Dynamic simulation of Texas City Refinery explosion for safety studies

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The catastrophic explosion that occurred at Texas City on 23 March 2005, during the start-up of the raffinate splitter led to an estimated 15 deaths and 180 injuries. Since the incident, several studies have investigated the root causes of the incident, and wider organisational, process safety management and human elements that contributed to the incident. There have also been some attempts to model the sequence of events before the incident, and consequences of the resulting fire and explosions.

This study attempts to model the sequence of events and replicate the reported process variables during the isomerisation unit start-up on the day of the incident. The resulting dynamic simulation model is used as the framework for further safety studies for similar processes, and will hopefully contribute towards safer operations in the chemical and oil & gas process industries.

Keywords: Texas City, Dynamic simulation, HAZOP, Process Hazard Analysis

Introduction

The explosion that occurred at BP's Texas City Refinery on 23 March 2005 remains one of the most catastrophic incidents in the history of the process industries. The explosion resulted in the deaths of 15 persons, and over 180 were injured. It is estimated that economic losses of up to US$1.5b resulted from the incident and follow-up efforts at reconstruction. This incident has been the subject of several process safety analyses, and lessons learned from the incident investigation have been used by process industries around the world for improving technical safety performance, and for developing better process safety management processes. The US Chemical Safety Board (US CSB, 2005) reports that the incident was due to organisational and safety deficiencies at all levels of the organisation as warning signs were present for many years.

The incident happened as the raffinate splitter section of the isomerisation unit (ISOM unit) was being re-started after a turnaround maintenance. Evidence suggests that the start-up process was being carried out contrary to instructions in BP's start-up procedures following major turnaround maintenance work. The operators proceeded to fill the raffinate tower with hydrocarbon liquid feed for over three hours with no liquid outflow from the tower. This resulted in the tower overheating and overfilling as a result of thermal liquid expansion. In addition to operator negligence and error, incorrect functioning of critical instrumentation and control devices also contributed to the catastrophic explosion.

Although the accident occurred over 8 years ago, valuable lessons have been deduced from analysing the operational and process design errors that led to the catastrophic loss of containment and subsequent explosion. The incident is widely used as an important case study in academic circles, and informs the process safety management approach adopted in many high profile chemical process industries. Numerous published academic papers on the subject also exist.

This work aims to extend the application of the lessons learned from the explosion by offering an approach to hazard and operability studies that will contribute to safer operations in the process industries. The UK Health and Safety Executive acknowledges the need to "*engage an industry whose performance in risk management and control is not what it should be*" (HSE, 2007). The methodology used involves an initial dynamic simulation of the sequence of events that led to the explosion using commercially available process simulation software. This is followed by a further simulation scenario that explores an alternative sequence of events and likely consequences.

Literature review

Shortly after the explosion that occurred at Texas City, the US Chemical Safety Board initiated a thorough investigation that concluded with the publication of an investigation report in 2007. The report provides detailed accounts of the refinery's operations, ISOM unit description, and technical details of the accident sequence. The US CSB report also presented its findings on BP's process safety management approach, and alluded to lapses in the company's safety culture, and a lack of management commitment to process safety improvement. Amongst other things, the report identified a flawed management of change process which meant that temporary trailers were located too close to the ISOM and NDU (Naptha Desulfurization) units. Another concern raised by the report was the poor implementation and close-out of actions arising from process hazard analysis (PHA) studies.

The Baker panel report (Baker et al. 2007) commissioned by BP was also published in 2007. The Baker report offered 10 recommendations for process safety improvement in the process industries. It is noteworthy that most of the recommendations from the CSB and Baker panel reports have been adopted across the process industries around the globe. For instance, there is considerable emphasis on identifying and managing relevant process safety metrics (process safety performance indicators) as opposed to occupational safety data that give no indication of the process safety performance of a facility. However, a lot more needs to be done in order to achieve the aspiration of transforming the process industries into high reliability organisations.

