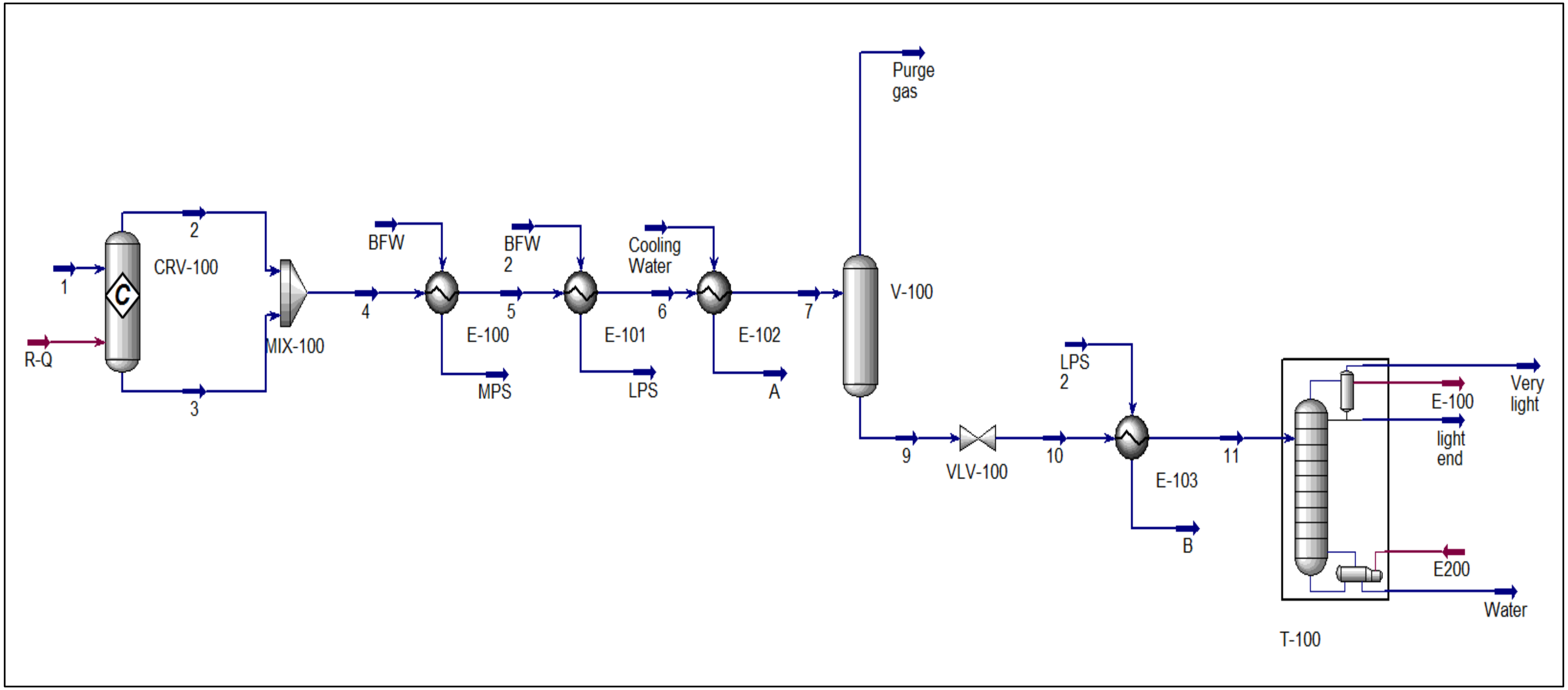

FIGURE 7 Case Study 2 – HYSYS Simulation of Methanol Plant before Applying Inherent Safety Principles

The initial release rate, m was kept constant at 50,000 kg/s. The selected streams resulted in different flammability limits (C_{LFL} and C_{UFL}) as well as flammable mass fraction (m_f/m) due to different compositions. Comparisons were made for the possible destructions observed from a hypothetical explosion from each of the streams. TNO correlation method was used to estimate the consequence from explosion and the results are tabulated in Table 9. The results indicated that any explosion due to leakage from these streams could cause damage up to 100 m.

