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EXPLOSION POTENTIAL ASSESSMENT OF HEAT EXCHANGER NETWORK AT THE PRELIMINARY DESIGN STAGE

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Abstract

The failure of Shell and Tube Heat Exchangers (STHE) is being extensively observed in the chemical process industries. This failure can cause enormous production loss and have a potential of dangerous consequences such as an explosion, fire and toxic release scenarios. There is an urgent need for assessing the explosion potential of shell and tube heat exchanger at the preliminary design stage. In current work, inherent safety index based approach is used to resolve the highlighted issue. Inherent Safety Index for Shell and Tube Heat Exchanger (ISISTHE) is a newly developed index for assessing the inherent safety level of a STHE at the preliminary design stage. This index is composed of preliminary design variables and integrated with the process design simulator (Aspen HYSYS). Process information can easily be transferred from process design simulator to MS Excel spreadsheet owing to this integration. This index could potentially facilitate the design engineer to analyse the worst heat exchanger in the heat exchanger network. Typical heat exchanger network of the steam reforming process is presented as a case study and the worst heat exchanger of this network has been identified. It is inferred from this analysis that shell and tube heat exchangers possess high operating pressure, corrected mean temperature difference (CMTD) and flammability and reactive potential needs to be critically analysed at the preliminary design stage.

Keywords: Explosion potential, Inherent safety indices, Inherent safety level and Preliminary design stage.

1. Introduction

Inherent safety provides an efficient platform to substantially reduce hazards. Traditional approaches to safety management cope process hazards by incorporating add-on safety measures such as alarms, interlocks and relief system. These additional layers could potentially create complications and complexities in the process design. Inherently safer design (ISD) reduces hazard by altering the basic technology in process design. ISD delivers a simple and economical solution for chemical processes [1]. Moreover, ISD reduces the frequency of credible events such as explosion, fire and toxic release [1]. ISD is considered as a primary prevention tool to avoid accident and its risk mitigation strategy differs from the secondary safety measures. Secondary methods reduce the probability rather than the possibilities of chemical accidents [2]. It provides a cost optimal solution throughout the process plant lifecycle [3].

STHE are being extensively deployed in the chemical process industries. These heat exchangers serve as a heater, cooler, partial or total condenser, evaporator, decomposer and boiler. Tubular exchanger manufacturing association (TEMA) is responsible for issuing and updating the design guidelines of STHE on a regular frequency [4]. The latest version of these guidelines updated by TEMA in 2007 [5]. These standard guidelines provide benchmark criteria for the designing, sizing and rating of STHE.

Leaks are commonly observed in STHE. Failure of STHE could potentially make a cause of major production loss and have a potential of credible events such as explosion and fire. Occasionally, failure of these heat exchangers have also been observed just after a few months of service [6, 7]. On 2nd April 2010 an enormous explosion and fire was observed by the catastrophic rupture of STHE in Tesoro Anacortes Refinery Washington, United States [8]. Seven fatalities were reported in this accident. This tragedy proclaimed as the largest fatal incident in the US petroleum refinery since after the BP Texas City accident [8]. An inadequately designed process system could potentially have a high failure potential. Design error is one of the major cause of accidents in chemical process industries [9, 10]. Moreover, the failure of heat transfer equipment is significantly observed due to inadequate design. Initial leak frequency of STHE is significantly higher than the other process equipment for various loss of containments events such as a small leak, continuous release and catastrophic rupture [11]. Therefore, it is essential to carefully execute the process design and safety analysis of these heat exchangers. Meagre safety analysis could possibly become a failure cause of the heat transfer equipment [10].

The failure of STHE is being widely experienced in the chemical process industries. Inherently safer design provides a gateway to reduce the explosion potential of STHE. Possibilities to implement ISD strategies are reduced as the design proceeds from early design stage to the operation phase [12]. Current work provides a coherent and user-friendly platform for assessing the explosion potential of STHE at the preliminary design stage.

2. Inherent Safety Indices

Inherent safety indices are the quantitative tool for assessing the inherent safety. Safety indices facilitate the design engineer to select an inherently safe chemical process route selection, and the inherent safety level assessment of various process equipment. These indices work swiftly than the conventional safety methods such as hazard and operability analysis, event tree analysis, fault tree analysis and failure mode effect analysis. Safety indices assist in decision-making at the various design cycles and do not require high proficiency and expertise [13].

