QUANTITATIVE RISK ASSESSMENT (QRA) OF A PETROLEUM REFINERY

A DISSERTATION

Submitted in partial fulfillment of the
requirements for the award of the degree
of
MASTER OF TECHNOLOGY
in
CHEMICAL ENGINEERING
(With Specialization in Industrial Safety and Hazards Management)

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CANDIDATE'S DECLARATION

I hereby declare that the work which is being presented in the dissertation entitled "**QUANTITATIVE RISK ASSESSMENT (QRA) OF A PETROLEUM REFINERY**", in partial fulfillment of the requirements for the award of the degree of Master of technology in Chemical Engineering with specialization in "**Industrial Safety and Hazards Management**", and submitted to the **Department of Chemical Engineering, Indian Institute of Technology, Roorkee**, is an authentic record of the work carried out by me during the period June 2007 to June 2008, under the guidance of **Dr. Nidhi Bhandari** and **Dr. Kailas L. Wasewar**. The matter embodied in this work has not been submitted for the award of any other degree.

Date: 02·06·08

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Sivakumar

(Menedi Siva Kumar)

CERTIFICATE

This is to certify that the above statement made by the candidate is correct to the best of my knowledge.

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ABSTRACT

The growing importance in the area of environment protection is forcing Chemical Process Industry (CPI) to use the state-of-the-art technology. This in turn increased the complexity of the process. The diversity in the products manufactured by CPI has made it more and more common for these industries to use reactors, conduits and storage vessels in which hazardous substances are handled at extreme conditions. The accidental release of the inventory results in toxic release, fires and explosions. The increased public awareness towards this issue has prompted the authorities to develop tools and techniques for carrying out risk analysis and assessment of CPIs. This Quantitative Risk Assessment (QRA) does not end with the identification and analysis of hazards but it also provides valuable inputs for the formulation of the Emergency Procedures and the Risk Management Plans.

In this report, a comprehensive literature review along with a general QRA methodology is presented. Risks associated with Hindustan Petroleum Corporation Limited refinery facilities were assessed as a case study. This was the first oil refinery on the East Coast and the first major industry in the city of Visakhapatnam, Andhra Pradesh.

Initially major hazards in the plant were identified. Based on this, consequence analysis and frequency analysis were performed for all the events which include the selection of methodology to describe the consequence and attaching probability to estimate the frequency. For some events fault trees were also drawn to find the frequency.

Finally, the results from the consequence analysis were presented in terms of damage distance from the release point. Individual risk contours were plotted for LPG storage, which was the worst case scenario. The F-N curve was drawn to present the societal risk. Various changes were recommended to ensure that the risk posed to the society comes into the acceptable region.

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ACRONYMS

AIChE	American Institute of Chemical Engineers
ARU	Amine Recovery Unit
BLEVE	Boiling Liquid Expanding Vapour Explosion
CCPS	Center for Chemical Process Safety
CDU	Crude Distillation Unit
CPI	Chemical Process Industry
DHDS	Diesel Hydro Desulphurisation Unit
FCCU	Fluidized Catalytic Cracking Unit
FMEA	Failure Mode Effect Analysis
FTA	Fault Tree Analysis
HAZOP	Hazard and Operability Study
HPLH	Heavy Lift Pump House
HSE	Health and Safety Executive
IChemE	Institution of Chemical Engineers (Great Britain)
IDLH	Immediately Dangerous to Life and Health
LC	Lethal Concentration
LEL	Lower Explosion Limit
LPG	Liquefied Petroleum Gas
MMTPA	Million Metric Tonnes Per Annum
NFPA	National Fire Protection Association
NUREG	Nuclear Regulatory Commission
ORA	Optimal Risk Analysis
OREDA	Offshore Reliability Data Handbook
PERD	Process Equipment Reliability Data
PRA	Probabilistic Risk Analysis
PRU	Propylene Recovery unit
PSM	Process Safety Management
QRA	Quantitative Risk Analysis
SMS	Safety Management System
SRU	Sulphur Recovery Unit
SWSU	Sour Water Stripping Unit

TLV	Threshold Limit Values
TNO	Netherlands Organization for Applied Scientific Research
TNT	Trinitrotoluene
UEL	Upper Explosion Limit
VCE	Vapour Cloud Explosion
VREP	Visakh Refinery Expansion Project

CHAPTER 1

INTRODUCTION

The growing importance in the area of environment protection is forcing CPI to use the state-of-the-art technology. This in turn increased the complexity of the process. The diversity in the products manufactured by CPI has made it more and more common for these industries to use reactors, conduits and storage vessels in which hazardous substances are handled at extreme conditions. Accidents in such units are caused either by material failure (such as crack in the storage vessels), operational mistakes (such as raising the pressures temperature/flow-rate beyond critical limits), or external perturbation (such as damage caused by a projectile) can have serious-often catastrophic-consequences. The accidental release of the inventory results in toxic release, fires and explosions.

The most gruesome example of such an accident is the Bhopal Gas Tragedy of 1984 which killed or maimed over 20,000 persons but there have been numerous other accidents (Lees, 2005), Flixborough-1974, Basel-1986, Antwerp-1987, Pasadena-1989, Panipat-1993, Mumbai-1995, and Visakhapatnam-1997, in which the death toll would have been as high as in Bhopal if the areas where the accidents took place were not sparsely populated.

Along with the rapid growth of industrialization and population the risk posed by probable accidents also continued to rise. This is particularly so in the third world where population densities are very high around the industrial areas. The growth in the number of such industrial areas and in the number of industries contained in each of the areas gives rise to increasing probabilities of 'chain of accidents' or cascading/domino effects wherein an accident in one industry may cause another accident in a neighbouring industry which in turn may trigger another accident and so on. Some of the past experiences like Mexico-1984, Antwerp-1987, Pasadena-1989 and recently Visakhapatnam-1997 are examples of such disasters.

In order to prevent-or at least reduce the frequency of occurrence of such accidents, major efforts are needed towards raising the level of safety, hazard management and emergency preparedness. This realization and the increased public awareness towards this issue, has prompted the authorities to develop tools and techniques for carrying out risk analysis and assessment of CPIs.

1.1 THE CONCEPT OF RISK AND RISK ASSESSMENT

The concept of Risk and Risk assessment is not a new art. It has been practiced by the insurance markets for at least two centuries as a commercial activity. But long before industrialization, man was familiar with taking and accepting risk. In the context of chemical accident prevention, preparedness and response, codes and standards are the foundations of risk assessment and risk control.

Risk has been defined in various ways by different authors.
IChemE defined it as 'a function of probability (or frequency) and consequence – is sometimes termed 'expected loss.'
The Royal Society, London described it as 'the probability of undesired consequences.'
AIChE/CCPS (2000) coined it as 'a measure of human injury, environmental damage, or economic loss in terms of both the incident likelihood and the magnitude of the loss or injury.'

The risk assessment process, regardless of the method or techniques used includes the following components:

- *hazard:* an inherent property of a substance, agent, source of energy or situation having the potential to cause undesirable consequences (e.g., properties that can cause adverse effects or damage to health, the environment or property),
- *hazard identification:* systematic investigation of the possible hazards associated with an installation, particularly identification of the hazards of the chemical(s) that can cause injury or death to people or damage to property by the release of the chemical or by the release of energy in the event of an accident,
- *events identification:* determination of the different situations that can occur and lead to harm, including estimation of the probable quantity, concentrations, transport, and fate of the hazardous substance(s) or energy released in each specified situation, as determined in part by the environmental conditions at the time of the event,
- *events frequency:* an estimate of the number of times a specified phenomenon (event) occurs within a specified interval,

- *consequence assessment:* a calculation or estimate of the nature and extent of the damage caused by all specified hazardous events, including the influence of environmental factors and the probability of exposure of individuals, populations or ecosystems,
- *risk characterization:* integrates the previous components into an estimation of the combination of the event frequencies and consequences probabilities of the hazardous events specified (for each event and the sum of all events),
- *determining significance:* evaluation of the significance of the risk estimation and each of the components of the risk assessment process, including elements of risk perception and cost/benefit considerations.

The components of the risk assessment process defined above lead to a definition of risk assessment - a value judgment that combines the results of risk characterisation and the estimate of the significance of the risk.

1.2 RISK ACCEPTANCE CRITERIA

Risk acceptance criteria are based on current international practices in certain developed countries. Indian authorities have not yet specified any risk acceptance criteria. Hence Health and Safety Executive (HSE) proposed criterion for Individual Risk is adopted as:
10^{-5} per year for Intolerable risk
Lower than 10^{-6} per year for negligible risk

1.3 QUANTITATIVE RISK ASSESSMENT METHODOLOGY

The concept of QRA, which has emerged in recent years with ever-increasing importance being attached to it, deals with the following key aspects of accidents in CPIs (Khan and Abbasi, 1998).

- Development of tools and techniques to forecast accidents.
- Development of tools and techniques to analyse consequences of likely accidents. Such consequence analysis fulfils two objectives:
    - It helps in siting of industries and management of sites so as to minimize the damage if accident do occur;
    - It provides feedback for other exercises in accident forecasting and disaster management.

- Development of managerial strategies for 'emergency preparedness' and 'damage minimization'.

The risk is defined in QRA as a function of probability or frequency and consequence of a particular accident scenario:

$$\text{Risk} = F(s, c, f)$$

s = hypothetical scenario

c = estimated consequence(s)

f = estimated frequency

The QRA Methodology used in present work is given in Figure 1

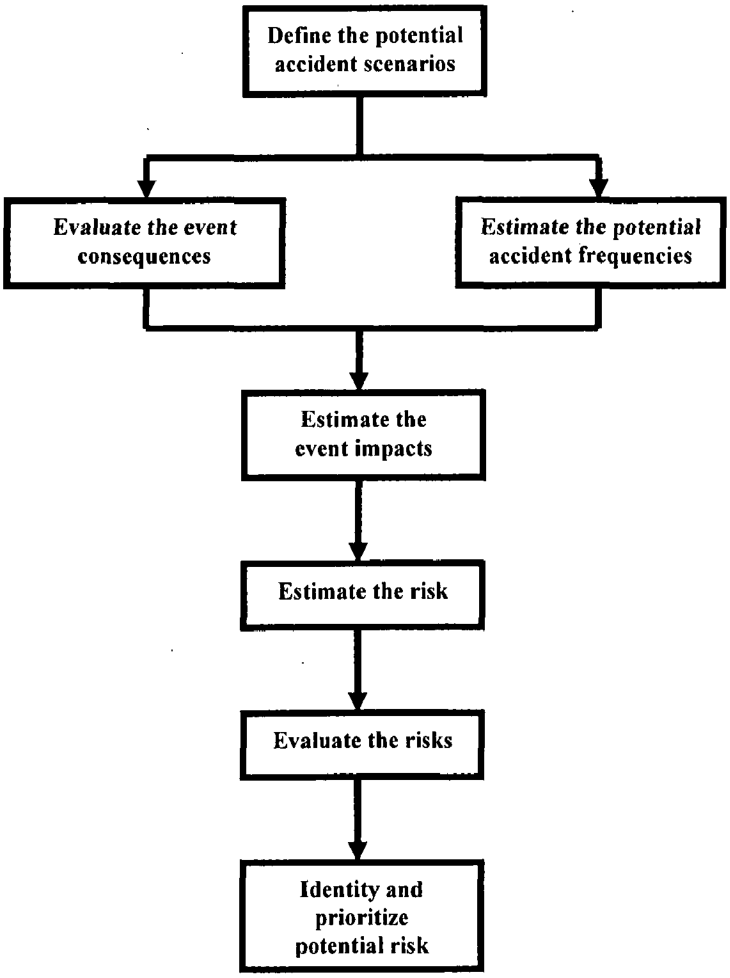


Figure 1.1: QRA Methodology

1.4 AIMS AND OBJECTIVES

From the available information, it was found that the fruits of QRA have not been fully exploited in Indian Refineries. Hence it was thought desirable to study the QRA for Indian Refinery with following objectives:

- To present a comprehensive literature review on risk assessment in chemical process industry.
- To assess the risk posed by refinery facilities with focus on the process units and the inventory using QRA Methodology.
- To suggest risk reduction measures if needed.
- To suggest the emergency preparedness plans.
- To critically examine the results obtained.
- To identify the shortcomings in the methodologies used and suggest improvements and alternatives.

CHAPTER 2

LITERATURE REVIEW

The present chapter discusses the various literatures that were taken to carry out this QRA study. When performing QRA to the large process facilities like petroleum refineries we usually end up with large number of hazards but the careful analysis of these hazards using the guidelines provided in the following literature will enable us to prioritize and eliminate the unwanted hazards. Analysis of major accidents and their causes in the past 20-30 years also provide valuable inputs to make the task less complicated.

Arendt (1990) discussed growing concern about the risk of major chemical accidents. It was pointed out that as new process technologies were developed and deployed, less of the historical experience base remained pertinent to safety assurance giving example of space industry and novel processes in CPI. Focus was on differentiating it from other methods which were deterministic where as QRA for CPI was mainly probabilistic in nature. Further there was a discussion on risk management and risk perspective. A number of misconceptions relating to data acquisition, accuracy etc., were discussed. It was concluded that Quantitative Risk Assessment was an important tool for the CPI and when used judiciously, the advantages of QRA outweighed the associated problems and costs.