TABLE 9 Calculations on the Probabilities of Consequence Impacts as a Function of Overpressure in the TNT Equivalent Method

Heat Exchanger	Loss of Containment (LOC)	Diameter of Leak (m)	Flammable Release Rate (kg/s)	Heat of Combustion (kJ/kg)	Actual Distance, r (m)	Overpressure (Pa)	Probability			
							Structural Damage	Glass Breakage	Fatalities	Injury
E-100	Small Leak	0.0124	29114.82	45272.63	100	19810.11	0.54	0.99	0	0.07
E-101	Small Leak	0.0124	29114.82	45272.63	100	19810.11	0.54	0.99	0	0.07
E-102	Small Leak	0.0124	29114.82	45272.63	100	19810.11	0.54	0.99	0	0.07
E-103	Small Leak	0.0124	25848.81	16424.48	100	16674.69	0.34	0.99	0	0.03
E-100	Small Leak	0.0124	25848.81	16424.48	100	16598.07	0.33	0.99	0	0.03
E-200	Small Leak	0.0124	25848.81	16424.48	100	16522.49	0.33	0.99	0	0.03

The iRET tool prediction of potential structural damage, glass breakage, fatalities and injuries to humans are shown in Table 9. These results were obtained by using the probit functions, based on the overpressure calculated by the TNT equivalent method. Table 9 shows that the probability of fatalities is zero at a distance of 100m from the centre of explosion for all streams. However, the probabilities of structural damage, glass breakage and injury are high and need to be reduced. The probabilities of consequence impacts are affected by the heat of combustion whereby the higher the heat of combustion, the higher the probabilities of consequence impacts.

Heat exchangers E-100, E-101 and E-102 have the highest heat of combustion with a value of 45,272.63 kJ/kg compared to other heat exchangers as shown in Table 9. Moreover, these heat exchangers have the highest probability of injury as well at 0.07 due to its high heat of combustion. As there are three heat exchangers with the same heat of combustion, hence, the heat exchanger with the highest log mean temperature difference was taken as the most critical heat exchanger. The log mean temperature difference was 187.9 ^{o}C, 146.6 ^{o}C and 40.88 ^{o}C for heat exchangers E-100, E-101 and E-102 respectively.

Therefore, the first heat exchanger or E-100 could be considered as the most critical heat exchanger in this analysis. By adopting inherent safety principles, the process conditions can be modified to achieve a safer design at preliminary design stage and lower the risk and consequence levels of the first heat exchanger which has the highest failure potential.

4.2.2 Methanol Plant after Applying Inherent Safety Principles

The modified HYSYS simulation and results after applying inherent safety principles are shown in Figure 8 and Table 10 respectively:

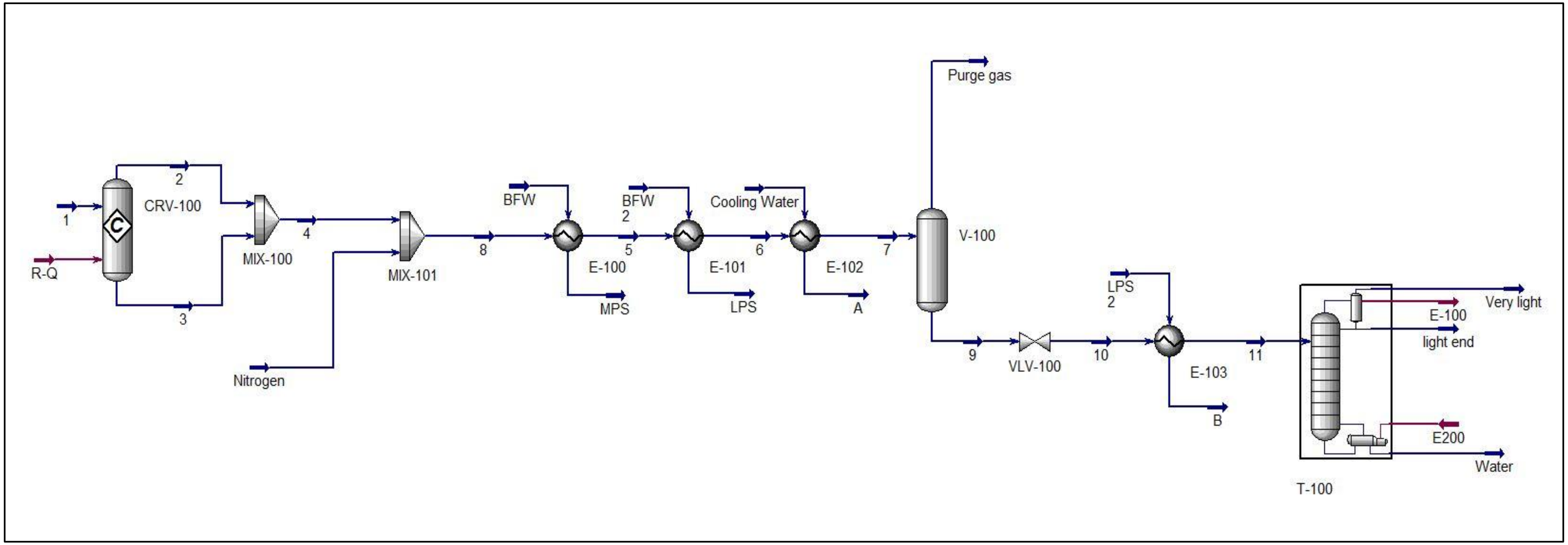

FIGURE 8 Case Study 2 – Modified HYSYS Simulation of Methanol Plant after Applying Inherent Safety Principles

TABLE 10 Modified Calculations on the Probabilities of Consequence Impacts as a Function of Overpressure in the TNT Equivalent Method after Applying Inherent Safety Principles

Heat Exchanger	Loss of Containment (LOC)	Diameter of Leak (m)	Flammable Release Rate (kg/s)	Heat of Combustion (kJ/kg)	Actual Distance, r (m)	Overpressure (Pa)	Probability			
							Structural Damage	Glass Breakage	Fatalities	Injury
E-100	Small Leak	0.0124	40303.04	9585.69	100	16433.85	0.32	0.99	0	0.03
E-101	Small Leak	0.0124	40303.04	9585.69	100	16433.85	0.32	0.99	0	0.03
E-102	Small Leak	0.0124	40303.04	9585.69	100	16433.85	0.32	0.99	0	0.03
E-103	Small Leak	0.0124	16248.45	16248.48	100	16668.09	0.34	0.99	0	0.03
E-100	Small Leak	0.0124	16248.45	16248.48	100	16591.49	0.33	0.99	0	0.03
E-200	Small Leak	0.0124	16248.45	16248.48	100	16515.95	0.33	0.99	0	0.03

The modification of HYSYS simulation of the methanol plant is done by adopting inherent safety principle known as attenuation whereby another mixer (MIX-101) is added with a feed of nitrogen to the front of the first heat exchanger as illustrated in Figure 8. This modification was necessary in order to reduce the heat of combustion and subsequently reduce the probability of injury. As shown in Table 10, the heat of combustion has reduced by 78.8% from 45272.63 kJ/kg to 9585.69 kJ/kg and this reduction has managed to reduce the probability of injury from 0.07 to 0.03.

Therefore, this case study has successfully demonstrated the potential application of iRET tool in process design to estimate the failure potential of heat exchanger network at preliminary design stages. Hence, design engineers can compare the potential damages to acceptable criteria by the implementation of integrated Risk Estimation Tool. If estimated risk does not meet the requirements, process conditions can be modified instantaneously by adopting inherent safety principles to achieve acceptable risk levels as well as a safer design.