A dynamic model of the events leading up to the Texas City refinery explosion is presented by Palacin-Linan (2005) and implemented in gPROMS. Palacin-Linan argues that because of the complexity of the distillation column operation on the day of the incident, it is not possible to replicate the sequence of events using commercially available simulation tools such as Aspen HYSYS or UNISIM. However, on the day of the incident, there was no distillation taking place in the column as it was simply being filled with liquid (Manca and Brambilla, 2012). Commercial simulators thus provide the capability to replicate the column filling dynamics, with reasonable approximations of the vapour-liquid equilibriums at the published heating rates and the physicochemical properties of the column components. In addition to a dynamic simulation of the column filling dynamics, Manca and Brambila unsuccessfully attempted to simulate the dynamics of the two-phase flow of sub-cooled liquid in the blowdown pipe using the Homogenous Equilibrium Model, HEM. The inability to successfully explain the dynamics of events downstream of the pressure relief valves (PRVs) is attributed to lack of critical information such as the type of PRVs, and the inlet and outlet process conditions to the PRVs. Although it is suggested that future research efforts be focused on using alternative models to explain the complete evolution of the accident dynamics, it is doubtful if such work will inherently provide substantial value in improving process safety. It is therefore not the intention of this paper to extrapolate or validate the results of dynamic process simulations of the Texas City explosion. In fact, close examination of the different failed barriers that led to the catastrophic explosion will reveal that concentrating our efforts at making sure that process safety barriers (both passive and active barriers) function on demand will be far more beneficial. The initial step in having the right barriers is the correct identification of the hazards present in the process from the design phase and throughout the asset lifecycle.

A few technical publications have attempted to model the impacts of the accident in order to quantify the resulting damage from the fires and explosions. Khan and Amyotte (2007) adopt a consequence modelling approach that uses the quantitative risk assessment methodology to quantify the extent of damage from fires and explosions following the incident. Similarly, Kalantarnia et al. (2010) describe a predictive model that uses the Texas City explosion event for consequence assessment as a useful tool to forecast the impacts of loss of containment incidents.

One of the outcomes of the Baker panel report into the Texas City refinery incident is the recommendation that robust process hazard analysis (PHA) is carried out on process plants. Hazard and operability study (HAZOP) is regarded as one of such PHA tools for the systematic identification and assessment of process and operability hazards (Herbert, 2011). The effectiveness of a HAZOP team depends on the creativity and experience of the participants in discovering unforeseen effects that might result in a major accident hazard. This is no easy task as the team works with final design documents prepared by clever engineers who would have thought of most credible hazard scenarios. In addition to being creative, the team therefore needs to be highly critical in order to identify latent errors in design. To do this, the team must systematically probe the process design in a sustained manner over many hours of debating process design errors.

It has been suggested (Mahnken, 2001) that case histories from previous accidents can be a powerful tool in stimulating and sustaining the creativity of HAZOP team members. Mahnken suggests that in addition to presenting the basic sequence of events of the incident, a hypothetical HAZOP worksheet that illustrates how the accident might have been foreseen in a HAZOP study should also be considered. Kletz (1999) argued that *"It is better to illuminate the hazards we have passed through than not illuminate them at all, as we may pass the same way again, but we should try to see them before we meet them…unfortunately, we do not always learn from the hazards we have passed through."* It is therefore the intention of this paper to explore a method for illuminating the hazards inherent in distillation column operations, and hopefully start a discussion on the usefulness of quantitative HAZOPs in promoting management buy-in in the often expensive process of satisfactorily closing out PHA recommendations.

The HAZOP methodology is qualitative industry best practise procedure for identifying and controlling process safety hazards. Like other PHA tools, HAZOPs help to identify unacceptable hazards. They do not reduce the hazards to acceptable levels. Reducing the hazards only happens when HAZOP actions are effectively closed out. However, because it involves considerable investments of time, cost and personnel expertise to close out HAZOP actions, it is usually not a top priority in many organisations. The challenge for process safety practitioners, therefore, is to ensure that management support is obtained and necessary resources are committed into closing out PHA recommendations. Unlike other PHA tools such as Layers of Protection or Fault Tree Analysis, HAZOP studies do not quantify risk levels (as this is not the primary aim of a HAZOP). Typically, high hazard scenarios identified during a HAZOP study are subjected to more rigorous PHA studies such as LOPA and the required Safety Integrity Level is identified to mitigate the identified hazard. This is often a lengthy process. Johnson (2010) proposes the use of a combined HAZOP/LOPA study to immediately quantify the risk reductions possible with available safeguards and assign a target SIL. This approach, according to Johnson, will avoid the duplication of efforts involved in conducting separate HAZOPs and LOPA studies. The practical application of this approach is however doubtful as the HAZOP process is inherently time consuming and requires focused deliberation over a lengthy period of time in order to correctly identify hazards.