Pietersen et al. (1992) presented a brief overview of risk assessment methodologies, focusing on safety in design and the different methods that can be combined for a full QRA. The results of a QRA were normally presented as individual risk and group or societal risk. The calculated individual risks were often presented in the form of an individual risk graph (IRG) or individual risk contours (IRC). The authors calculated the IRGs and IRCs for an example accident scenario: a release of ammonia at 2 kg/s for 1800s, during different types of meteorological conditions. The example clearly illustrated the development of graphs and curves. Finally, it was concluded that a QRA (inclusive risk contour mapping) can be a useful tool, for example in land use planning. However, for the safety of a chemical plant, other methods for risk assessment were also available, and are indispensable.

Tweeddale (1992) retrieved the importance of the 'Rapid ranking' that had been used for several years as a method of determining the priority that should be given to formal investigation of the wide range of hazards and risks present on major process industry sites. They discussed some historical, recent applications and the nature of the developments which they resulted in. Particular topics discussed include: development of the method such that it can be used by operating managers rather than specialists; use to raise risk consciousness; adaptation for use as an aid to hazard identification; use to define both which risks should be studied and which risks, even if low, should be incorporated in routine monitoring and periodic auditing programmes; use as a basis for comparison of the relative risks presented by a variety of different industrial installations; use for a range of different types of risk: and extension beyond ranking to include 'shortlisting' i.e., deciding which risks to include in a study programme and which to leave out. Philosophical difficulties including the danger of omitting a risk, from detailed study, which was erroneously shown as low and the uncertainty of the absolute level of the ranked risks and the approaches to minimize these problems were discussed.

Khan and Abbasi (1998) presented a state-of-the-art review of the available techniques and methodologies for carrying out risk analysis in CPI like checklists, HAZOP, FTA, FMEA and Hazard Indices. They also presented a set of methodologies developed by them to conduct risk analysis effectively and optimally like optHAZOP, PROFAT, HAZDIG etc.

Khan and Abbasi (1999) discussed briefly some of the major accidents in CPI which occurred during 1926–1997. They analysed the case studies with a view to understand the damage potential of various types of accidents, and the common causes or errors which had led to these disasters. An analysis of different types of accidental events such as fire, explosion and toxic release had also been done to assess the damage potential of such events. They stressed the greatest risk of damage posed by VCE. Throughout their study they highlighted the need for risk assessment in CPI.

Ditali et al. (2000) presented the prototype of the computer code, Atlantide, developed to assess the consequences associated with accidental events that can occur in LPG storage/transferring installations. The characteristic of Atlantide was simple enough and at the same time adequate to cope with consequence analysis as required by Italian legislation in fulfilling the Seveso Directive. The models and correlations implemented in the code were relevant to flashing liquid releases, heavy gas dispersion and other typical phenomena such as BLEVE/Fireball. The computer code allows, on the basis of the operating/design characteristics, the study of the relevant accidental events from the evaluation of the release rate (liquid, gaseous and two-phase).in the unit involved, to the analysis of the subsequent evaporation and dispersion, up to the assessment of the final phenomena of fire and explosion.

Mercx et al, (2000) discussed the various models available for the vapour cloud explosion blast modeling. They presented the fundamental objections in applying the TNT Equivalency method for vapour cloud explosion blast modelling. They also discussed the other types of models which do not have the fundamental objections like TNO Multi-Energy method that is increasingly accepted as a more reasonable alternative to be used as a simple and practical method. Computer codes based on computational fluid dynamics (CFD) like AutoReaGas, developed by TNO and Century Dynamics, could be used in case a more rigorous analysis was required. They stressed that a CFD approach, in which the actual situation is modelled, supplies case-specific results. An overview of the key aspects relevant to the application of the Multi-Energy method and CFD modeling were also provided and demonstrated with an example problem involving the calculation of the explosion blast load on a structure at some distance from the explosion in an offshore platform complex.

Khan and Abbasi (2001) discussed a different methodology i.e., Optimum Risk Analysis (ORA). They had gone ahead and applied it on a chemical plant sulfolane manufacturing unit and enumerated the advantages of this technique. ORA aimed to identify and assess hazards and to estimate the risk factors due to any mishap/accident in the chemical process industry. The ORA framework enabled modelling of probable accidents based on the chemical and process characteristics, evaluation of mode of occurrence of these accidents, estimation of detailed consequences and finally prediction of risk factors. ORA had normal steps like risk identification, ranking, estimation and assessment however the way these are done were little bit modified. After assessing the risks to sulfolane unit under consideration the authors made a number of suggestions like instead of one or two large-capacity vessels, several vessels of smaller capacity should be used for storage. Adequate space should be kept between the storage vessels and buffers provided between them so that adverse consequences of failure in one of them do not cause second or higher order. A thorough emergency preparedness strategy should always be kept in position, fortified by periodic drills or 'dry runs' so that the damage is contained in case an accident occurs. According to authors the methodology optimal risk analysis is swift, less expensive to implement, less time-consuming, and is as (or possibly more) accurate and precise, as existing methodologies

Melchers and Feutrill (2001) stated that Quantified risk analysis (QRA) was used for the revision of regulatory separation distances associated with medium size liquefied petroleum gas (LPG) refuelling facilities used in automotive service (gas) stations. The facility consisted of a 7.5 kl pressure vessel, pump, pipework, dispensing equipment and safety equipment. Multi-tank installations were relatively uncommon. They described the hazard scenarios considered, the risk analysis procedure and the selection and application of data for initiating events and for rates of failure of mechanical components and of the pressure vessel. Human errors and intervention possibilities were also considered. Because of the inapplicability of established consequence models and the relatively small scale of the facilities, a number of tests were performed to estimate flame length, flame impingement effects, ignition probabilities and the effectiveness of screening devices.

Tixier et al. (2002) presented an overview of 62 methodologies that has been developed, in the last decade, to undertake risk analysis on an industrial plant. The methodologies were separated into three different phases (identification, evaluation and hierarchisation). In order to understand their running, it seemed necessary to examine the input data, methods used, output data and to rank them in several classes. First, all the input data were grouped together into seven classes (plan or diagram, process and reaction, products, probability and frequency, policy, environment, text, and historical knowledge). Then, the methods were ranked in six classes based on the combination of four usual criteria (qualitative, quantitative, deterministic and probabilistic). And finally, the output data were classified into four classes (management, list, probabilistic and hierarchisation). This classification permits the appraisal of risk analysis methodologies. With the intention of understanding the running of these methodologies, the connections between the three defined previously criteria (determinist, probabilistic and determinist and probabilistic) were brought to the fore. They dealt with the application fields and the main limitations of these methodologies. They also highlighted the difficulties in taking into account all risks for an industrial plant and suggested that there was not only one general method to deal with the problems of industrial risks.

Alonso et al. (2006) recalled the importance of TNO Multi-Energy model in calculating the overpressure and impulse from accidental industrial explosions. From the curves given by this model, data was fitted to obtain equations showing the relationship between overpressure, impulse and distance. These equations, referred herein as *characteristic* curves, can be fitted by means of power equations, which depend on explosion energy and charge strength. *Characteristic* curves allowed the user to determine overpressure and impulse at each distance.

Beerens et al. (2006) once again reiterated the importance of QRA in CPI and, in some countries, also in land-use planning. In QRA calculations the frequency of an accident scenario was most often assessed by a generic failure frequency approach. For this purpose, there was a strong need for failure data that were actual, reliable and generally accepted. They showed the difficulty of tracing back failure data due to the lack of univocal terminology. To obtain validated failure data they initiated a study based on recent data and a fault-tree-based model with algorithms and modification factors. This initiative ultimately lead to failure data that could be easily applied in current-day practice and gives opportunity to obtain appropriate data for non-standard applications. Furthermore, the possibility of how these generic data could be revised and updated was presented.

Chang and Lin (2006) reviewed 242 accidents of storage tanks that occurred in industrial facilities over the last 40 years. They applied Fishbone Diagram to analyze the causes that lead to accidents. The authors also provided corrective actions to help operating engineers handling similar situations in the future. The results show that 74% of accidents occurred in petroleum refineries, oil terminals or storage. Fire and explosion account for 85% of the accidents. There were 80 accidents (33%) caused by lightning and 72 (30%) caused by human errors including poor operations and maintenance. Other causes were equipment failure, sabotage, crack and rupture, leak and line rupture, static electricity, open flames etc. Most of those accidents had been avoided if basic engineering principles were practiced.

Early (2006) discussed the recent implementation of a database management system at a chemical plant and chronicles the improvements accomplished through the introduction of a customized system. According to review while programming techniques still remain a bit complex and cumbersome, they do allow practitioners familiar with the workflow to model using software. This was a major change from the historical systems where the software designer had no knowledge of the workflow requirements.

Moosemiller (2006) described the use of failure rate data but also emphasizes the need to be careful in interpretation. He stated that as companies move progressively toward quantifying the risks of releases of hazardous materials, there becomes a greater need for developing the data necessary to populate the risk analysis. Sophisticated mathematical models have been developed to predict the consequences of a hazardous material release. But the effort devoted to the frequency side of the "risk equation" has been much disorganized by comparison, with inconsistent or non-existent definitions of "failure", mixing of incompatible data, application of data from one industry to a completely different industry, and a host of other problems. Nonetheless, through judicious assembly and analysis of a variety of data sources, a useful failure rate database can be developed. Pitfalls in interpreting failure rate data are also illustrated.

Tasneem Abbasi and S.A. Abbasi (2007) presented an overview of the mechanism, the causes, the consequences, and the preventive strategies associated with BLEVEs. The authors reiterated the importance of understanding the BLEVE mechanism as it is the most devastating of accidents likely in CPI. The list of BLEVE incidents that occurred between 1926 and 2004, the theories behind the BLEVE, various formulae available in the literature to calculate the fireball diameter, height of the fireball from the ground and BLEVE preventive measures are also presented.

Table 2.1: Summary of literature review

S No.	Name of the paper	Authors and Year of Publication	Journal	Conclusions Drawn
1.	Using Quantitative Risk Assessment in the Chemical Process Industry	Arendt, J. S. (1990)	Reliability Engineering and System Safety	Importance of QRA in the CPI. Basic elements of risk assessment and management. Misconceptions about QRA.
2.	Risk assessment and risk contour mapping	Pietersen, C. M., van het Veld, B. F. P. (1992)	Journal of Loss Prevention in the Process Industries	The presentation of the QRA results as Individual risk and Societal risk. Mapping of contours for the individual risk.
3.	Some experiences in hazard identification and risk shortlisting	Tweeddale, H.M., Cameron, R.F., Sylvester, S.S. (1992)	Journal of Loss Prevention in the Process Industries	Importance of risk ranking, shortlisting and the development of a method that can be used by the managers rather than specialists.

S No.	Name of the paper	Authors and Year of Publication	Journal	Conclusions Drawn
4.	Techniques and methodologies for risk analysis in chemical process industries	Khan, F.I., Abbasi, S.A. (1998)	Journal of Loss Prevention in the Process Industries	State-of-the-art review of the available techniques and methodologies for carrying out risk analysis in CPI.
5.	Major accidents in process industries and an analysis of causes and Consequences	Khan, F.I., Abbasi, S.A. (1999)	Journal of Loss Prevention in the Process Industries	Common cause of errors which resulted in the accidents. Greater potentiality of a VCE to result in damage. Need for QRA in CPI is highlighted.
6.	Consequence analysis in LPG installation using an integrated computer package	Ditali, S. et al (2000)	Journal of Hazardous Materials	Prototype of the computer code, Atlantide, developed to assess the consequences associated with accidental events that can occur in a LPG storage/transferring installations.

S No.	Name of the paper	Authors and Year of Publication	Journal	Conclusions Drawn
7.	Developments in vapour cloud explosion blast modeling	Mercx, W.P.M. et al (2000)	Journal of Hazardous Materials	Available models on vapour cloud explosion modeling and the recent developments in the earlier models.
8.	Risk assessment of LPG automotive refuelling facilities	Melchers, R.E., Feutrill, W.R. (2001)	Reliability Engineering and System Safety	Application of the Quantified risk analysis (QRA) to liquefied petroleum gas (LPG) refuelling facilities used in automotive service (gas) stations. for the revision of regulatory separation distances .
9.	Risk analysis of a typical chemical industry using ORA procedure	Khan, F.I., Abbasi, S.A. (2001)	Journal of Loss Prevention in the Process Industries	The need of elaborate safety arrangements to reactor and storage units as they are highly vulnerable to accidents.

S No.	Name of the paper	Authors and Year of Publication	Journal	Conclusions Drawn
10.	Review of 62 risk analysis methodologies of industrial plants	Tixier, J., Dusserre, G., Salvi, O., Gaston, D. (2002)	Journal of Loss Prevention in the Process Industries	The 62 methodologies underline the difficulty in taking into account all risks for an industrial site. Some methods are qualitative and some are quantitative. A single method may not be sufficient to draw a conclusion but the combination of these methods yields better results.
11.	Characteristic overpressure–impulse–distance curves for vapour cloud explosions using the TNO Multi-Energy model	Alonso, F.D. et al (2006)	Journal of Hazardous Materials	Relationship between overpressure- impulse-distance.
12.	Database management systems for process safety	Early W.F (2006)	Journal of Hazardous Materials	Discusses the recent implementation of database management systems at chemical plants

S No.	Name of the paper	Authors and Year of Publication	Journal	Conclusions Drawn
13.	The use of generic failure frequencies in QRA: The quality and use of failure frequencies and how to bring them up-to-date	Beerens, H.I., Post, J.G., Uijt de Haag, P.A.M. (2006)	Journal of Hazardous Materials	Strong need for failure data that are actual, reliable and generally accepted for the QRA studies.
14.	Avoiding pitfalls in assembling an equipment failure rate database for risk assessments	Moosemillar A (2006)	Journal of Hazardous Materials	Importance of databases in quantifying risks, judicious analysis required, description of a number of data sources
15.	A study of storage tank accidents	Chang, J.I., Lin, C.C (2006)	Journal of Loss Prevention in the Process Industries	Various causes for the occurrence of storage tank accidents. The storage of LPG in the sphere form is more dangerous.