There are two ways of analysing the inherent safety level of a process system via indices. In the first method, various attributes of process system such as flammability, toxicity, explosiveness, and process conditions are estimated and the outcome expressed in the form of a single value. This value is expressed the inherent safety level. Prototype inherent safety index (PIIS) [14], inherent safety index (ISI) [15], í-safe [16], integrated inherent safety index (I2SI) [17], process route index (PRI) [18], inherent occupational health index (IOHI) [19], health quotient index (HQI) [20], inherent safety key performance indicator (IS-KPI) [21], risk-based inherent safety index (RISI) [22] and numerical descriptive inherent safety technique (NuDIST) [23] were the typical illustrations of such sort of indices.

The second approach is based on the risk estimation. The severity level and frequency of credible events are estimated to evaluate the risk. Inherent risk assessment (IRA) similar to quantitative risk assessment (QRA) was introduced for assessing the inherent risk of an explosion [24]. In this method, the Malaysian 2-region FN curve was used to examine the risk level. Similarly, this concept extended for estimating the risk of toxic release event by using the toxic release inherent risk analysis (TRIRA) [25]. Two region risk matrix was implemented to analyse this risk level. Recently, risk based inherent safety index (RISI) developed for the optimum design selection [22]. This index is an extension of integrated inherent safety index (I2SI) [17].

The safety weighted hazard index (SWeHI) [26] and integrated inherent safety index (I2SI) [17] can be used for assessing the inherent safety of the STHE. However, the detailed process information such as material balance, operating conditions, process flow diagram (PFD), piping and instrumentation diagram (P&ID) and plant layout design were required for the estimation of these indices [11]. Therefore, it is difficult to estimate the inherent safety level of the STHE by using the existing inherent safety indices at the preliminary design stage. Moreover, no index is particularly developed for analysing the inherent safety level of STHE. Therefore, a newly developed Inherent Safety Index for Shell and Tube Heat exchanger (ISISTHE) is presented for assessing the inherent safety level of a STHE at the preliminary design stage.

ISISTHE is composed of preliminary design variable and integrated with the process design simulator such as Aspen HYSYS V (8.0). This integration provides an easy way of transferring the process information from process design simulator to MS Excel spreadsheet. This concept was introduced in the development of integrated risk estimation tool (IRET) [27]. A similar approach was adopted to develop inherent safety Index module (ISIM) [28] , process route index (PRI) [18], toxic release consequence analysis tool (TORCAT) [29], process stream index (PSI) [30],toxic release inherent risk assessment (TRIRA) [25] and inherent fire consequence estimation tool (IFCET) [31].

3. Methodology to Organise ISISTHE

Several failure case histories of STHE were thoroughly examined to identify the failure contributing factors and variables of STHE. The details of failure cases are given in Table 1. The methodological framework to develop an inherent safety index for STHE is given in Fig. 1. Basic failure cause and relevant variables were segregated in each failure case. Failure contributing variables acknowledged according to the relevant stage. These failure histories of STHE were acquired from various literature such as engineering failure analysis and material and design journals, investigation reports issued by chemical safety and hazard investigation board of the United States. All information is illustrated in Table 1.

Table 1. Failure analysis of STHE through various case histories.

Sr, No	Exchanger Service	Failure Causes	Contributing variables	Variable defining stage	Future direction	Ref.
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	[8]
			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	[32]
			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	-	[33]
			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 exchanger	[34]
			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 safe material	[6]
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	[35]
7.	Process gas at shell and BFW at tube side	Thermal fatigue	Excessive heating	Preliminary design stage	Timely inspection	[36]
8.	Ammonia in the shell side and process chemical in the tube side.	Over pressurization	Pressure	Preliminary design stage	Emphasize on workers safety training	[37]
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.	[38]

