Physical explosion analysis in heat exchanger network design

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Abstract. The failure of shell and tube heat exchangers is being extensively experienced by the chemical process industries. This failure can create a loss of production for long time duration. Moreover, loss of containment through heat exchanger could potentially lead to a credible event such as fire, explosion and toxic release. There is a need to analyse the possible worst case effect originated from the loss of containment of the heat exchanger at the early design stage. Physical explosion analysis during the heat exchanger network design is presented in this work. Baker and Prugh explosion models are deployed for assessing the explosion effect. Microsoft Excel integrated with process design simulator through object linking and embedded (OLE) automation for this analysis. Aspen HYSYS V (8.0) used as a simulation platform in this work. A typical heat exchanger network of steam reforming and shift conversion process was presented as a case study. It is investigated from this analysis that overpressure generated from the physical explosion of each heat exchanger can be estimated in a more precise manner by using Prugh model. The present work could potentially assist the design engineer to identify the critical heat exchanger in the network at the preliminary design stage.

1. Introduction
The explosion is defined as the sudden and violent release of energy. Violence of explosion depends upon the energy released rate [1]. Blast effect generated from the explosions can create transient change in the gas density, velocity and pressure of the surrounding air. Moreover, the loss of containment of pressurized gas, boiling liquid, flammable material and the reaction runaway could potentially lead to various types of explosions. Few well-known explosion mechanisms are presented in the Table 1.

Shell and Tube Heat Exchangers (STHE) are being extensively deployed in the chemical process industries. These heat exchangers serve as heater, cooler, partial or total condenser, evaporator, decomposer and boiler. Tubular Exchanger Manufacturing Association (TEMA) is responsible for issuing and updating its design guidelines on a regular frequency [2]. The latest version of these guidelines was updated by TEMA in 2007. The failure of STHE could potentially cause an enormous production loss and

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have a potential of credible events such as explosion and fire. Occasionally, the failure of these heat exchangers has also been observed just after a few months of service [3, 4]. On 10th June, 2008, physical explosion observed through bursting of heat exchanger due to over pressurisation in rubber industry located in Houston, United State [5]. Recently, the catastrophic failure of STHE in Tesoro Anacortes Refinery, Washington caused seven fatalities [6].

### Table 1 Various types of explosion

| Explosion Type                               | Characteristics                                                                 |
|----------------------------------------------|---------------------------------------------------------------------------------|
| Vapor cloud explosion (VCE)                  | Resulted from the combustion of flammable loss of containment                   |
| Boiling liquid expanding vapor explosion (BLEVE) | Resulted with the rapid released of pressurized liquid above from its atmospheric boiling point |
| Physical explosion                           | Caused by the catastrophic failure of pressurized system                        |
| Chemical explosion                           | Explosion resulted from the uncontrollable chemical reaction                    |

An inadequately designed STHE could potentially have a high failure potential. The design errors frequently cause accidents in the chemical process industries [7, 8]. The failure frequency of STHE is relatively higher as compared to other process equipment [9]. Therefore, lack of consideration in performing safety analysis could possibly lead towards the failure of STHE. Hence, there is a need to properly explore the consequent impacts of explosions during the heat exchanger network design. Process design can easily be modified in the preliminary design stage. Although the level of process information is lower at this design stage, but process design can be changed smoothly and economically at this initial design stage [10]. The over pressure effect resulted from the physical explosions of heat exchangers is primarily discussed in this work.

## 2. Physical Explosion Models

### 2.1 Baker’s Model

Baker et al. (1983) developed a model for estimating the burst pressure. In this method energy of explosion is evaluated by using Brode equation [11]. It is represented by below equation

\[
E = \frac{(P_f - P_o) \times V}{\gamma - 1}
\]  

\[E\]  

\[
E = \frac{(P_f - P_o) \times V}{\gamma - 1}
\]  

\[
P_i\] and \(P_o\) are the initial and final (ambient) pressure of expanding gas (Pa), \(V\) is total volume of gas (m³), \(\gamma\) is heat capacity ratio and \(E\) is energy (J) of expanding gas. This expression indicates the required energy to raise the gas pressure from atmospheric pressure to initial pressure by keeping volume constant. However, the real gas explosions behave differently from this assumption. The expansion energy is partly
dissipated in vessel deformation and rupture. The remaining explosion energy is utilized to generate blast over pressure. Peak overpressure \((P')\) can be estimated from the graph as mentioned in Figure 1 by using Scaled distance \((R')\).

\[
R' = r \times \left( \frac{P_o}{E} \right)^{\frac{1}{7}}
\]  \hspace{1cm} (2)

Where \(r\) is absolute distance (m) and \(P_o\) is ambient pressure (Pa).