S No.	Name of the paper	Authors and Year of Publication	Journal	Conclusions Drawn
16.	The boiling liquid expanding vapour explosion (BLEVE): Mechanism, consequence assessment, management	Tasneem Abbasi, S.A. Abbasi (2007)	Journal of Hazardous Materials	BLEVE mechanisms, consequence assessment and its preventive measures.

2.1 CONCLUSIONS FROM THE LITERATURE REVIEW

Following conclusions were derived from the above literature survey:

- Risk is a subjective concept which varies with respect to the context and the depth of study.
- Guidelines for QRA are available from such sources as AIChE/CCPS, British HSE, and NUREG etc provide a structure to be adopted for risk assessment but leave choice of methods and steps to the assessors.
- Cost benefit analysis is a crucial factor in risk assessment
- Risk assessment can be qualitative as well as quantitative. Quantitative methods are being given more stress since they allow for a better comparison of risk levels and reduce subjectivity in decision making process.
- The major accidents in the history of the petroleum refineries revealed that they originated from the inventories.
- To perform the frequency analysis we can go with plant specific data if available. If not we can trust the generic data from the sources like TNO, PERD, OREDA depending on the situation that matches to the conditions.
- The major problem in QRA is the presence of trustworthy data.
- There is no possibility of eliminating risk but we can put efforts to bring it under the acceptable region.

CHAPTER 3

DEFINITION OF THE PROBLEM

To perform the Quantitative Risk Assessment studies, a petroleum refinery is considered for the analysis. Hindustan Petroleum Corporation Limited (HPCL)-Visakh Refinery (VR) had an initial installed capacity of 0.675 MMTPA in 1957. The crude processing capacity was raised to 7.5 MMTPA throughput level over a period of years by various modifications and adding two units of 3.0 MMTPA capacity of crude oil processing. Refinery was capable of processing both imported & indigenous Crude's. Refinery has processed various types of Bituminous and Non-Bituminous Crude's since its inception. DHDS (Diesel Hydro De Sulphurisation) and related utilities/offsite facilities were added for enhancing the quality of diesel product to meet Environment norms.

The product slate of the refinery included Liquefied Petroleum Gas (LPG), Naphtha, Propylene, Motor Spirit (Petrol), Mineral Turpentine Oil (MTO), Aviation Turbine Fuel (ATF), Jute Bleaching Oil (JBO), Superior Kerosene Oil (SKO), High Speed Diesel (HSD), Light Diesel Oil (LDO), Fuel Oil (FO), Low Sulphur Heavy Stock (LSHS) and Bitumen. Also Sulphur is generated from Sulphur Recovery Unit. The complete configuration of the refinery was presented in Appendix A

The Refinery was located at latitude of 17°41' N and longitude of 83°17" E on an area taken on a 99 years lease from Visakhapatnam Port Trust. The refinery spanning an area of 515 acres was situated at about 1 km (northwest) from the foot of Yarada hills. The refinery was flanked by HPC Terminal & HPC LPG bottling plant on eastern side, Coromandel Fertilizers on the western side, residential colonies on the southern side and Andhra Petrochemicals, East India Petroleum on the northern side. HPCL had additional tankage project (ATP) on the northern side of the refinery covering an area of 215 acres. The Refinery has road access from three sides, form the Southern side (Opposite INS Kalyani hospital), from Eastern side (LLPH gate) and from Northern side (through ATP area).

From the refinery configuration it was clear that the southern side of the refinery was densely populated and was surrounded by other industries. Any accident in this industry may trigger the dominos effects. The refinery also consisted large storage inventories rather than process inventory. Chang and Lin (2006) described that the storage of LPG in

Horton sphere pose greater risk than storage in any other form. The refinery consists of five LPG and two Propylene Horton spheres.

In the present work, Quantitative Risk Assessment study was performed on HPCL-VR with the aims and objectives described earlier.

CHAPTER 4

HAZARDS IDENTIFICATION AND SCENARIO SELECTION

The present chapter deals with the hazardous chemicals handled in the refinery and their threshold values, selection of the release sources from the process and storage facilities using the methodology described in the Appendix B. Finally, the potential release sources are shortlisted.

4.1 INTRODUCTION

A Risk Assessment study starts with the identification of hazards and selection of scenarios that are to be addressed for further analysis.

Hazard (Crowl and Louvar, 2002) was defined as 'a chemical or physical condition that has the potential for causing damage to people, property or the environment.' A number of techniques were available till date for hazard identification depending upon the depth and objective of study.

Accidental release of flammable or toxic vapours can result in severe consequences like Boiling Liquid Expanding Vapour Explosion (BLEVE), Vapour Cloud Explosion (VCE) or Toxic Vapour Cloud. A BLEVE occurs when there is a sudden loss of containment of a pressure vessel containing a superheated liquid or liquefied gas. Delayed ignition of flammable vapours can result in blast overpressures covering large area. Toxic clouds may cover yet larger distances due to the lower concentration threshold value for potentially lethal situations in relation to those in case of explosive clouds.

In contrast, fires have localised consequences. The extent of damage to people depends on the heat flux and duration of exposure. Fires can be put out or contained in most cases, but there are few mitigating actions one can take once a vapour cloud gets released. Major accident hazards arise, therefore, consequent upon the release of flammable or toxic vapours.

This chapter deals with the identification of hazards and scenarios leading to BLEVE, Vapour Cloud Explosion/flash fire and Toxic Vapour Cloud, which potentially can have off-site consequences.

HPCL-VR has several plants and storage areas. Identification of the hazardous materials handled in the various units was the first step. Typically the focus is on identification of those chemicals which on release can get air-borne and have off-site consequences.

4.2 HAZARDOUS CHEMICALS HANDLED AND THEIR PROPERTIES

It is essential to have comprehensive information on all the chemicals handled in the Refinery. An understanding of their physico-chemical properties is also essential. Being a petroleum refinery all units handle various hydrocarbons, liquid or gaseous. These are not pure components but mixtures of several hydrocarbons, especially the liquid hydrocarbon streams that comprise of a large number of hydrocarbons of varying Carbon number. Besides hydrocarbons other hazardous chemicals are also handled in some of the process units of the refinery.

Liquefied Petroleum Gas

LPG poses both fire and explosion hazard. Boiling liquid like LPG (or Butane/Propane) on accidental release would result in significant flash evaporation. On delayed ignition flash fire/vapour cloud explosion would result.

Liquid Hydrocarbons

The liquid hydrocarbons are not pure components but mixtures of a large number of hydrocarbons of varying Carbon number. Therefore, no stream has a definite boiling point but instead each has a boiling range. A particular stream may be defined in terms of the various Carbon number cuts it contains as specified by its initial and final boiling points.

The hazard posed by these hydrocarbons is largely that of fire. Some of the streams with relatively high vapour pressure may possibly generate enough vapour on release to pose an explosion or flash fire hazard e.g. Crude, naphtha (CDU top product), motor spirit. All other liquid hydrocarbon streams in the refinery give rise to only pool fires upon ignition, with the heat radiation hazard. From experience it may be concluded that releases of these hydrocarbons will not give rise to off-site consequences.

Inventories of liquid hydrocarbons in the process facilities are relatively small and the largest inventories are in the storage tanks.

Other Substances

In addition to hydrocarbons, there are other hazardous substances handled in various process facilities of the plant. These are for the most part toxic chemicals, and are listed as follows.

Table 4.1 List of hazardous chemicals and the facility where they are used

Hazardous Substance	Facility where Used
Hydrogen (H_2)	Diesel Hydro Desulphurisation Unit (DHDS)
Hydrogen Sulphide (H_2S), Sulphur dioxide (SO_2)	Sulphur Recovery Unit (SRU)
Chlorine (Cl_2)	Heavy Lift Pump House (HPLH), Drinking water package unit
Carbon Monoxide (CO)	Fluidized Catalytic Cracking Unit (FCCU) I/II
Ammonia (NH_3)	Crude Distillation Unit (CDU)

Hydrogen sulphide is absorbed in amine solution and then recovered from it by regenerating the amine solution. The recovered hydrogen sulphide is sent to the sulphur recovery unit where it is burnt to elemental sulphur. Sulphur dioxide is formed as an intermediate product in this process. There is no storage of either hydrogen sulphide or sulphur dioxide. Chlorine is used in tonners for water treatment in the above locations. In FCCU combustion of residual CO is completed in a CO Boiler where steam is generated. Aqueous ammonical solution (2-5%), by bubbling gaseous ammonia through demineralised water, is used for dosing into atmospheric column overhead line, atmospheric column reflux line and vacuum cool overhead system in CDU.

The properties of these chemicals are briefly described below.

Hydrogen Gas

In case of an accidental release of hydrogen or hydrogen bearing mixture, due to very high buoyancy the cloud will rise very fast and will either get dispersed or an immediate ignition will occur due to reverse Joule-Thomson effect/high operating temperature. There is no storage of hydrogen in the plant premises.

Hydrogen Sulphide

Hydrogen sulphide is a dense, colourless, highly flammable water-soluble gas with an offensive odour of rotten eggs. The Lower Explosion Limit (LEL) is 2% while Upper Explosion Limit (UEL) is 46%. On combustion the gas emits highly toxic fumes of SO_2. H_2S itself is an acutely toxic gas. When the amount of hydrogen sulphide absorbed into the blood stream exceeds that which is readily oxidised, systemic poisoning occurs and respiratory paralysis may follow immediately. This situation corresponds to an atmospheric concentration of approximately 700 to 1000 ppm. The 30-minute LC_{50} value is 987 mg/m^3.

A release of hydrogen sulphide could occur during transfer of the amine acid gas from the amine regenerator to the sulphur recovery unit, sour water stripper gas from Sour Water Stripping unit (SWSU) to SRU and from the combustion chamber of the SRU in DHDS. It poses both toxic as well as explosive vapour cloud hazard.

Sulphur Dioxide

Sulphur dioxide is highly toxic and an irritant. The normal boiling point is -10°C. The vapours cause irritation of the eyes and lungs, with severe choking. Exposure to liquid sulphur dioxide can cause frostbite. The 30-minute LC_{50} value is 5784 mg/m^3. Sulphur dioxide is present as a gas at high temperatures in the combustion chamber of the SRU furnace of DHDS.

Ammonia

Ammonia is highly toxic and an irritant. It is also a flammable material, the limits in air being 16-25% and the ignition temperature is 651°C. However, in normal circumstances it is extremely difficult to ignite the gas unless another flammable material is also present. Therefore only the toxicity is to be considered. The normal boiling point of ammonia is -33.4°C. On contact with the skin ammonia produces severe burns compounded by frostbite. It causes irritation of the mucous membrane, eye surface and any moist skin. The gas tends to prevent respiration on the upper respiratory tract. In high concentrations if inhaled it causes unconsciousness from halted respiration or throat becoming closed at the windpipe. The Immediately Dangerous to Life and Health (IDLH) value of Ammonia is 500 ppm and the 30-minute LC_{50} value is 6164 mg/m^3.

Chlorine

Chlorine is a toxic gas (B.P. is -34°C). In liquefied form chlorine is a clear amber dense liquid. The gas is greenish-yellow, about 2.5 times as dense as air, and non-flammable. Liquid chlorine causes severe irritation and blistering of skin. The gas has a pungent suffocating odour and irritates to the nose and throat. It is an extremely powerful vesicant (blister producing) and respiratory irritant.

Typically exposure to chlorine concentrations of 3-6 ppm results in a stinging and burning sensation in the eyes. Exposure for 0.5-1 hour to concentration ranging between 14-21 ppm causes pulmonary oedema, pneumonitis, emphysema and bronchitis. This is usually associated with marked bronchospasm, muscular soreness and headache. Whilst there is inevitably a variation in individual susceptibility, typically 4 ppm is the maximum concentration that can be breathed for one hour without damage, 40-60 ppm is dangerous for a 30-minute exposure and a concentration of 1000 ppm is likely to be fatal after a few breaths.

Chlorine tonners were in use at various locations in HPCL-VR, which represents a toxic vapour cloud hazard.

4.3 METHODOLOGY FOLLOWED FOR SELECTION OF RELEASE SOURCES

Selection of release sources is based on IPO (Inter Provinciaal Overleg, a governmental body in The Hague, Netherlands) guidelines (KLG-TNO, 2000) currently in use in the EC countries. Most of the applied directives in European countries and the USA were based on a selection by use of the inventory pressure, temperature and the threshold value for that specific compound or class to which the compound belongs. The threshold values were chosen in such a way that inventories below the threshold will not form a major hazard with off-site consequences.