Dynamic simulation of raffinate splitter

The simulation of the sequence of events that led to the catastrophic explosions at the BP Texas City refinery is based on the hypothesis that the raffinate splitter (distillation column) overfilled. Subsequently, liquid in the column spilled over the top into the overhead pipe, leading to over pressurisation and opening of the relief valves upstream of the blowdown vessel. The simulation approach adopted in this study replaces the distillation column model in Aspen HYSYS with a tank separator. Prior to the explosion on 23 March 2005 at 01:13pm, it is reported that the column was completely liquid full (flooded) with a layer of sub-cooled liquid at the top, with the condenser and reflux drum filled with pressurised nitrogen. No vapour flowed out of the column, and there was no reflux into the column. Indeed, laboratory tests of liquid samples recovered from the column feed, bottoms product, and overhead samples demonstrated that no separation took place in the column on the day of the incident (US CSB, 2005). This justifies the approach used in this study where the distillation tower is replaced with a tank separator that is simply filling up with liquid feed. It has also been suggested that the behaviour of the vapour generated in the column on the day of the incident cannot be accurately predicted, as the prevailing process conditions are outside the range of standard empirical models of distillation tray separation available in commercial simulation packages such as Aspen HYSYS.

The tank separator is modelled as a vertical vessel with a total liquid volume of 583 m3. The volume of the separator is taken as the difference between the total internal volume of the actual distillation column less the volume occupied by column internals such as trays, weirs and downcomers (Manca and Brambilla, 2012). The dynamic simulation sought to demonstrate the following: Liquid over-fill in the distillation column; Liquid thermal expansion in the column as a result of feed pre-heat; and feed vaporisation dynamics in the distillation column. According to data provided by BP (Palacin-Linan, 2005), the feed to the column was a mixture of 35 light hydrocarbons. However, for the purpose of this simulation, the components have been lumped into four hydrocarbon categories; C5, C6, C7 and C8. All the components with 5 carbon atoms have been lumped in the C5 category, while components with 6, 7 and 8 carbon atoms have been lumped into the hexane, heptanes and octane categories. In addition to hydrocarbons, nitrogen is included in the components list as nitrogen was used to pressurise and test the system for leaks prior to the introduction of hydrocarbon feed. Also, water is included in the component list, as the column was steamed out following nitrogen purge.

The process flow sheet (PFS) used for the dynamic simulation is shown in Figure 1 (adapted from Manca and Brambilla, 2012). It comprises the tank separator used to model the distillation column and a simple TEE separator used to simulate the separation of light raffinate from nitrogen used to pressurise the system prior to the introduction of feed. For the purpose of dynamic simulation, two flow control valves are included in the flow sheet: FIC-1 is used to control the inflow of feed into the column, and FIC-2 is used to control the outflow of product from the column. The actual distillation column had a level controller for manipulating the flow of product from the bottom of the column to storage but on the day of the incident, this level controller malfunctioned.