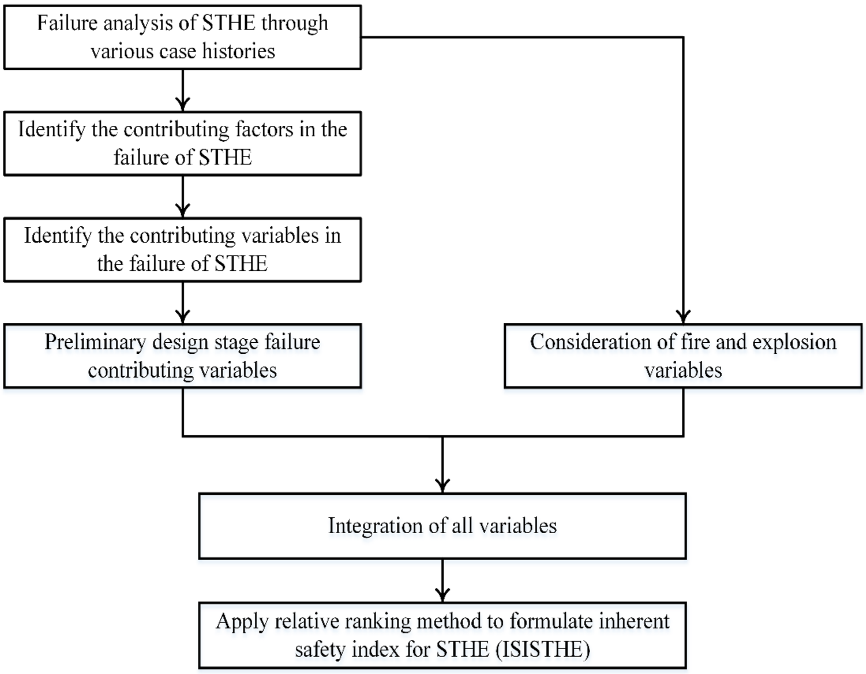

Fig. 1. Methodological framework to develop new index for STHE.

Several factors could become a failure cause of STHE. Corrosion, mechanical vibrations, inappropriate design, meagre fabrication, inappropriate construction of material, flow and heat transfer related issues were frequently observed failure contributor of STHE. These factors are influenced by several variables. It includes heat transfer rate, pressure, flow velocities and chemical reactivity. The failure of STHE frequently observed due to flow and thermal instabilities [32-34, 36]. Heat transfer and flow variables can easily be defined by the process conditions. Pressure and velocity are the most commonly used flow variable. Pressure could easily be monitored from the process condition while velocity required the exact configuration of STHE. Corrected mean temperature difference (CMTD) describes the driving force for heat transfer in STHE. Moreover, pressure (P) and corrected mean temperature difference (CMTD) can easily be evaluated from the given process conditions of STHE.

The explosions are commonly observed credible events in the chemical process industries. Catastrophic failure of STHE in Tesoro Anacortes Refinery, United Sates [8] and Sodegaura-Refinery, Japan [39] were the typically observed explosions illustrations from the failure of a STHE. The heating value (H_v) and combustibility potential (ΔFL) can be implemented to estimate the explosion potential [18]. Both of parameter can easily be estimated at the preliminary design stage. The ISISTHE is developed by the integration of the failure variables with the explosion parameters. The relative ranking approach is used to formulate this index. The same concept was adopted to develop the process stream index (PSI) [30]. The above information can be summarised by the following equations

$$\text{ISISTHE} = f(\text{Pressure, CMTD, Heating value, Combustibility potential}) \tag{1}$$

$$\text{ISISTHE} = f(P, CMTP, H_V, \Delta FL) \tag{2}$$

where P, X, H_v and ΔFL represented pressure, corrected mean temperature difference, heating value and combustibility potential with the units of kPa, ^oC, kJ/kg and % respectively. ΔFL represents the combustibility range and it can be estimated by the difference between upper and lower flammability limit.