Detailed methodology has been discussed in Annexure B.

In accordance with the methodology the first step was the shortlisting of units for S-factor calculation. For this purpose it was convenient to consider the refinery facilities as being in one of three groups' viz., Process Units, Storage, and Run down lines from process and

Transfer Pipelines to/from Refinery. For the identified release sources, A-factor and S-factor was calculated with the help of spreadsheet software provided by the HPCL-VR and finally the list of hazards is tabulated as follows.

Selection of release sources according to the described methodology is given in the following section.

4.3.1 Selection of release sources from process units with potential off-site consequences

From the literature review (Chang and Lin, 2006), it is clear that off-site consequences from release of hydrocarbons may result only from releases of LPG or the relatively more volatile liquid hydrocarbons which are potentially capable of generating explosive vapour cloud. Process inventories are relatively small when compared to product storage. Only those process inventories, which may result potentially in explosive vapour cloud, have been considered. Therefore the largest inventory (>10 MT) of volatile liquid hydrocarbons from each process unit are listed. No process hydrocarbons inventories from the existing process units have S factor even equal to one and these process units do not contribute to any significant off-site risk. Among flammables, in addition to hydrocarbons, hydrogen inventories are also present but these were clearly too small in relation to their location from their refinery boundary to merit consideration.

Toxic chemicals handled in the process units are ammonia, carbon monoxide, chlorine, hydrogen sulphide, and sulphur dioxide. Barring chlorine, which is handled in ton cylinders, the other inventories are present as "flow inventories" rather than as hold-ups in some vessel or container. All these inventories were considered for S factor calculation for the sake of simplicity in short-listing. For the estimation of the total quantity of toxic gas, which may result potentially in toxic vapour cloud, e.g. H_2S from Amine Regenerator to SRU and CO from CO Boiler, regeneration gas line; maximum release duration has been assumed of 20 minutes; closure of the respective valves or stopping of the leak is assumed to take place within that time.

Table 4.2 Calculated A and S factors for the Process units

Location	Material	Release Source	A-factor	S-factor
CDU I	Primary Naphtha	Atmospheric overhead drum	1.0	0
CDU II	Primary Naphtha	Atmospheric overhead drum	2.1	0
CDU III	Primary Naphtha	Atmospheric overhead drum	2.4	0
DHDS	H_2S	Acid gas line from ARU to SRU	26.7	>1
DHDS	H_2S	Off gas from SWSU to SRU	0.53	0
DHDS	NH_3	Gas from SWSU to SRU	0.3	0
FCCU I	CO	CO boiler	1.4	0
FCCU II	CO	CO boiler	2.88	0
HPLH	Cl_2	Chlorine tonner	30	>1
SRU	H_2S	Combustion Chamber	20	>1

It is observed that all release sources from process units except Chlorine tonner at different locations, H_2S lines from ARU to SRU and SRU combustion chamber result in S-factor less than 1. These release sources are selected for Hazard Analysis.

4.3.2 Selection of Storage units with potential off-site consequences

Hydrocarbon products were stored in the respective storage tanks in the tank farm area including ATP within the refinery battery limit. There were about 67 tanks within HPCL refinery premises, while the ATP has 26. Additional 14 tanks were being provided by VREP - II and DHDS projects. A mix of floating roof and fixed roof tanks were in use.

Table 4.3 Calculated A and S factors for the Storage units

Location	Material	tank number	Capacity (kl)	A-factor	S-factor
Tank farm	Crude oil	20-D-5	63,622	59	0
Tank farm	Crude oil	120-T-01E	65,597	59	>1
Tank farm	Naphtha	20-D-160	12,844	10.02	0
Tank farm	Naphtha	120-T-7A	35,430	27.6	>1
Tank farm	LPG	120-T-12A	1200 MT	120	>1
Tank farm	Propylene	120-T-180A	750 MT	75	>1

4.4 LISTING OF RELEASE SOURCES

The following were the list of release sources for which S factor = or >1 and were the major hazards that have the offsite consequences. The listing is not complete i.e. if there were two or more tanks of the same capacity for one material at one location, both were potential release sources if S-factor = or >1 for the listed unit.

Table 4.4 Release sources for Process and Storage units

Location	**Material**	**Release source**
Process Units		
DHDS	H_2S	Acid gas line from ARU to SRU
VREP-II	H_2S	Acid gas line from ARU to SRU
HPLH	Cl_2	Chlorine tonner
Storage Units		
Tank farm	Crude oil	Storage tank/120-T-01E
Tank farm	Naphtha	Storage tank/120-T-7A
Tank farm	LPG	Horton sphere/120-T-12A
Tank farm	Propylene	Horton sphere/120-T-180A

CHAPTER 5

CONSEQUENCE ANALYSIS

The present chapter deals with the consequences that result from the potential hazards that were indentified in Chapter 4 and the failure modes of the release sources.

5.1 INTRODUCTION

Accidents begin with an incident, which usually results in the loss of containment of material from the process. The material has hazardous properties, which might include toxic properties and energy content. Typical incidents might include the rupture or break of a pipeline, a hole in a tank or pipe, runaway reaction, fire external to the vessel, etc. The subsequent effects upon release depend on large number of factors e.g. type and quantity of material released, proximity to the ignition source, meteorological conditions etc. Once the incident was known, source models were selected to describe how materials are discharged from the process. The source model provided a description of the rate of discharge, the total quantity discharged (or total time of discharge), and the state of the discharge-solid, liquid, vapour, or a combination. A dispersion model was subsequently used to describe how the material was transported downwind and dispersed to some concentration levels. For flammable releases, fire and explosion models converted the source model information on the release into energy hazard potentials such as thermal radiation and explosion overpressures. Effect models converted the incident-specific results into effects on people (injury or death) and structures.

In a brief, these were the outcomes from this chapter

Table 5.1 Outcome from consequence analysis based on incident

Incident	**Outcome**
Fire	Radiant heat flux
Explosion	Overpressure
Toxic release	Toxic effects

5.2 FAILURE MODES FOR SELECTED RELEASE SOURCES

Plant inventory releases to environment due to Loss of Containment. Various causes and modes for such an eventuality were well documented in AIChE/CCPS (2000). A leak can range in size from a pinhole leak to a catastrophic failure. In general smaller leaks have higher accident likelihood but lower consequence distances. On the other hand larger releases have lower accident likelihood but longer consequence distance.

For all the release sources listed above release consequence calculations have been done assuming a catastrophic failure. For vessels, including storage tanks, a catastrophic failure is defined as failure of the largest diameter nozzle connection to that vessel. This establishes the worst case consequences from loss of containment of that inventory.

For QRA, we need to consider likely failure modes that could result in off-site consequences. Often, two equivalent hole sizes are considered for a given failure mode (particularly piping failure). One hole size represents a small (more likely) leak and the other a large (less likely) leak, for example 20% and 100% of line diameter for piping failure. From the literature (Chang and Lin, 2006; KLG-TNO, 2000) on inventories of flammable chemicals viz., LPG, we find that only large leaks are significant for off-site consequences. Therefore, in the present work, only a large leak size had been considered for any particular failure mode. For the same reason failure modes such as flange leak, valve leaks etc., which are most unlikely to cause large leaks have not been considered for flammable inventories. To summarise, the following equivalent hole sizes and modes have been considered for QRA scenarios

1. Catastrophic failure of vessel (failure of largest diameter nozzle connection)
2. Failure of bottom line:
    a. Full bore rupture for piping dia. < 6"
    b. 6" equivalent hole size for piping dia. > 6"
    c. For toxic inventories, a small leak with equivalent hole size = 20% of piping dia.

5.3 LISTING OF SCENARIOS FOR RELEASE CONSEQUENCE CALCULATIONS

From the hazards that were listed in Table 4.4, the consequence analysis was performed based on the following failure mode and equivalent hole size and outcome scenario.

Table 5.2 Failure mode and outcome scenario for the release sources

S.No	Release Source	Failure Mode and Equivalent Hole Size	Outcome Scenarios
1.	DHDS/ H_2S line from Amine regenerator to SRU	Full bore rupture	Toxic vapour cloud dispersion
2.	Naphtha storage/120-T-7A	Catastrophic failure, 6"	Explosive vapour cloud formation with delayed ignition
3.	LPG Horton sphere/120-T-12A/9A/B/C/D	Catastrophic failure	BLEVE
4.	Propylene Horton sphere/120-T-180A/B	Catastrophic failure	BLEVE

The above listed scenarios were modelled based on the incident outcome and their effects were presented in the chapter 8

CHAPTER 6

FREQUENCY ANALYSIS

This chapter finds the frequency of occurrence of the events that were considered for QRA in Chapter 5. The frequency of potential risk contributors were calculated using the fault tree analysis where as the other risk contributors were excerpted from the generic data available in the literature.

6.1 INTRODUCTION

Failure frequencies may be classified as plant specific data and generic data. Failure rate data generated from collecting information on equipment failure experience at a plant are referred to as plant-specific data. A characteristic of plant-specific data was that they reflect the plant's process, environment, maintenance practices, and choice and operation of equipment. Data accumulated and aggregated from a variety of plants and industries, such as nuclear power plants, CPI or offshore petroleum platforms, and are called generic data. With inputs from many sources, generic failure rate data can provide a much larger pool of data. However, generic data are derived from equipment of many manufacturers, a number of processes, and many plants with various operating strategies. Consequently, they are much less specific and detailed.

The generic data can be obtained from the sources like PERD, OREDA, TNO. The frequencies were given in **frequency per year**, except for the pipes they were given in **frequency per year and per meter.**

6.2 ACCIDENT LIKELIHOOD CALCULATIONS FOR QRA SCENARIOS

Scenario 1: Failure of H_2S line from DHDS Amine Regenerator to DHDS Sulphur Recovery Unit

Outcome: Toxic vapour cloud of H_2S

Scenario Description

A major leak in the pipeline will result in a release of H_2S to the atmosphere. A toxic cloud will be formed which will disperse in the prevailing wind direction. Release duration was assumed to be ten minutes as it was assumed that actions to stop the release will be initiated within ten minutes of the release taking place.

Accident likelihood

Piping failure frequency: 1.8E-6 m^{-1}yr^{-1}
Line length: 50 m
Outcome likelihood: 9.0E-5 yr^{-1}

Scenario 2: Failure of Naphtha Tank

Outcome: Vapour Cloud Explosion / Flash fire

Scenario Description

Catastrophic failure (largest nozzle dia.-6") results in release of naphtha, which will be confined in a pool. There was no possibility of isolation and release duration is taken as 5 minutes. On release, a pool results that on evaporation a vapour cloud will be formed which may ignite resulting in a flash fire or VCE.

Accident likelihood

Storage tank catastrophic failure frequency: 1.0E-6 yr^{-1}
No. of Naphtha tanks: 2
Probability of delayed ignition: 0.04 (Appendix D.3.5)
Outcome likelihood: 8.0E-8 yr^{-1}

Scenario 3: BLEVE of LPG Horton sphere

Outcome: BLEVE

Scenario Description

LPG from Crude Distillation Unit (CDU) and Fluidized Catalytic Cracking Unit (FCCU) is Amine treated and caustic washed in the Merox plant and routed to storage spheres in the plant. There are 5 spheres of similar capacity in the same geographical location.

LPG was stored under pressure and catastrophic failure leading to BLEVE was a possibility.

Accident likelihood

Horton sphere catastrophic failure frequency: 3.6E-7 yr^{-1} (Figure 6.1)
No. of Naphtha tanks: 5
Outcome likelihood: 1.80E-6 yr^{-1}