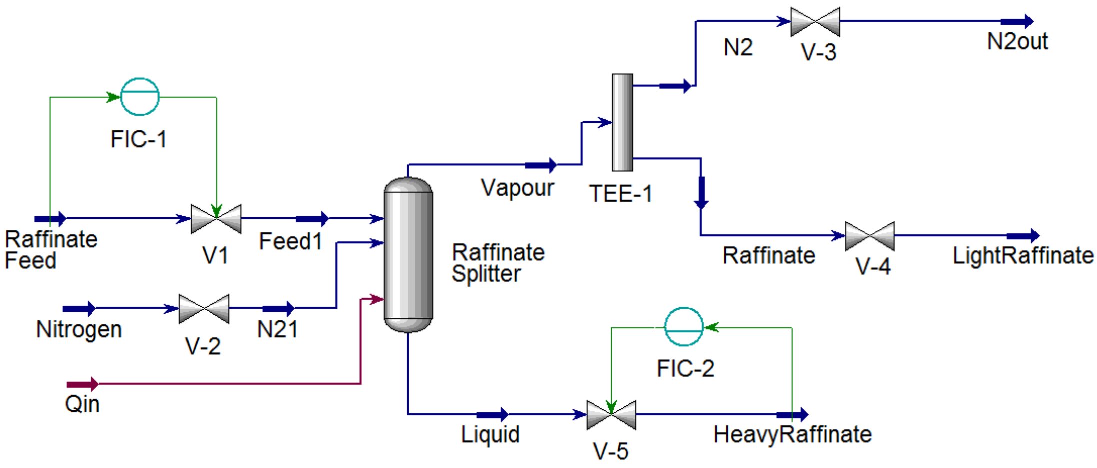

Figure 1: Process Flowsheet for dynamic simulation

The initial conditions used for the steady state simulation are those available in published documents about the process conditions at the start-up of the raffinate splitter tower on 23 March at the end of the turn-around maintenance (Palacin-Linan, 2005; US Chemical Safety Board, 2005, Baker, Levenson, and Bowman, 2007). The simulation covers the period from 0213hrs to 1320hrs on 23 March 2005, corresponding to the time raffinate feed was introduced into the tower and when the explosion occurred, respectively.

The first simulation is for the sequence of events from 0213hrs to 1000hrs when the night shift crew started introducing raffinate feed into the raffinate splitter from the aromatics recovery unit. Because of variations in the reported feed rate to the column during the start-up procedure, an average feed flowrate of 16.3 Kg/s was used for the simulation. An isothermal temperature regime is assumed for this stage of the simulation as no heat input to the splitter is reported. The feed enters the column at a temperature of 23°C. Following crew changeover at 0600hrs, the raffinate splitter start-up procedure was resumed at approximately 1000hrs on the morning of 23 March, 2005. At this stage, the fuel gas supply to the reboiler furnace was initiated and the first two burners were lit. Feed pre-heat subsequently occurred by convective heat transfer in the top section of the furnace, increasing the temperature of the feed at a rate of 8.5°C per hour. During this period in the start-up sequence, the level control valve (LCV) at the outlet of the raffinate splitter remained closed, contrary to BP’s start-up procedures. The correct procedure required that the level control valve was left at 50% open in automatic mode, to establish heavy raffinate flow to storage. Operators had previously complained that leaving the level at 50% was not ideal (contrary to what is stated in the start-up procedure). They maintained that if the level was left at 50%, a drop in liquid level could result in completely loosing heavy raffinate flow from the tower, resulting in a trip to the feed supply, and costly interruption to the start-up procedure.

A dynamic simulation of the overall tower filling dynamics from 1000 to 1320 when the explosion occurred revealed that the feed to the column vaporised at approximately 1310 hrs. This happened as a result of the additional heat input into the column though the feed-product heat exchanger. The liquid level in the column from 1000hrs to 1320 hrs is shown in Figure 2.

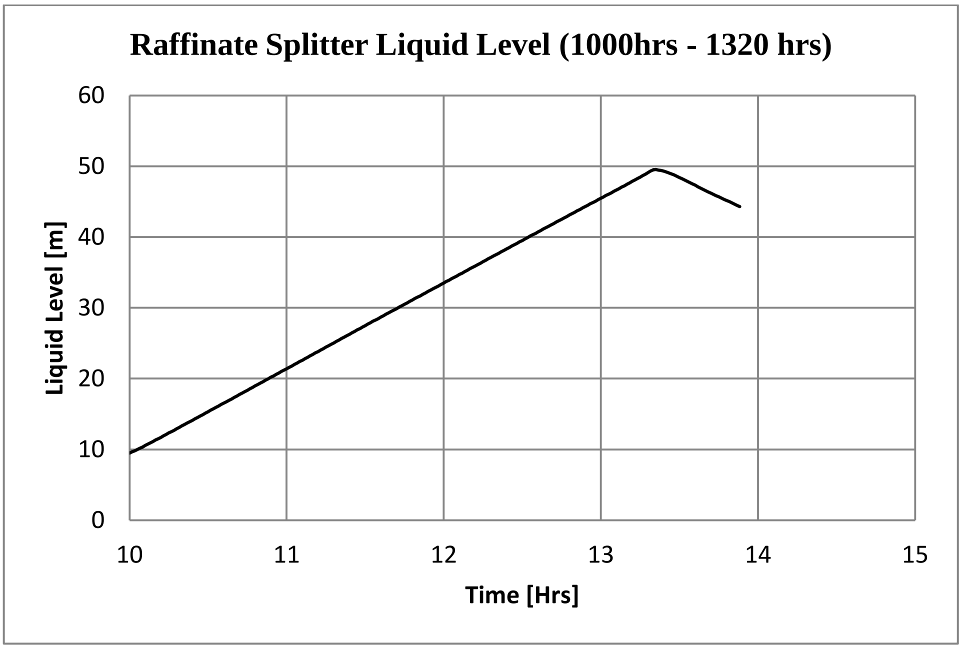

Figure 2: Column liquid filling dynamics

The dynamics of the feed vapourisation as the rate of feed pre-heat increased at 1300hrs is shown in Figure 3. The feed completely vaporises at 1310hrs. This is expected to occur and agrees with reported observations on the day of the incident, and from simulation results by other researchers.