The dimensionless number of each variable is originated by taking the ratio of that variable value for a selected heat exchanger to the average value of that variable in the whole heat exchanger network. This technique is based on the principle of relative ranking. The relative ranking is an appropriate method to analyse hazardous attributes, process conditions and operating parameters at the conceptual design stage [40]. Relative ranking can be utilized for the development of numerical index at the preliminary design stage. These numbers can be expressed by the following below equations.

$$I_p = \frac{\text{Pressure } (P) \text{ for a specific STHE}}{\text{Average value of } (P) \text{ for all heat exchangers}} \tag{3}$$

$$I_X = \frac{\text{CMTD}(X) \text{ for a specific STHE}}{\text{Average value of } (X) \text{ for all heat exchangers}} \tag{4}$$

$$I_{Hv} = \frac{\text{Heating value} (Hv) \text{ for a specific STHE}}{\text{Average value of } (Hv) \text{ for all heat exchangers}} \tag{5}$$

$$I_{\Delta FL} = \frac{\text{Combustibility potential } (\Delta FL) \text{ for a specific STHE}}{\text{Average value of } (\Delta FL) \text{ for all heat exchangers}} \tag{6}$$

These numbers can be combined to formulate an index that reflects the severity of a heat exchanger in the case of a loss of containment scenario leading to a fire and explosion. The product of these dimensionless numbers (I_P, I_X, I_{Hv} and $I_{\Delta FL}$) can be used to represent the inherent safety level of a STHE. It would result in the inherent Safety Index for STHE (ISISTHE). It can be expressed by the following equation

$$\text{ISISTHE} = I_p \times I_X \times I_{Hv} \times I_{\Delta FL}$$
(7)

The numerical value of this index represents the inherent safety level of a STHE at the preliminary design stage. It should be kept in mind that a heat exchanger would be less inherently safe if the value of the pressure, CMTD, heating value and combustibility potential are relatively higher.

The explosion is selected as the credible events in the formulation of ISISTHE. Therefore, the fluid stream possesses the higher combustibility potential (ΔFL) and heating value (Hv) would be considered in the evaluation of ISISTHE. For example, if the combustibility potential and heating value of a shell side fluid is higher than a tube side fluid, then the pressure (P) of a shell side fluid would be used to evaluate ISISTHE.

4. Case Study

Steam reforming is a widely used industrial process for the production of hydrogen gas. The process simulation diagram of a typical steam reforming unit is presented in Fig. 2. The R-4501 and R-4502 are the primary and secondary reformers used for the reforming reaction. High and low-temperature shift convertors (R-4503 and R-4504) are also installed to convert carbon monoxide (CO) into the carbon dioxide (CO_2). A network of eight heat exchangers is deployed to recover the heat of process gas leaves from the reforming section. Eventually, two phase separator (V-4501) is installed to separate out the product gases. The final product stream indicates the product gases from during this process.

This PFD is comprised of eight STHE. The process streams FEED1 to FEED8 of heat exchangers E-4501 to E-4508 are nominated for estimating the ISISTHE. These streams are selected on the basis of high heating value and combustibility potential. The inherent safety level assessment of each heat exchanger is carried out by using ISISTHE. The results are presented in Table 2. The E-4503 retains the highest ISISTHE value in this heat exchanger. Therefore, this heat has the lowest inherent safety level and can be considered as the worst heat exchanger. Moreover, the loss of containment from this heat exchanger could have a high potential of creating an explosion. This heat exchanger is required censorious attention while developing the piping and instrumentation diagram (P&ID) and layout design.

Table 2. ISISTHE analysis for the heat exchanger network of the steam reforming unit.

Heat Exchanger	Selected Stream	Pressure	Heating Value	Combustibility	CMTD
		kPa	kJ/kg	%	^{o}C
E-4501	Shell	3530.00	7710.95	63.50	210.25
E-4502	Shell	3530.00	7710.95	63.50	126.79
E-4503	Shell	3530.39	7710.95	63.50	524.45
E-4504	Tube	3530.00	7710.95	63.50	99.04
E-4505	Tube	3530.00	7710.95	63.50	21.54
E-4506	Tube	3200.00	7506.21	63.80	56.20
E-4507	Tube	3200.00	7475.50	63.84	35.91
E-4508	Tube	3200.00	7475.50	63.84	57.76
Average		3406.30	7626.50	63.62	141.49
Heat Exchanger	I_P	I_{Hv}	$I_{\Delta F}$	I_{CMTD}	ISISTHE
E-4501	1.04	1.01	1.00	1.49	1.55
E-4502	1.04	1.01	1.00	0.90	0.94
E-4503	**1.04**	**1.01**	**1.00**	**3.71**	**3.88**
E-4504	1.04	1.01	1.00	0.70	0.73
E-4505	1.04	1.01	1.00	0.15	0.16
E-4506	0.94	0.98	1.00	0.40	0.37
E-4507	0.94	0.98	1.00	0.25	0.23
E-4508	0.94	0.98	1.00	0.41	0.38