Scenario 4: BLEVE of Propylene Horton sphere

Outcome: BLEVE

Scenario Description

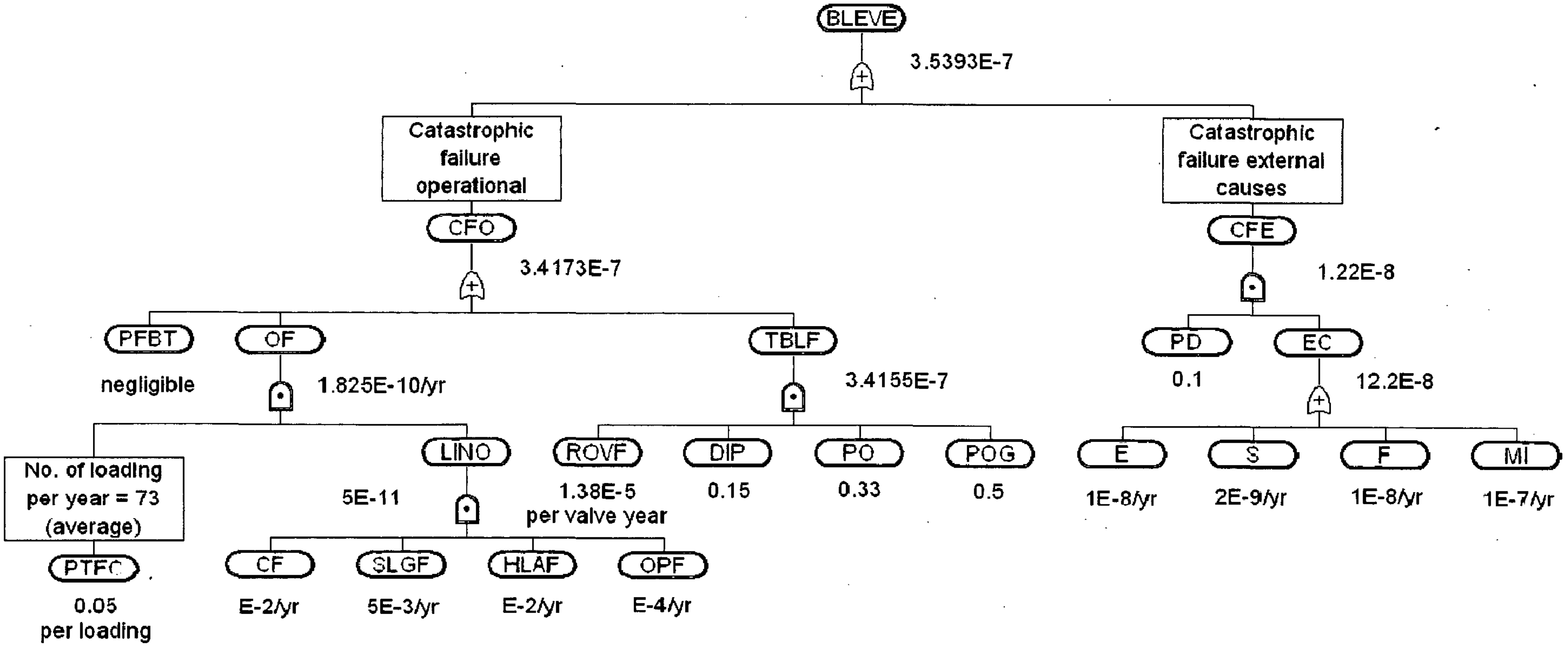
Propylene from cracked LPG is a mixture if propane, propylene, butane, butylenes with some amounts of C2 and C5 hydrocarbons. Cracked LPG from Merox plant will be routed to the Propylene Recovery unit (PRU), where it is segregated into propylene and saturated LPG. Saturated LPG is routed to LPG storage spheres. Propylene is purified, water washed and routed to 2 nos. Propylene Horton spheres for storing and shipping. Propylene is stored under pressure and catastrophic failure leading to BLEVE is a possibility.

Accident likelihood

Horton sphere catastrophic failure frequency:	3.6E-7 yr^{-1}
No. of Naphtha tanks:	2
Outcome likelihood:	7.2E-7 yr^{-1}

Table 6.1 Summary of Frequency Analysis

#	Scenario Description	Accident Frequency
1	BLEVE of LPG Horton sphere	1.8E-6 /yr
2	BLEVE of Propylene Horton sphere	7.2E-7 /yr
3	Failure of Naphtha Tank	8.0E-8 /yr
4	Failure of H_2S line from DHDS Amine Regenerator to DHDS Sulphur Recovery Unit	9.0E-5 /yr



CFO: Catastrophic failure (operational causes)
CFE: Catastrophic failure (external causes)
PFBT: Prolonged fire below tank
OF: Over filling
TBLF: Tank/bottom line failure due to blast overpressure
PD. Probability of major damage
EC: External causes
LINO. Level increase not observed
ROVF: Remote operating valve failure
DIP: Delayed ignition probability
PO: Probability of over pressure
POG: Probability of over pressure greater than 0.5 bar
E: Earthquake
S: Subsidence
F: Flooding
MI: Mechanical impact
PTFC: Probability of tank filling to full capacity
CF: Cell failure
SLGF: Servo level gauge fails
HLAF: Head-Head level alarm fails
OF: Operator failure

Figure 6.1: Fault Tree of Horton sphere

Note: **i.** Horton spheres were generally not filled to more than 80% of the full filling capacity. Nevertheless, loading to full capacity cannot be ruled out. This probability had been assumed as 5% i.e., 5 out of 100 loadings reach the full capacity of the sphere.

ii. It was assumed that no damage to the sphere, its supporting structure, bottom line and upstream of the ROV will occur below the blast overpressures of 0.5 bar.

CHAPTER 7

RISK MAPPING

This Chapter deals with the presentation of the risk in terms of contours i.e., individual risk contours and the societal risk contours.

7.1 INTRODUCTION

The results of a QRA are most often presented in terms of individual and group or societal risk. Risk was defined in Crowl and Louvar, 2001as a measure of human injury, environmental damage, or economic loss in terms of both the incident likelihood and the magnitude of the loss or injury. Risk thus comprises of two variables: magnitude of consequences and the probability of occurrence. Risk concepts and Risk acceptance criteria were discussed in chapter 1

Individual Risk is the probability of death occurring as a result of accidents at a plant, installation or a transport route expressed as a function of the distance from such an activity. Such a risk actually exists only when a person is permanently at that spot (out of doors). The individual risk was illustrated with the aid of risk-curves or iso-risk contours.

Group risk (Societal risk) is the probability of a certain number of victims per year In calculating the group risk demographic data relating to the presence of humans is necessary. The societal risk was represented as an F-N curve, which depicts the frequency of occurrence per year F of a certain number of fatalities, N.

Another measure of societal risk, viz. Potential Loss of Life (PLL) is also computed. PLL is expressed in terms of expected number of fatalities per year. PLL enables a single number comparison of alternate situations. Its drawback is that it does not permit discrimination between more likely accidents with fewer numbers of fatalities and less likely accidents with greater numbers of fatalities.

The individual and societal risks from a plant are the result of the cumulating of risks connected with all possible scenarios. From the standpoint of what constitutes acceptable risk levels from a complex, both the Individual as well as Societal risk should be within the acceptance criteria.

7.2. INDIVIDUAL RISK

The individual risk contours were plotted on the plant layout. If the outcome was BLEVE then a circle with diameter equivalent to the fireball radius gave the contour. If the outcome was VCE or toxic release then the circle with diameter equivalent to the maximum puff/plume width and distance travelled by the puff/plume in the estimated time as centre gave the contour. The individual risk for LPG storage was shown is figure 7.1.

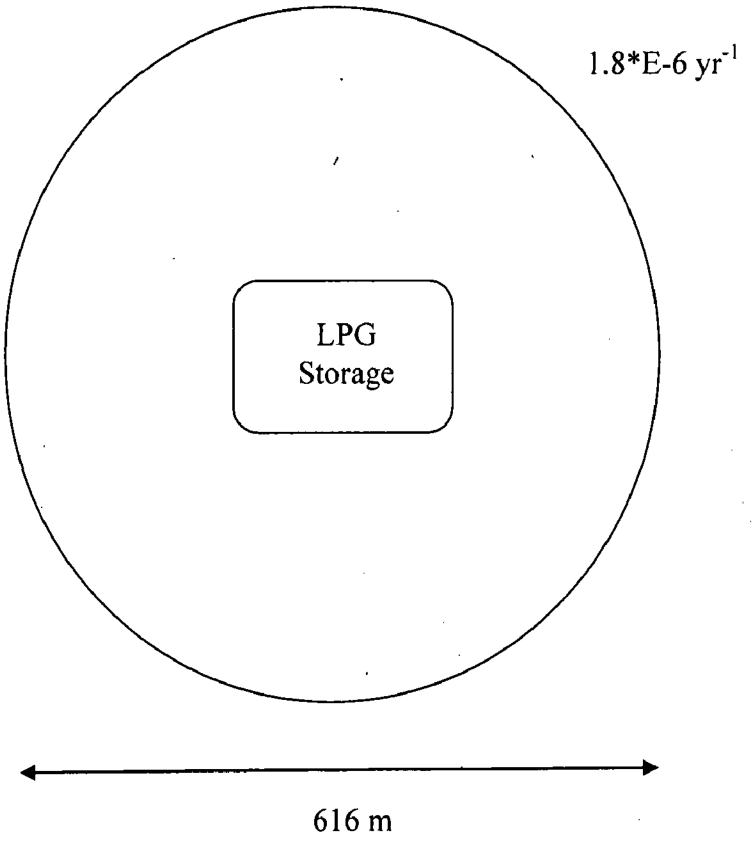


Figure 7.1 Individual Risk Contour of LPG BLEVE

7.3. SOCIETAL RISK

The individual risk values for each location when combined with population at that location gave the group risk or societal risk. The societal risk is represented as F-N curve.

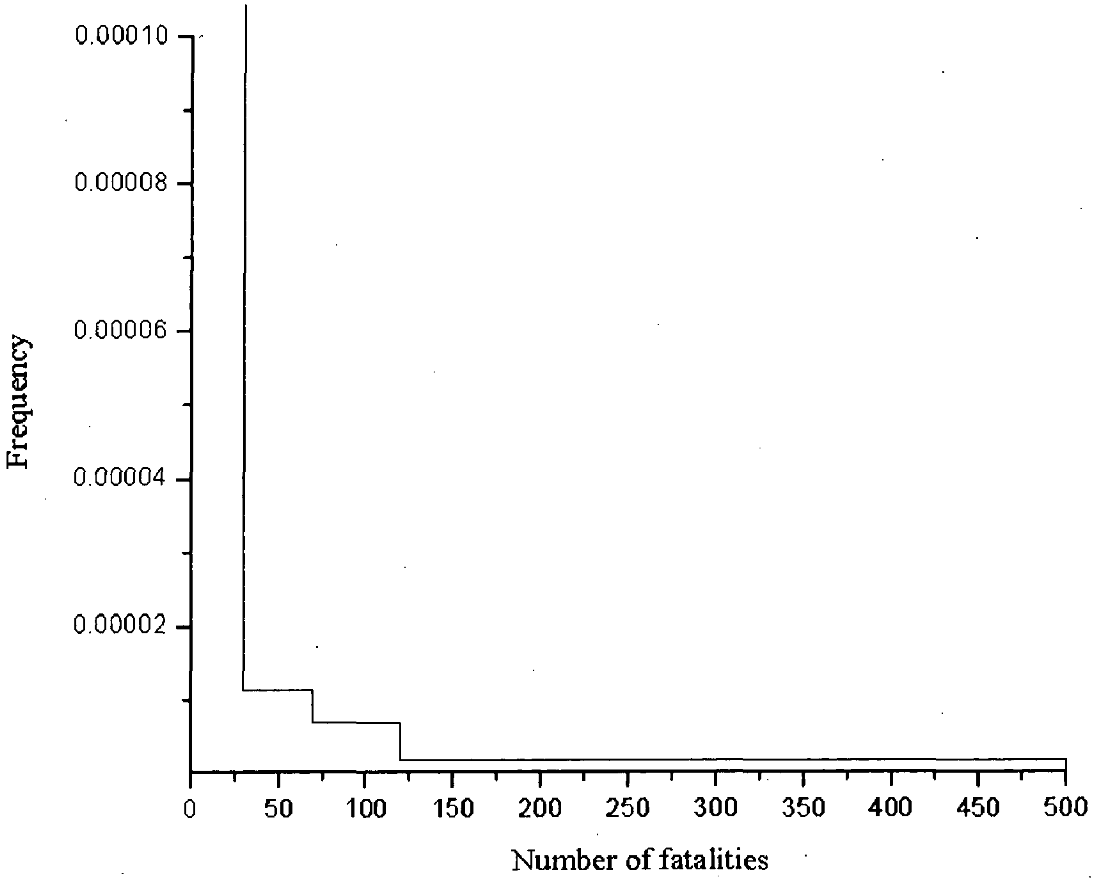


Figure 7.2 F-N Curve for Societal risk

CHAPTER 8

RESULTS AND DISCUSSION

This chapter describes the results of QRA by using the models quoted in the literature.

8.1 RESULTS FROM CONSEQUENCE ANALYSIS

Scenario 1: Failure of H_2S line from DHDS Amine Regenerator to DHDS Sulphur Recovery Unit

H_2S recovered in the Amine regeneration unit will be transported by pipeline to the SRU. A major leak in the pipeline will result in a release of H_2S to the atmosphere. A toxic cloud will be formed and will disperse in the prevailing wind direction. Release duration was estimated to be 10 min as it was assumed that actions to stop the release will be initiated within 10 min of the release taking place.

H_2S was also flammable. However, the release rate was so low that no explosive cloud of 50 kg or more in size was unlikely to form. Therefore, dispersion of flammable cloud was not modelled.

The dispersion of the toxic cloud was modelled using the Pasquill-Gifford puff model (AIChE/CCPS, 1999) with the following input details.

Input Details:

Dia. of the line:	10"
Length of the line:	50m
H_2S feed flow rate from ARU:	4365 kg/hr (1.2 kg/s)
Detection time and isolation time:	10 min
Average flow rate:	1.2 kg/s
Duration of release:	600 s
Total mass released:	360 kg
Wind speed:	2 m/s (stability class F)

Calculated Results:

Distance downwind:	1200 m
Isopleth concentration:	13.91 kg/m^3

Maximum puff width: 92.78 m

From the calculated results it was clear that the vapour cloud will be concentrated at 1200m in the downwind direction, which was a densely populated area. As the wind speed was estimated at 2 m/s, it takes near about one minute time for the cloud to pass that location.

The following figure shows the isopleth at a distance downwind.