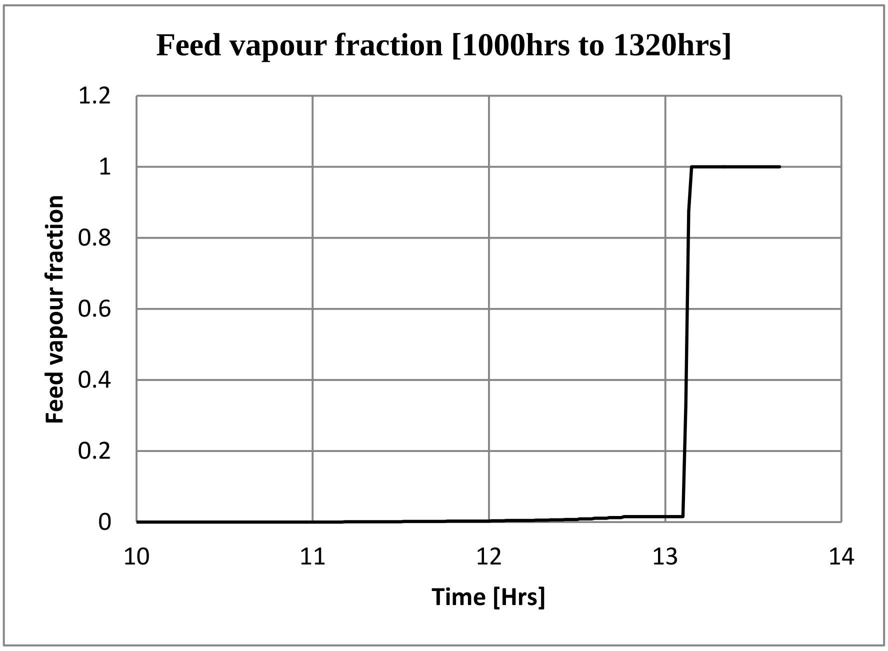

Figure 3: Feed vapourisation dynamics

The overall feed temperature increase obtained from the dynamic simulation is shown in figure 4.

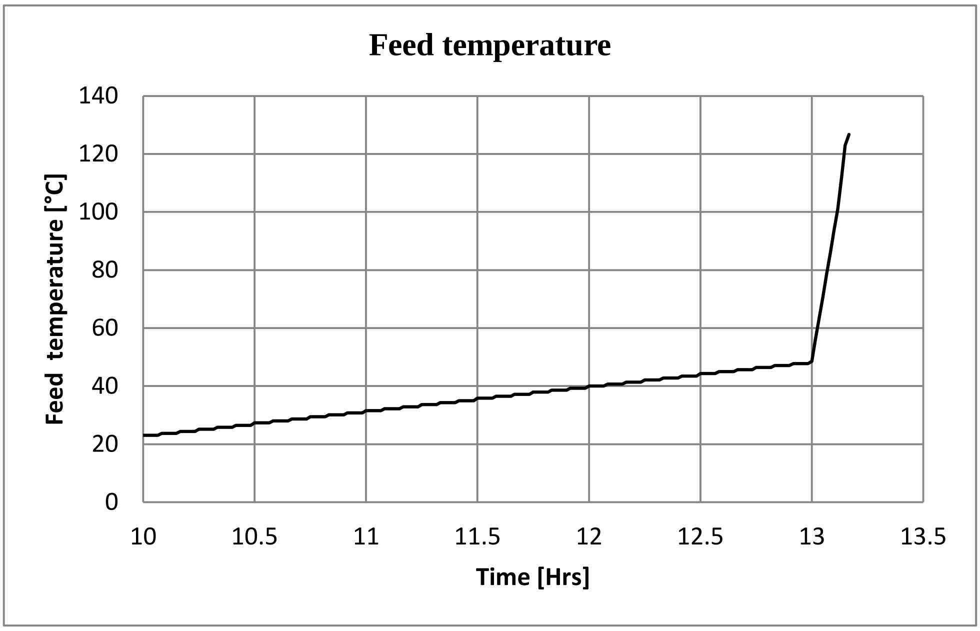

Figure 4: Feed temperature dynamics

Quantitative HAZOP

In order to extend the use of the dynamic simulation, an additional simulation scenario was investigated. It has been established that the simulation approach employed in this study is a tank model that replicates the dynamic filling of the raffinate splitter like it happened on the day of the incident. This approach is also used in the dynamic HAZOP study. The initial dynamic simulation has been used to establish that it is possible to replicate the column temperature and pressure profiles, and the liquid level dynamics observed on the day of the accident.