![Figure 1. Estimation of overpressure from scaled distance in Baker Model [11]](image)

2.2 Prugh’s 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 [11].

\[
E = \frac{V}{V_R} \times \frac{P_1}{P_R} \times R \times T_1 \times \ln \left( \frac{P_1}{P_0} \right)
\]  \hspace{1cm} (3)

Where \(R\) is ideal gas constant, \(V\) is initial volume of compressed gas, \(V_R\) and \(P_R\) are volume and pressure at standard conditions. Explosion energy \((E)\) is converted into equivalent mass of TNT and it can be used to estimate the blast effects.

\[
W = 1.39 \times 10^{-6} V \left( \frac{P_1}{P_0} \right) R T_1 \ln \left( \frac{P_1}{P_2} \right)
\]  \hspace{1cm} (4)

Where \(W\) is equivalent mass of TNT, \(V\) is volume of compressed gas, \(P_1, P_2\) and \(P_0\) are the initial, final and standard pressure and \(T_1\) is the initial temperature of the expanding gas respectively. The analogy of
the explosion of a pressurized gas container to a condensed phase point source of TNT is not adequate in the near field. As vessel is not a point source. Prugh suggested a modified method for estimating the virtual distance from explosion center. It can be estimated by subtracting the geometrical distance between center and external surface of the vessel. The actual distance is obtained by adding virtual distance to selected distance from the source. Maximum overpressure at the contact surface between the expanding gas and air can be evaluated by following equation [11].

\[
P_b = P_a \left[ 1 - \frac{3.5(\gamma - 1)(P_a - 1)}{\sqrt{\frac{\gamma T}{M}}(1 + 5.9P_a)} \right]^{\frac{\gamma - 1}{\gamma - 1}}
\]

Equation (5)

\(P_b\) is the burst pressure of the vessel and \(P_a\) is the pressure at the surface of the vessel in bar. Scaled distance (Z) would be required to evaluate scaled over pressure (\(P_s\)), and it can be evaluated by TNT equivalent mass (W) and source distance (r). Over pressure (\(P_s\)) can be estimated by using Figure 2.

\[
Z = \frac{r}{(W)^{\frac{1}{2}}}
\]

Equation (6)

**Figure 2.** Estimation of overpressure from scaled distance in Prugh Model [11]

3. Case study of typical heat exchanger network
Steam reforming and shift conversion process are widely used for the production hydrogen (H\(_2\)) and carbon dioxide (CO\(_2\)) on the commercial scale. The simulated PFD of the process is given in the Figure 3.
Figure 3. PFD of steam reforming and shift conversion process

The physical explosion effect is analysed for each heat exchanger by using both explosions models. The overpressure magnitude is evaluated at a constant distance from source and presented in the Table 2.

Table 2

| Heat exchanger | Bakers Model | Prugh Model |
|----------------|--------------|-------------|
| E-4201         | 1.02E+07     | 2.22E+10    |
| E-4202         | 1.08E+07     | 1.84E+10    |
| E-4204         | 1.15E+07     | 3.66E+10    |
| E-4205         | 9.63E+06     | 1.71E+10    |
| E-4206         | 9.00E+06     | 1.35E+10    |
| E-4207         | 8.53E+06     | 1.33E+10    |
| E-4208         | 9.78E+06     | 1.25E+10    |
| E-4209         | 1.29E+07     | 1.34E+10    |

The over pressure value estimated from Bakers model is much higher than the Prugh model. Baker model is highly influenced by the physical properties of the working fluid and Prugh model is less influenced by the physical properties, but much sensitive to temperature and may be preferred for high temperature process equipment [12]. The trends of explosion energy and over pressure values are plotted to analyse the behaviour of both parameters in both explosion models. These trends are presented in Figure 4 and 5.
Figure 4 Comparison of explosion energy and over pressure in Baker Model

Wide variation is observed in the trends obtained from the Baker Model. However, similar configuration of the both parameter trends is observed in the Prugh model. The overpressure generated by the physical explosion of E-4204 is found maximum among the all the heat exchangers. Therefore, this heat exchanger is the critical one in the network and required a detailed safety review.

Figure 5 Comparison of explosion energy and over pressure in Prugh Model

4. Conclusion
Overpressure effect generated from physical explosion is analysed for the heat exchanger network by using two well-known explosion model. Prugh model can deliver a better approximation of overpressure as compared to the Baker model due to its consistent response for temperature sensitive equipment. Moreover, this analysis could potentially assist the design engineer to figure out the critical heat exchanger in the heat exchanger network.
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