CHAPTER 5

CONCLUSION AND RECOMMENDATION

A prototype iRET tool based on the framework in Figures 3 and 4 was used to study the failure potential assessment of heat exchanger network (HEN) at the preliminary design stage. The risk and consequence estimation for worst explosion scenario can easily be done in the initial process design stage with iRET tool. Initially, iRET was limited for piping systems only. Through this study, it has enhanced the scope of iRET for Heat Exchanger Network. Heat exchanger possesses high heat of combustion or heating value of process fluid which results in high consequence impacts. Hence, further development of iRET tool was done and it provided the opportunity of inherently safer design of heat exchanger network by the application of inherent safety principles and produced a comprehensive inherent safety tool for process plant design. Finally, one of the recommendations is iRET tool has a potential to commercialize due to its easy understandable approach and widespread application in process industry. Besides that, the same methodology can be used to develop index for other process equipment.

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APPENDIX

Appendix-A

HEAT EXCHANGER FAILURES CASE STUDIES/FAILURE ANALYSIS REPORTS.

Exchanger Service	**Basic cause of failure**	**Incident Exposure**	**Incident Report/Publication Name and year**
To preheat Feed gas to Reactor by using outlet gas of same reactor.	High Temperature Hydrogen Attack.	Heavy fire included seven Fatalities.	Catastrophic rupture of heat exchanger Investigation report of Tesoro Anacortes refinery issued by U.S. Chemical safety and hazard investigation board May 2014.
Industrial water at shell side and coolant at tube side.	Impingement attack lead towards Erosion corrosion	9 tubes were leaked out of 118 and unit was shut down.	Erosion-corrosion of heat exchanger tubes "B.Kuznicka" Engineering failure analysis 16 (2009) 2382-2387.
Cooling water on tube side and steam is on shell side.	Low velocity led towards fouling and flow induced erosion. Inappropriate tubes material.	Not Mentioned	Effect of flow induced corrosion and erosion on failure of tubular heat exchanger "Khalil Ranjbar" Material and design 31 (2010) 613-619.
Flue gas at shell side and Boiler Feed Water (BFW) at tube side.	Poor water treatment leads towards creep attack due to overheating of tubes.	Not Mentioned	Failure analysis of ammonia plant heat exchanger 101-C "Jahromi, AliPour and Beirami" Engineering failure analysis 10 (2003) 405-421.
Flue gases outside and thermal fluid inside of tubes of vertical heat exchanger	Thermal cycling leads towards thermal fatigue (corrosion fatigue).	Not Mentioned	Failure analysis of a heat-exchanger serpentine " Azevedo and Alves" Engineering Failure Analysis 12 (2005) 193–200
Four gas coolers, gas is inside of tube and seawater is on shell side.	Crevice corrosion	Unit was shut down after 01 year.	Failure analysis of heat exchanger tubes of four gas coolers " Allahkaram, Zakersafaee and Haghgoo" Engineering Failure Analysis 18 (2011) 1108–1114
Cooling water flows inside of tube and hot oil flows on shell side.	Galvanic corrosion and incompatible material of gaskets.	Not Mentioned	Failure analysis of a shell and tube oil cooler " Mousavian , Hajjari et al " Engineering Failure Analysis 18 (2011) 202–211
Process gas and boiler feed water in boiler.	Pitting attack due to excess H2S and SO2 in system	Not Mentioned	Failure of a HE in Sulfur Recovery Unit of a petroleum refinery "V.F.C. Linsa,, E.M. Guimara˜" Journal of Loss Prevention in the Process Industries 20 (2007) 91–97
Process gas at shell and Boiler feed water at tube side	Cycling heating and cooling caused thermal fatigue	Not Mentioned	Failure analysis of heat exchanger tubes " Usman,Nasir A.Khan" Engineering Failure Analysis 15 (2008) 118–128
Ammonia heat exchanger	Over pressurization	One person killed and six injured.	Heat exchanger rupture and ammonia release in Houston, Texasissued by U.S. Chemical safety and hazard investigation board June 2008.