The significance of this methodology is not in the rigorous simulation of the sequence of events, but in the extrapolation of the model for use in dynamic HAZOP studies. It can be argued that the benefits obtainable from such an approach exceed whatever gain may be obtained from developing more accurate models. Such benefits include a visual representation of the quantified level of risks resulting from incorrectly designed equipment, or not following adequate operational procedures.

A conventional HAZOP involves using guidewords to qualitatively investigate the hazards and operability issues that would result from incorrect operations. Examples of such guidewords include “more flow”, “more temperature”, “no flow”, “less pressure”, and so on. For the purpose of this study the guideword used is “more temperature”. According to the accident investigation report, the observed temperature profile occurred as a result of heat input from two burners in the reboiler which were lit at approximately 1000hrs.Two additional burners were lit at 1117hrs. The HAZOP scenario explored in this study therefore considers the case where increased heat input to the column began at 1000hrs with all four burners lit. A qualitative assessment of this scenario would obviously identify the associated hazards of increased temperature input and likely overpressure of the column. Appropriate safeguards would be the pressure relief valves and the blowdown drum and stack which will protect the column from any overpressure scenario. The additional lapses that occurred on the day of the incident such as leaving the LCV at the bottom of the column closed are also included in this scenario. The resulting dynamic process conditions from this simulation are shown in figures 5 to 7.