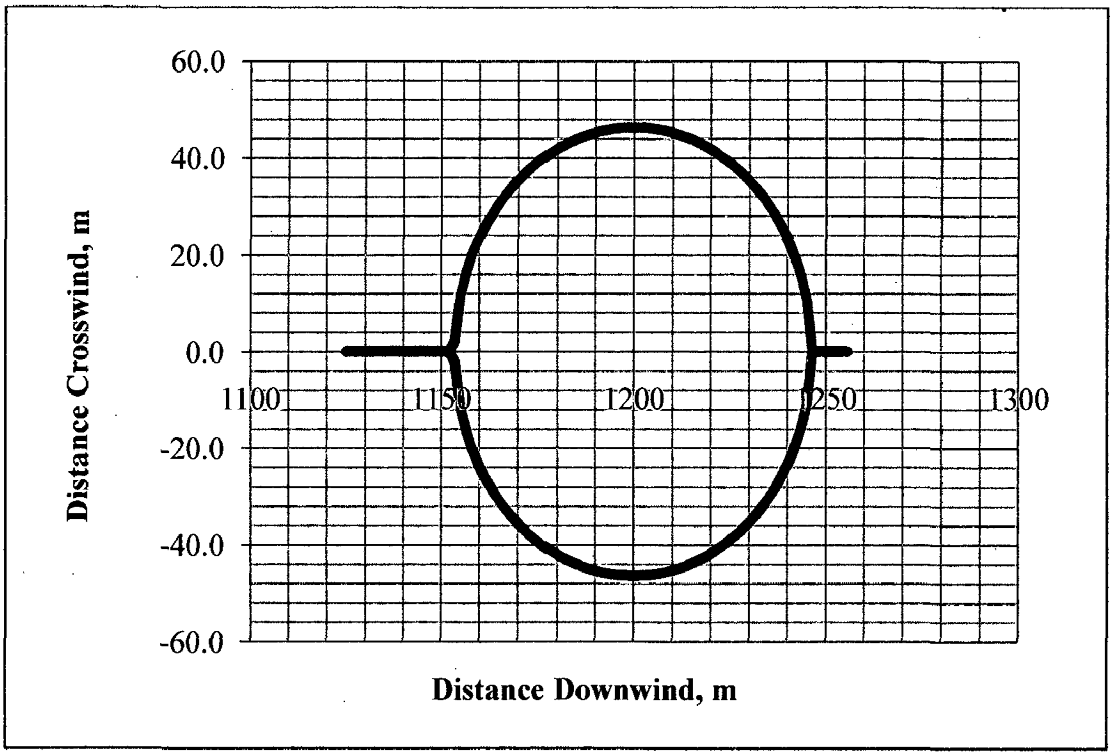


Figure 8.1 Isopleth according to Pasquill Gifford Model (10 ppm)

Scenario 2: Failure of Naphtha Tank

The catastrophic failure (largest nozzle dia.-6") results in release of naphtha, which will be confined in a pool. There is no possibility of isolation and release duration was taken as 5 minutes. On release a pool results that on evaporation a vapour cloud will be formed which may ignite resulting in a flash fire or VCE.

From the following input details the VCE behaviour was modelled using TNT equivalency model (AIChE/CCPS, 1999) to estimate the overpressure generated, duration time and arrival time to a particular distance.

Input details:

Tank Capacity:	35430 kl
Tank Diameter:	58 m
Leak Size:	6"
Average flow rate:	131 kg/s
Duration of release:	300s

Calculate results:

Total mass released:	131*1200 = 39300 kg
TNT equivalent of the fuel:	19683.8 kg TNT
Scaled distance, z:	3.7037 m/kg$^{1/3}$

Distance R(m)	Scaled distance z (m/kg$^{1/3}$)	Overpressure		Impulse (Pas)	Duration time (ms)	Arrival Time (ms)
		(kPa)	(psig)			
100	3.7037	75.33	10.93	77.14	3.28	5.09
500	18.5183	6.68	0.969	17.09	5.81	45.61
1000	37.0365	2.63	0.382	8.57	6.98	99.12
1500	55.5548	1.81	0.262	5.78	8.15	152.82

From the above overpressure results, the damage effects caused to the common structures and the process equipment were estimated.

From Table D.4a, b it was clear that

Distance, m	Damage
100	Probable total destruction of buildings; heavy machine tools (7000 lb) moved and badly damaged; very heavy machine tools (12,000 lb) survive
500	Partial demolition of houses, made uninhabitable; reactor units over turned and pipes were broken
1000	Limited minor structural damage; Storage tanks gets uplifted (half tilted)
1500	Typical pressure for glass breakage; Fired heater bricks crack; Chemical reactors windows and gauges broken

The calculated results overpressure, impulse and arrival time were plotted with respect to distance in the following figures 8.1, 8.2, 8.3. They clearly shoe that the impact in the 500m radius will be more.

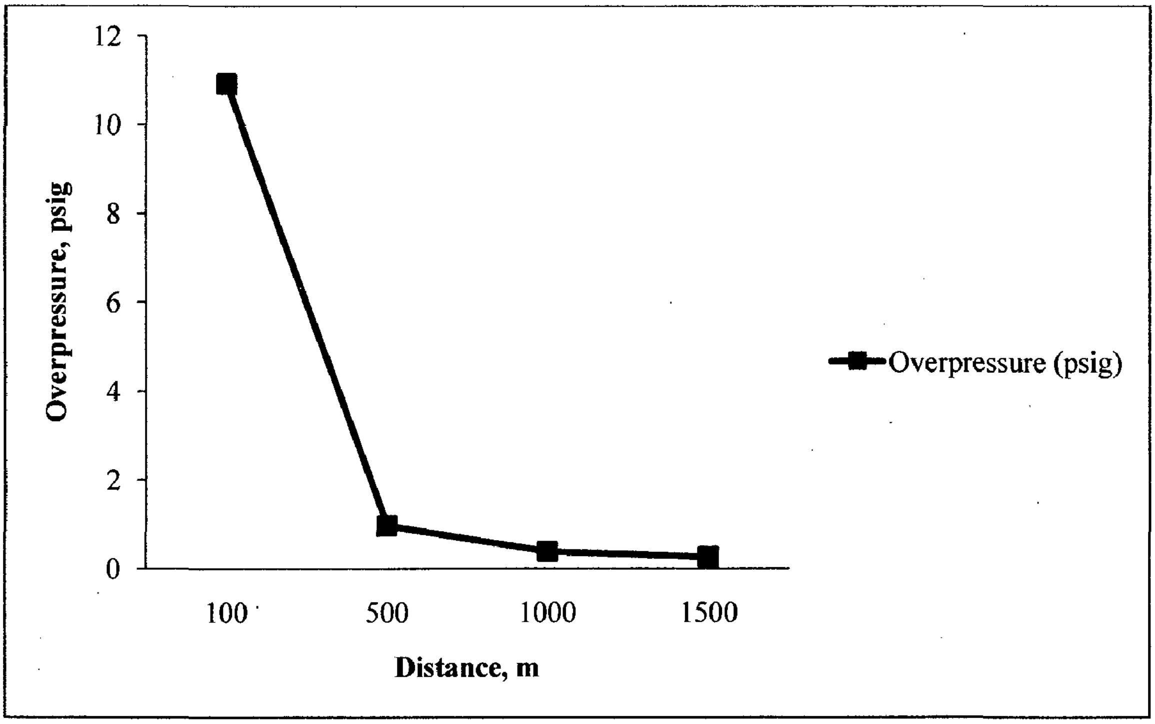


Figure 8.2 Overpressure versus distance

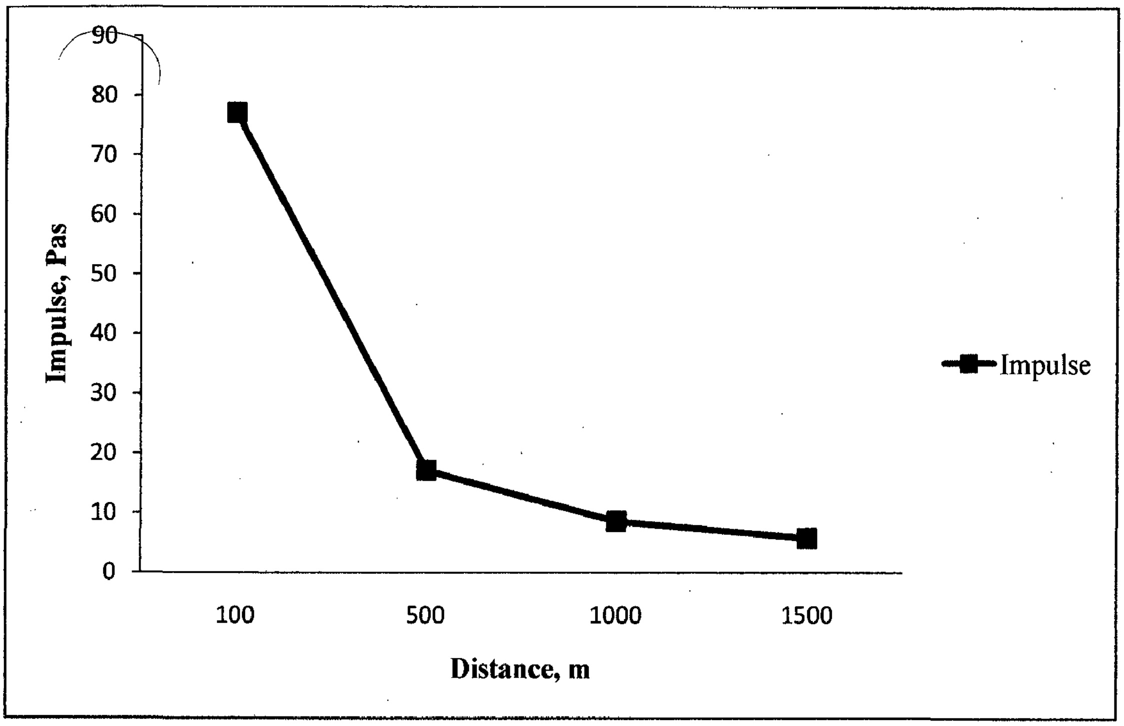


Figure 8.3 Impulse versus Distance

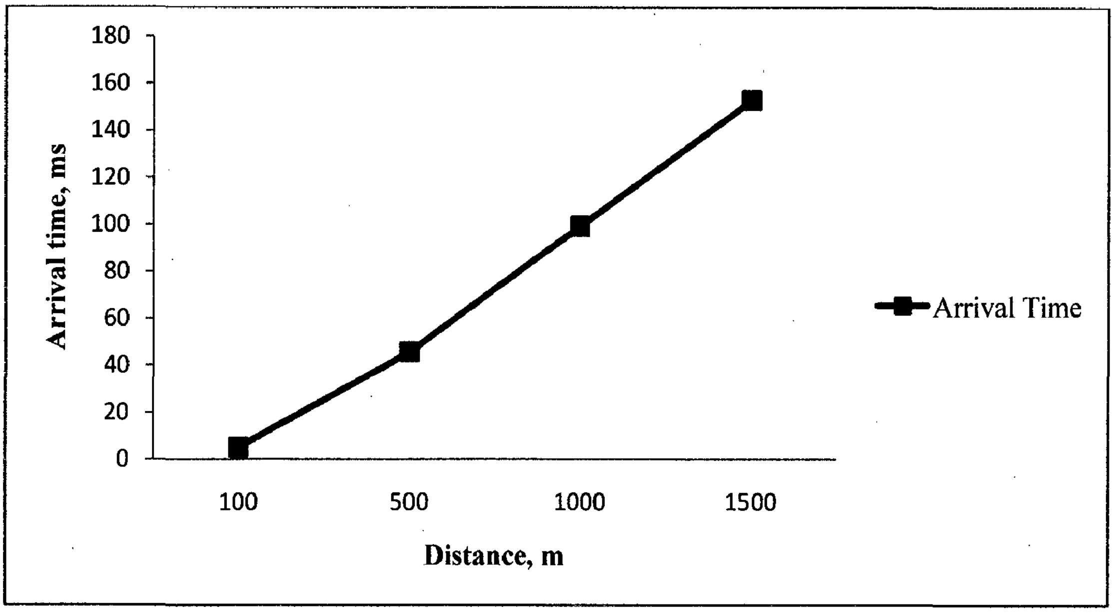


Figure 8.4 Arrival time versus Distance

Scenario 3: Failure of LPG Horton sphere

LPG from Crude Distillation Unit (CDU) and Fluidized Catalytic Cracking Unit (FCCU) was Amine treated and caustic washed in the Merox plant and routed to storage spheres in the plant. There were 5 spheres of similar capacity in the same geographical location.

LPG was stored under pressure and catastrophic failure leading to BLEVE was a possibility. The BLEVE was modelled based on the methodology described in AIChE/CCPS (1999). Based on the following input details the fireball diameter, maximum duration of combustion of the fireball, surface emitted flux and flux at required distance in the horizontal and vertical direction were estimated.