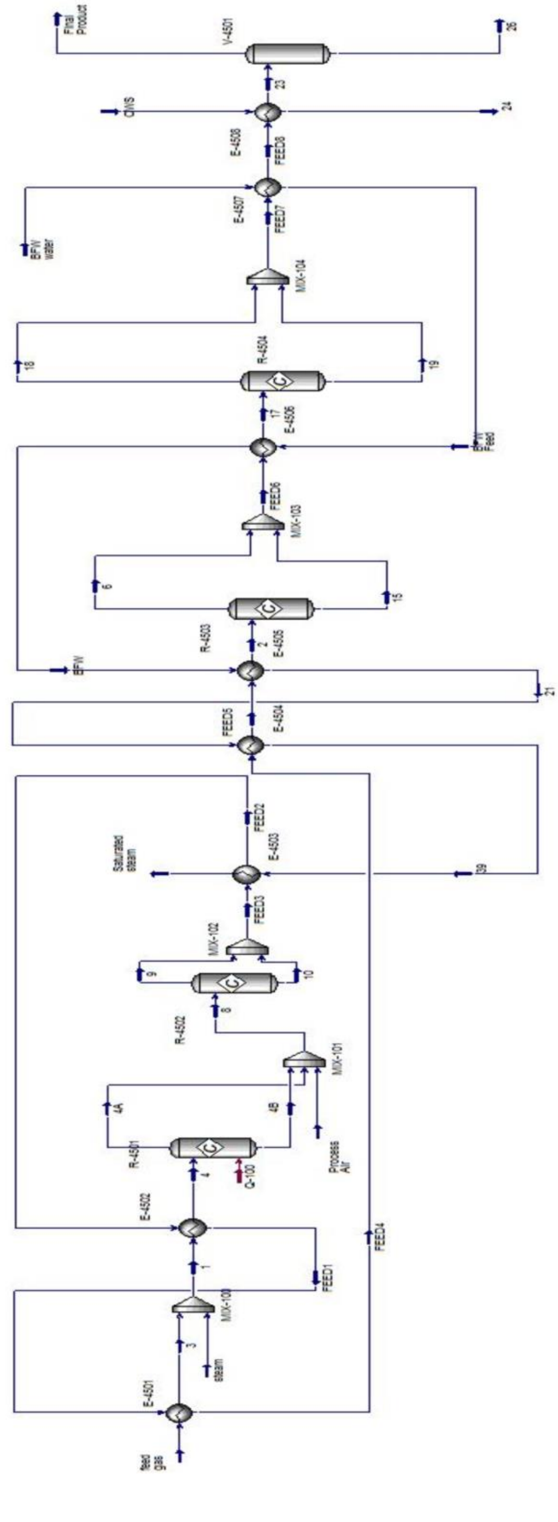

Fig. 2. Process flow diagram of steam reforming and shift conversion unit heat exchanger network .

5. Comparison of ISISTHE with Explosion Models

Explosions are deliberated as credible events in the formulation of ISITHE. Therefore, the magnitude of ISISTHE can be compared with the explosion energy released by STHE in a worst case scenario. The similar concept used to validate process stream index (PSI) [30].

5.1. Comparison with Baker Model

Baker model is frequently used for estimating the burst pressure. In this model, Brode equation is used for the estimation of explosion energy [41]. It is presented by the following equation

$$E = \frac{(P_1 - P_o) \times V}{\gamma - 1} \tag{8}$$

where P_1 and P_o are the initial and final (ambient) pressure of expanding gas, V is the total volume of gas and γ is the heat capacity ratio. The explosion energy of all heat exchangers is estimated by considering the worst-case release scenario. The estimated explosion energy and ISISTHE value of all heat exchangers are presented in Table 3. Moreover, the graph between the two parameters is given in Fig. 3. The similar configuration is observed of both trends.

Table 3. Estimated explosion energy by Baker Model and ISISTHE values.