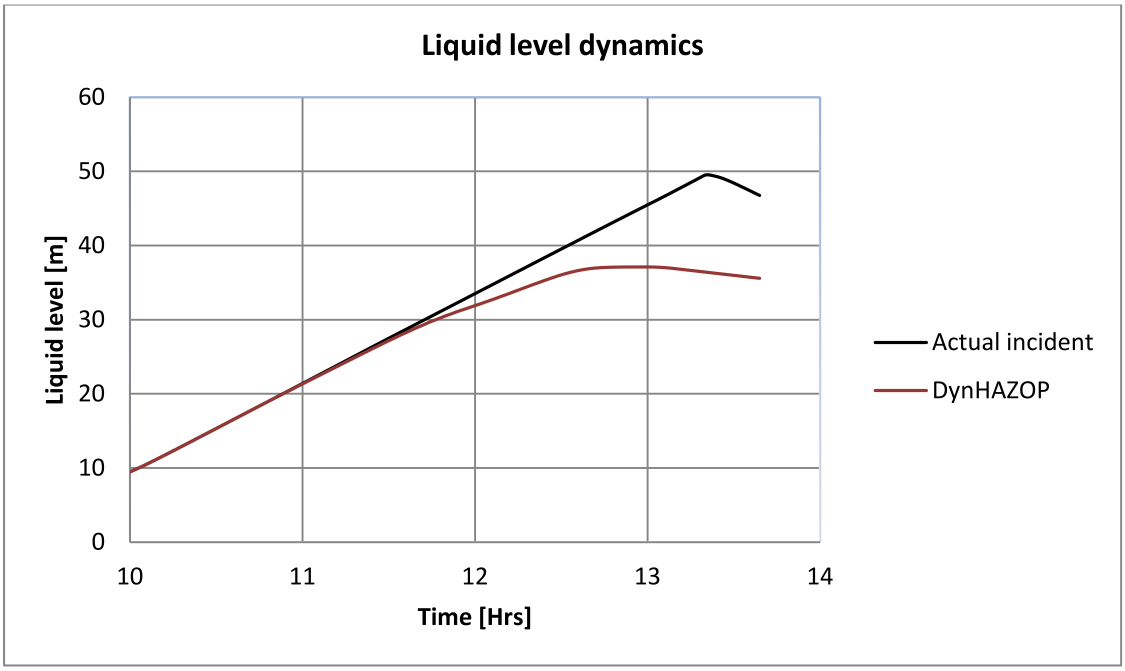

Figure 5: Dynamic HAZOP liquid level dynamics

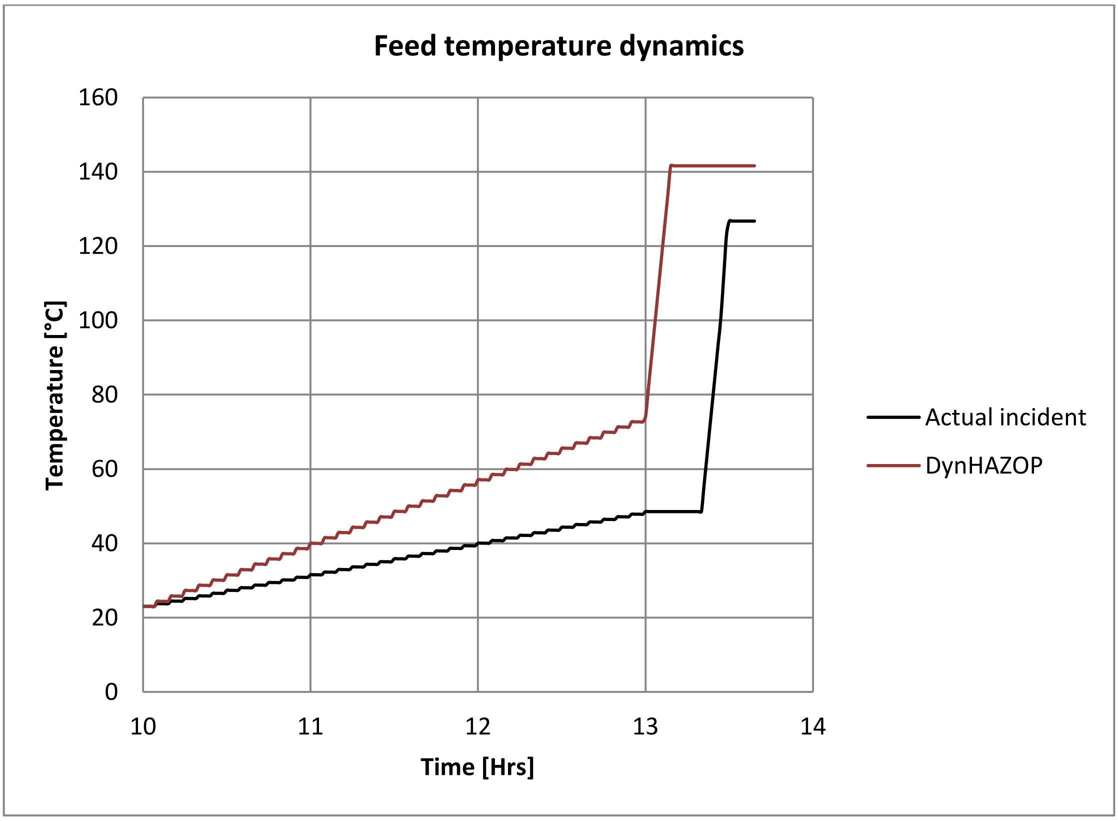

Figure 6: Dynamic HAZOP feed temperature dynamics

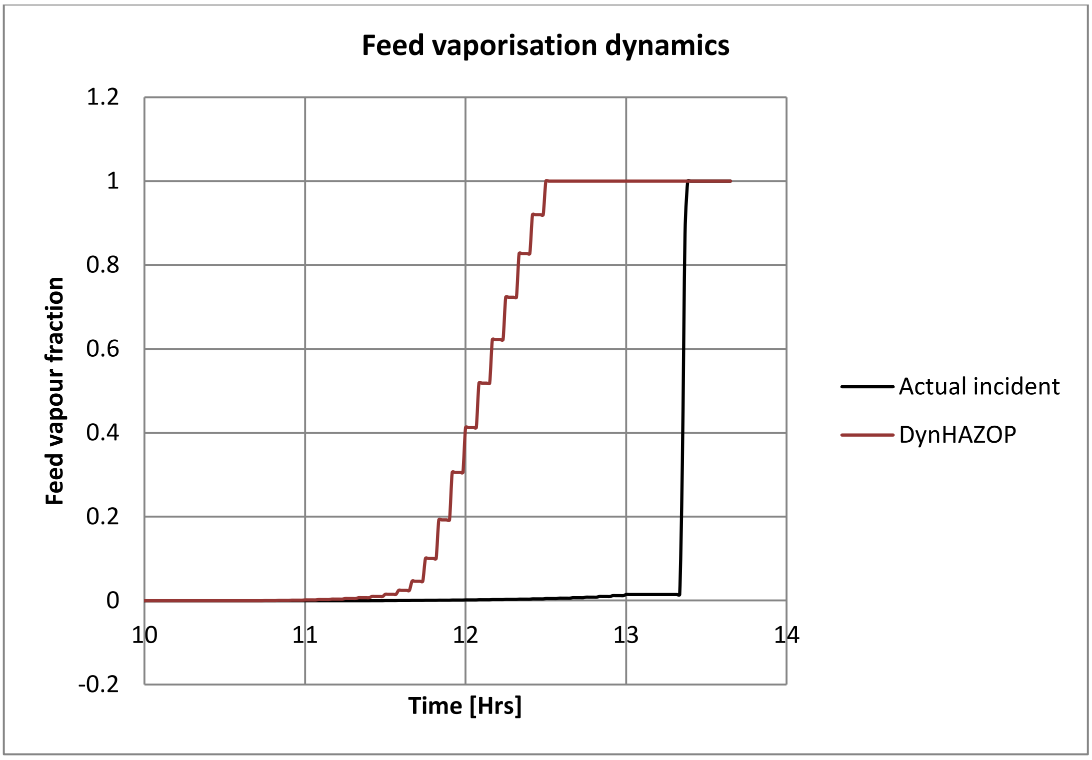

Figure 7: Dynamic HAZOP feed vapourisation dynamics

Discussion & Conclusion

The observed temperature profile obtained from Aspen HYSYS dynamic simulation closely matches that reported by the simulations of Manca and Brambilla (2012) which was implemented in UNISIM. At 1300hrs, the feed temperature had risen to 48.5°C, and the vapour fraction was 0.015. At 1:12pm, the liquid level obtained from the dynamic simulation was 47.9m, similar to a level of 48m at 1:02pm reported in the CSB report. At 1314hrs, it is reported that sub-cooled hydrocarbon liquids flowed out of the top of the column into the vertical overhead vapour line. At this stage, it is reported that the rate of temperature increase of the feed into the tower was at an additional 7.5 °C per minute. This is because the heavy raffinate product flow transferred additional heat to the feed through the feed-product heat exchanger located upstream of the reboiler. This led to a situation where the column contained a stream of heated liquid at the bottom of the tower while a layer of cold liquid remained at the top section of the column.

The dynamic simulation of the overall tower filling dynamics from 1000 to 1320 when the explosion occurred revealed that the feed to the column vaporised at approximately 1310 hrs. This happened as a result of the additional heat input into the column though the feed-product heat exchanger. Subsequently, thermal expansion of the liquid in the column led to the filling of the overhead vapour line with hydrocarbon liquids and an increase in pressure as a result of the hydrostatic liquid head. The pressure in the tower at this point is reported to have been about 63 psig. The combined