Input Details:

No. of LPG Spheres: 5 (120-T-09A, 09B, 09C, 09D, 12A)
Capacity of each sphere: 1200 MT
Volume of each sphere: 2570 m^3
Diameter of each sphere: 17 m
Thickness of shell plate: 34.5 mm
Mass released: 1200000 kg
Ambient temperature: 300 K
Operating Volume: 80%
Radiation Fraction, R: 0.3
Heat of Combustion of fuel: 45840 kJ/kg
Distance from fireball center on ground, L:

Calculated Results:

Maximum fireball diameter: 616.3 m
Fireball combustion duration: 26.8 s
Center height of fireball: 462.3 m
Initial ground level hemisphere dia.: 801.2 m
Surface emitted flux: 516 kW/m^2

Distance L (m)	Path length X_s (m)	Transmittivity τ_a	View factor		Received Flux (kW/m^2)	
			Horizontal	Vertical	Horizontal	Vertical
200	195.5	0.615	0.34	0.15	109.04	60.18
400	303.1	0.591	0.19	0.17	58.63	50.74
600	449.2	0.570	0.10	0.13	29.75	38.62
800	615.8	0.555	0.06	0.10	15.93	27.57
1000	793.5	0.542	0.03	0.07	9.19	19.87

From the above results and using Appendix D.3 it was concluded that

At a distance above 600m region 1% lethality was observed and in the region of 400-600m the lethality increases based on the increased exposure time.

The received heat fluxes in the horizontal and in the vertical direction were plotted with respect to distance from the event occurrence and were shown in figure 8.5. From the figure it is clear that the horizontal flux is greater than the vertical flux.

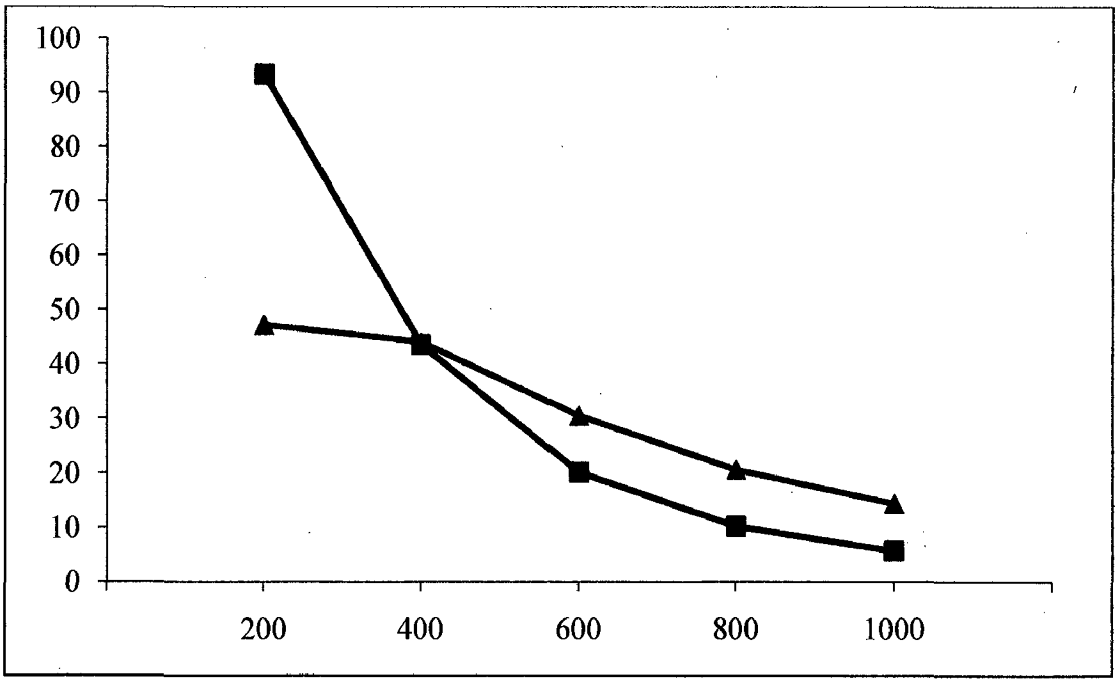


Figure 8.5 Distance vs. Radiant Heat Flux for LPG Horton sphere

Scenario 4: Failure of Propylene Horton sphere

Propylene from cracked LPG was a mixture if propane, propylene, butane, butylenes with some amounts of C2 and C5 hydrocarbons. Cracked LPG from Merox plant will be routed to the Propylene Recovery unit (PRU), where it was segregated into propylene and saturated LPG. Saturated LPG was routed to LPG storage spheres. Propylene was purified, water washed and routed to 2 nos. Propylene Horton spheres for storing and shipping.

Propylene was stored under pressure and catastrophic failure leading to BLEVE was a possibility.

Input Details:

No. of Propylene Spheres:	2 (120-T-180A, 180B)
Capacity of each sphere:	750 MT
Volume of each sphere:	1600 m^3
Diameter of each sphere:	14.5 m
Thickness of shell plate:	45.0 mm
Mass released:	750000 Kg
Ambient temperature:	300 K
Operating Volume:	80%
Radiation Fraction, R:	0.3
Heat of Combustion of fuel:	45799 kJ/kg
Distance from fireball center on ground, L:	

Calculated Results:

Maximum fireball diameter:	527 m
Fireball combustion duration:	24.8 s
Center height of fireball:	395.2 m
Initial ground level hemisphere dia.:	685.1 m
Surface emitted flux:	476.9 kW/m^2

The results were obtained in the same manner as described in the above scenario 3.

Distance L (m)	Path length X_s (m)	Transmittivity τ_a	View factor		Received Flux (kW/m^2)	
			Horizontal	Vertical	Horizontal	Vertical
200	179.5	0.620	0.32	0.16	93.28	47.2
400	298.8	0.592	0.15	0.16	43.55	44.07
600	455.0	0.570	0.07	0.11	20.10	30.52
800	628.8	0.553	0.04	0.08	10.19	20.63
1000	811.8	0.541	0.02	0.06	5.69	14.40

It was concluded that the distance in the radius of 200m observe 100% lethality which was low when compared to BLEVE of LPG because the capacity of LPG is more compared to Propylene.

The received heat fluxes in the horizontal and in the vertical direction were plotted with respect to distance from the event occurrence and were shown in figure 8.5. From the figure it is clear that the horizontal flux is greater than the vertical flux.

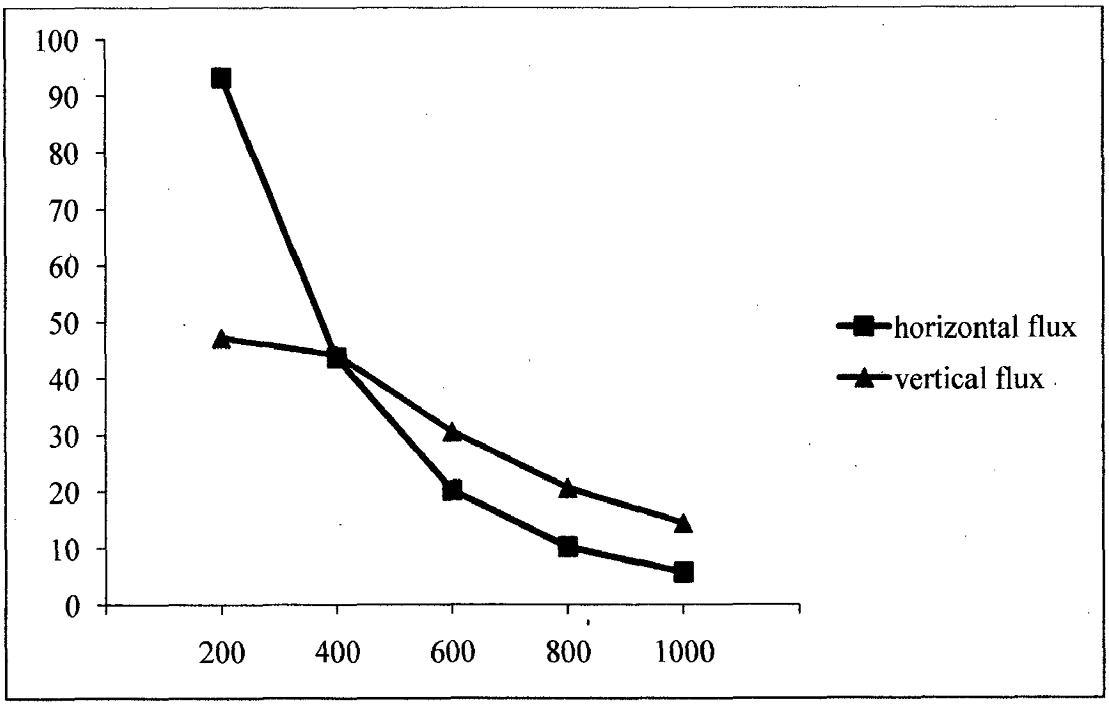


Figure 8.6 Distance vs. Radiant Heat Flux for Propylene Horton sphere

8.2 RESULTS FROM RISK MAPPING

1. The major contributors for risk to the society were from
   Failure of LPG Horton sphere (Outcome-BLEVE)
   Failure of Propylene Horton sphere (Outcome-BLEVE)
2. The individual risk lies in the acceptable region in most of the areas.
3. The F-N curve of societal risk lies to some extent in the unacceptable region. This was entirely due to LPG and Propylene Horton spheres.

CHAPTER 9

CONCLUSION AND RECOMMENDATIONS

9.1 CONCLUSION

Following conclusions were drawn from the results discussed in chapter 8.

- The dominant risk contributors were

  Scenario 3: Failure of LPG Horton sphere (Outcome-BLEVE)

  Scenario 4: Failure of Propylene Horton sphere (Outcome-BLEVE)
- The catastrophic failure frequency of a Horton sphere due to BLEVE was given by Delvosalle et al (2004) as 8.0E-5 yr^{-1}. The one calculated here using the fault tree analysis was found to be 3.6E-7 yr^{-1}. This shows that case specific analysis is to be used for the potential risk contributors rather than using the generic data.
- The individual risk is acceptable for the facility. Societal risk lies some extent in the unacceptable region. Thus the inventories of highly inflammable substances are to be installed away from the densely populated areas.

9.2 RECOMMENDATIONS

LPG and Propylene Horton spheres pose a major offsite risk. The offsite areas like Port Colony and Malkapuram town, which are towards the south of the refinery will be most affected, in case of potential BLEVE.

The following are recommended:

- The inventories of hazardous material should be reduced so as to make it an inherently safer plant or the process to be modified to reduce the operating temperatures and pressures.
- The Horton spheres are well designed to the current best practices and are state-of-art. There is no scope for further risk reduction through technological means. The accident likelihood depends upon the effectiveness of the Safety Management System (SMS). Therefore, LPG and Propylene storage handling system deserves special attention in terms of inspection, maintenance and SMS.
- Addition of Horton sphere will increase the BLEVE probability. Hence, no further addition should be made.

- In the long run, at an appropriate opportunity HPCL may consider alternate storage mode in the place of Horton spheres. Mounded bullets offer the best option. For mounded bullets, BLEVE is not a credible scenario and can be ruled out, thereby eliminating the dominant contributors of societal risk from the storage of LPG.

REFERENCES

AIChE/CCPS. (1992). Guidelines for Hazard Evaluation Procedures second edition with worked examples. Center for Chemical Process Safety, American Institute of Chemical Engineers, New York.

AIChE/CCPS. (2000). Guidelines for Chemical process Quantitative Risk analysis second edition. Center for Chemical Process Safety, American Institute of Chemical Engineers, New York.

AIChE/CCPS. (1999). Guidelines for Consequence Analysis of Chemical Releases. Center for Chemical Process Safety, American Institute of Chemical Engineers, New York.

AIChE/CCPS. (1997). Guidelines for Postrelease Mitigation Technology in the Chemical Process Industry. Center for Chemical Process Safety, American Institute of Chemical Engineers, New York.

Arendt, J. S. (1990). Using Quantitative Risk Assessment in the Chemical Process Industry. Reliability Engineering and System Safety, 29, 133-149.

Beerens, H.I., et al. (2006). The use of generic failure frequencies in QRA: The quality and use of failure frequencies and how to bring them up-to-date. Journal of Hazardous Materials, 130, 265–270.

Chang, J.I., and C.C. Lin (2006) A study of storage tank accidents Journal of Loss Prevention in the Process Industries 19, pp 51–59.

Hope et al.(1984a). Methodologies for Hazard Analysis and Risk Assessment in the Petroleum Refining and Storage Industry-Part I. Fire Technology, August.

Hope et al.(1984b). Methodologies for Hazard Analysis and Risk Assessment in the Petroleum Refining and Storage Industry-Part II. Fire Technology, November.

Khan, F.I., and S.A. Abbasi (1998). Techniques and methodologies for risk analysis in chemical process industries. Journal of Loss Prevention in the Process Industries, 11, 261-277

Khan, F.I., and S.A. Abbasi (1999). Major accidents in process industries and an analysis of causes and Consequences. Journal of Loss Prevention in the Process Industries, 12, 361–378

Khan, F.I., and S.A. Abbasi (2001). Risk analysis of a typical chemical industry using ORA procedure. Journal of Loss Prevention in the Process Industries, 14, 43–59.

Pietersen, C. M., and B. F. P. van het Veld (1992) Risk assessment and risk contour mapping. Journal of Loss Prevention in the Process Industries, Vol5, No 1.

Pikaar, J. (1995). "Risk Assessment and Consequence Models." 8th International Symposium on Loss Prevention and Safety Promotion in the Process Industries, June 6-9, 1995, Antwerp, Belgium, Keynote lecture on Theme 4.

Sam Mannan and Frank. P. Lees (2005). Lee's Loss Prevention in the Process Industries. (Vol. 1-3), Elsevier Butterworth-Heinemann.

Tweeddale, H.M., R.F., Cameron and S.S., Sylvester (1992). Some experiences in hazard identification and risk shortlisting. Journal of Loss Prevention in the Process Industries, Vol5, No 5, 279-288.