Exchanger	Explosion Energy J	ISISTHE
E-4501	4.09E+06	1.55
E-4502	4.31E+06	0.94
E-4503	9.23E+06	3.88
E-4504	3.85E+06	0.73
E-4505	3.60E+06	0.16
E-4506	3.50E+06	0.37
E-4507	4.03E+06	0.23
E-4508	5.32E+06	0.38

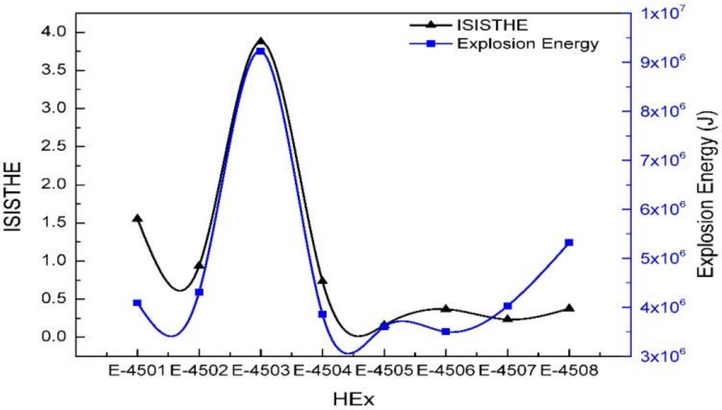

Fig. 3. ISISTHE comparison with Baker Model.

5.2. Comparison with Prugh model

Explosion energy can be estimated by using Brown equation in Prugh's model. It is based on isothermal gas expansion from atmospheric to initial conditions [41].

$$E = \frac{V}{V_R} \times \frac{P_1}{P_R} \times R \times T_1 \times \ln\left(\frac{P_1}{P_0}\right) \tag{9}$$

Where V and V_R are the initial and standard volume of gas, P_1, P_R and P_o are initial, standard and ambient pressure and R and T_1 are gas constant and the initial temperature of gas respectively. Similarly, the explosion energy of each heat exchanger is estimated by considering the worst case release event. The estimated explosion energy by using this explosion model and ISISTHE value are presented in Table 4. The trends of both parameters are given in Fig. 4. The configuration of both trends is found quite identical.

Table 4. Estimated explosion energy by Prugh Model and ISISTHE values

Exchanger	Explosion Energy J	ISISTHE
E-4501	8.88E+09	1.55
E-4502	7.35E+09	0.94
E-4503	2.93E+10	3.88
E-4504	6.84E+09	0.73
E-4505	5.42E+09	0.16
E-4506	5.47E+09	0.37
E-4507	5.15E+09	0.23
E-4508	5.51E+09	0.38

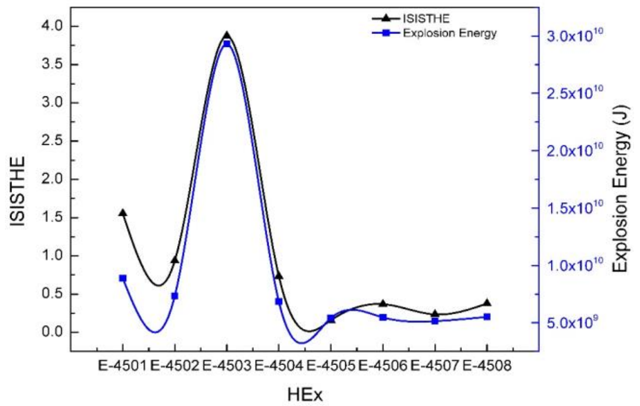

Fig. 4. ISISTHE comparison with Prugh model.

6. Conclusion

Newly developed index (ISISTHE) can easily be implemented for assessing the inherent safety level of a heat exchanger at the preliminary design stage. This index is user-friendly and required less process information as compared to other indices. Moreover, it is integrated with a process design simulator for easily transferring of process information. This index facility the design engineer to figure out the worst heat exchanger in the given heat exchanger network. This index has been validated with the well-known explosion models. Therefore, this index implicitly predicts the potential of explosion from the loss of containment of a heat exchanger. Heat exchanger retains high operating pressure, corrected mean temperature difference and the combustibility potential of process gas needs comprehensive safety analysis throughout the design process. The scope of this methodology can be extended for other consequences such as toxic release and fire.

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