liquid static head and the tower pressure led to an increase in pressure in the overhead vapour line higher than the relief valve lift set pressures of 40, 41, and 42 psig. Flammable hydrocarbon vapours subsequently flowed from the overhead line through the collection headers into the blowdown drum. It is reported that liquids flowed from the raffinate splitter tower into the blowdown vessel at a rate of 509, 500 gph, resulting in a discharge of approximately 51,900 galons of liquids into the blowdown drum in six minutes. Once the blowdown drum and stack overfilled, flammable liquid spilled to the ground and created a vapour cloud around the ISOM unit. The vapour cloud exploded at 1320hrs and the likely ignition source is reported to most likely have been an idling diesel pickup truck.

The dynamic HAZOP simulation demonstrated an alternative pathway for the sequence of events leading up to the accident. At 12noon, an additional temperature increase of 43% occurred, leading to a mixed phase feed flow with a vapour fraction of 0.4. It is noteworthy that the rise in liquid level in the column is less than that which supposedly took place on the day of the incident. Complete vaporisation of the feed takes place at 1230pm in the alternative accident pathway, much earlier than the reported feed vaporisation time on the day of the incident.

The dynamic HAZOP in this study has demonstrated an alternative pathway for the evolution of process variables if additional safety barriers were breached on the day of the incident. Establishing the evolution of the process variables downstream of the column has not been considered in this study. However, it has been demonstrated that a dynamic HAZOP may help secure management buy-in by demonstrating the likely consequences of deviations from standard operating procedures or other hazards in process plants.

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