Theodore, L., J. P. Reynolds, and F. B. Taylor (1989). Accident and Emergency Management. New York: John Wiley & Sons.

Clancey, V. J. (1972). Diagnostic Features of Explosion Damage. 6th International Meeting on Forensic Sciences. Edinburgh, Scotland.

Ditali, S et al. (2000). Consequence analysis in LPG installation using an integrated computer package. Journal of Hazardous Materials, 71, 159–177.

Abbasi, T and S. A. Abbasi (2007). The boiling liquid expanding vapour explosion (BLEVE): Mechanism, consequence assessment, management. Journal of Hazardous Materials, 141, 489–519.

Alonso, F. D. et al. (2006). Characteristic overpressure–impulse–distance curves for vapour cloud explosions using the TNO Multi-Energy model. Journal of Hazardous Materials, A137, 734–741.

Crowl, D. A., and J. F. Louvar (2001). Chemical Process Safety: Fundamentals With Applications. 2nd ed., Prentice Hall, NJ.

KLG-TNO (2000). Risk Analysis Study at HPCL-VR. KLG-TNO Safety Technology Limited.

Mercx, W.P.M. et al (2000). Developments in vapour cloud explosion blast modeling. Journal of Hazardous Materials, 71, 301–319.

Early, W.F. (2006). Database management systems for process safety. Journal of Hazardous Materials, 130, 53–57.

Moosemiller, M. (2006). Avoiding pitfalls in assembling an equipment failure rate database for risk assessments. Journal of Hazardous Materials, 130, 128–132.

Melchers, R.E., and W.R. Feutrill (2001). Risk assessment of LPG automotive refuelling facilities. Reliability Engineering and System Safety, 74, 283-290.

Cox, A.W., Lees, F.P. and Ang, M.L. (1990). Classification of Hazardous Locations. Institution of Chemical Engineers, Rugby.

Delvosalle C., Fiévez C. and Pipart A. (2004). Generic frequencies data for the critical events. ARAMIS D1C, Appendix 10.

APPENDIX A

CONFIGURATION OF THE REFINERY

PROCESS UNITS

Crude Distillation Units: Three nos.
CDU I : 1.5 MMTPA (30400 bbls/day)
CDU II : 3.0 MMTPA (60800 bbls/day)
CDU III : 3.0 MMTPA (60800 bbls/day)

Fluidized Catalytic Cracking Units: Two nos.
FCCU I : 0.95 MMTPA (18000 bbls/day)
FCCU II : 0.65 MMTPA (12300 bbls/day)

Vis-Breaker Unit: One no.
VBU : 1.0 MMTPA (17300 bbls/day)

Propylene Recovery Unit: One no.
PRU : 0.023 MMTPA (800 bbls/day)

Bitumen Blowing Unit: One no.
BBU : 0.225 MMTPA (3900 bbls/day)

Product Treating Units:
DHDS : 1.8 MMTPA (37000 bbls/day)
LPG Amine Treatment Unit
LPG, ATF and Petrol Merox Units
Amine Regeneration Unit

Environmental Control Facilities:
SRU : 3 nos.
Sour Water Striping Units: 2 nos.
Effluent Treatment Plants: 4 nos.
CO Boilers : 2 nos.

OFFSITE FACILITIES

HPCL-Visakh Refinery has fully fledged independent offshore crude receiving facilities, well established blending and shipping facilities.

PRODUCTS

LPG:	Domestic (Cooking Gas)
	Commercial (Cooking Gas)
	Industrial ([Fuel)
Petrol:	Transportation Fuel
	Normal petrol
	0.05 wt% Sulphur petrol
Kerosene:	Domestic (Cooking Fuel/ Lamp)
Diesel:	Transportation Fuel
	Normal Diesel
	0.05 wt% Sulphur Diesel
	Low Sulphur High Flash Diesel (Navy Grade Diesel)
	Light Diesel Oil (Marine Diesel: Shipping)
Aviation Turbo Fuel	
Propylene Petrochemical Feed Stock	
Naphtha:	Industrial, Power Generation
	Fuel
	Fertilizer
Mineral Turpentine Oil (Paint Industry Solvent)	
Jute Batching Oil (Jute Industry Solvent)	
Fuel Oil:	Light
	Heavy
Low Sulphur Heavy Stock	
Bitumen	
Sulphur	

APPENDIX B

METHODOLOGY FOR HAZARD ANALYSIS STUDY

STEP I: SELECTION OF HAZARDOUS ACTIVITIES WITHIN THE SITE

In Step I the selection of units or part of units that can pose a risk for the surroundings was to be carried out. Most of the applied directives in European countries and the USA were based on a selection by use of the inventory pressure, temperature and the threshold value for that specific compound or class to which the compound belongs. The threshold values were chosen in such a way that inventories below the threshold will not form a major hazard to the surroundings.

This methodology distinguishes the following subsequent steps:

1. Division of the installation into isolatable sections like: separated units in a process, storage units, transport units (road tankers, rail wagons, etc.).
2. Determination of threshold quantity G for the hazardous chemical.
3. Calculation of nomination factor, A. This factor accounts for type of chemical involved, potential hazard of the chemical, type of installation as well as process conditions like temperature and pressure. The A-factor is calculation for each unit or part of the installation identified in 1).
4. Calculation of selection factor, S. This factor, combining the A-factor and the distance L between the installation and a nearby residential area, results in a measure of the potential hazard to vulnerable area. The calculation of the S-factor is the final step in the selection procedure. The S-factor demands whether an installation or part of the installation can give a significant contribution to the potential risk outside the boundaries of the premises.

A detailed description of the selection methodology was given in steps 1 to 4.

Step 1: Division of the installation in isolatable units

First the installation must be divided in storage units and separated units in a process.

Storage units: Each storage tank with a capacity of equal or more than the threshold value of the considered compound can be considered as a single unit.

Process units: Process units were separated first by means of the location of different stages in the process. Besides that the process must be divided in coherent parts of one or more sections in which the same materials will be handled and/or units which can be isolated from each other in case of emergency by means of quick isolation valves.

Step 2: Determination of the Threshold Quantity, G

The threshold quantities of hazardous materials were determined by the amount that can cause a potential danger for human beings at a distance of 100 meter from the point of release.

The following threshold values were used:

Flammable products: For all flammable products a threshold quantity was used of G= 10,000 kg.

Toxic products: All toxic products (liquid/gases) were considered irrespective of the inventory. However, for calculation of nomination factor A, following threshold values are used:

The threshold quantity of chlorine was G=300 kg. (LC_{50}= 864 mg/m^3, 1 hr exposure)
The threshold quantity of H_2S was G=300 kg. (LC_{50}= 898 mg/m^3, 1 hr exposure)
The threshold quantity of ammonia was G=3,000 kg. (LC_{50}= 11,590 mg/m^3, 1 hr exposure)
The threshold quantity of CO was G=3000 kg. (LC_{50}= 7,949 mg/m^3, 1 hr exposure)

Step 3: Calculation of process circumstance factor, O

The potential danger of an installation will be formed by the physical, flammability and/or toxic properties of the compound and process parameters like pressure and temperature.

The process parameters were taken into account by a ratio to a reference installation. The deviation of the actual process conditions with respect to the reference conditions will be corrected by means of a so-called circumstance factors O1, O2 and O3. The different circumstance factors were listed in the table H1 and H2.

The total circumstance factor (O_{total}) was equivalent to the product of all relevant circumstance factors:

$$O_{total} = O1 * O2 * O3$$

Table B1. Determination of Circumstance Factors

Circumstances	Condition	O-factor
O1: Type of activity	Process installation Storage activity	O1=1 O1=0.1
O2: Location of activity	Enveloped [1]) Outdoors	O2=0.1 O2=1
O3: Physical state of chemical	Gas Liquid [2]) Solid (powder)	O3=10 O3= X_1+X_2 [3]) O3=0.1

Notes:

[1]) Also valid for bunds if process temperature equal or lower than atmospheric boiling point.

[2]) For values of X_1 and X_2 see Table B2.

[3]) Total of X_1 and X_2 should not exceed 10.

Table B2: Determination of correction factors X_1 and X_2

Process conditions [1])	X_1	X_2
T_{proc} [4]) > T_{boil} and P_{sat} > 3bar	10	
T_{proc} > T_{boil} and 1bar < P_{sat} < 3bar	1-10 (interpolate) [2])	
T_{proc} < T_{boil}	P_{sat} [bar] [3])	
T_{proc} < 25°C and -25°C < T_{boil} < 25°C		0
T_{proc} < 25°C and -75°C < T_{boil} < -25°C		1
T_{proc} < 25°C and -125°C < T_{boil} < -75°C		2
T_{proc} < 25°C and -75°C < T_{boil}		3

Notes:

[1]) T_{boil} was atmospheric boiling point of pure compound, or the starting point of a boiling range.

[2]) Linear interpolation of X_1 from 1-10 (In round figures).

[3]) P_{sat} was the partial vapour pressure at the process temperature; lowest value of X_1 = 0.1.

[4]) T_{proc} was the nominal process or storage temperature.

Step 4: Calculation of the Notification factor A

The A-factor was a (semi-) quantitative measure of the potential impact of releases of hazardous materials in the surroundings. In case the A-factor=1 this means that the consequence distance (potential fatality) is about 100 meter.

The A-factor was calculated by multiplying the total amount of hazardous material Q and the factor of circumstance O_{total} divided by the threshold quantity G of the compound:

$$A=(Q * O_{total})/G$$

For the calculation of the A-factor spreadsheet software developed by TNO was used. This spreadsheet enables the calculation of the A-factor rather easily and quickly, even for a large number of units.

Step 5: Calculation of the Selection factor S

The S-factor for a hazardous activity was calculated from the A-factor and the distance of the unit to nearby residential areas. The method distinguishes between toxic and flammable materials.

The S-factor is a measure for the potential consequences of a release of hazardous materials from that particular unit. The factor was used for the selection of units for which accidental releases of hazardous materials can be a danger for the surroundings and which consequently must be taken into account for hazard analysis.

For toxic materials, the S-factor was calculated according to the following equation:

$$S = (100/L)^2 * A \qquad (1)$$

For flammable materials, the S-factor was calculated according to the following equation:

$$S = (100/L)^3 * A \qquad (2)$$

In which:

L = distance from selected unit to plant boundaries, in [m]

A = calculated A-factor

If the distance is less than 100 m take L=100.

The S-factors need to be calculated for each of the identified units. In principle, all units were selected for further inclusion in the hazard analysis if:

At any location along the fence of the premises $S \geq 1$

It was not always necessary to calculate the S-factor for all units present. By use of quick scan of the calculated A-factor, a first pre-selection can identify and neglect the units for which it was obvious that they do not contribute to the potential danger for the surroundings. This quick scan can be carried out in the following way:

- Establish the shortest distance between the unit and the premises fence;
- Establish the minimum A-factor necessary to select a unit, i.e. $S > 1$ (by use of the appropriate formulae (1) and (2) above, for toxic or flammable materials);
- Select the relevant unit with an A-factor equal or higher than the "threshold" A-factor.
- In determination of the relevant distance, the different locations of the various units have to be accounted for.

Completion of steps 1 to 5 results in a list of release sources to be further evaluated in the effects and consequences assessments.

STEP II: IDENTIFICATION OF ACCIDENT SCENARIOS FOR THE SELECTED UNITS.

For the selected units release consequence calculations have been done assuming a catastrophic failure. For vessels, including storage tanks, a catastrophic failure is defined as failure of the largest diameter nozzle connection to that vessel. This establishes the worst case consequences from loss of containment of that inventory.

APPENDIX C

MODEL CALCULATIONS

Scenario 1: Failure of H_2S line from DHDS Amine Regenerator to DHDS sulphur Recovery Unit

Assume that the plant fence line is 500m away from the release

Assume E-2m/s

a). The time required to reach the fence line

$$t = \frac{x}{u}$$

$$= \frac{500}{2} = 250\,\mathrm{sec} = 4.2\,\mathrm{min}$$

This leaves very little time for evacuation

b). Maximum concentration outside the fence line

$$\langle c \rangle (u_t, 0, 0, t) = \frac{G^*}{\sqrt{2} * \pi^{3/2} * \sigma_x \sigma_y \sigma_z}$$

$$\sigma_x = \sigma_y = 0.04(x)^{0.92}$$

$$= 0.04(500)^{0.92} = 12.16\ \mathrm{m}$$

$$\sigma_z = 0.1 * \left(x^{0.65}\right) = 0.1(500)^{0.65}$$

$$= 5.68\ \mathrm{m}$$

$$\langle c \rangle (u_t, 0, 0, t) = \frac{360}{\sqrt{2} * 3.14^{3/2} * 12.16^2 * 5.68}$$

$$= 0.0545\ \mathrm{kg/m^3}$$

$$= 54520.75\ \mathrm{mg/m^3}$$

Which is a very high concentration near the plant boundary

c).The IDLH value for H_2S is 10 ppm = 14mg/m^3

$$14 * 10^{-6} kg/m^3 = \frac{360}{\sqrt{2} * \pi^{3/2} * \sigma^2{}_y \sigma_z}$$

$$\Rightarrow \sigma_y{}^2\sigma_z = 3.2654*10^6 m^3$$

$$\Rightarrow (0.04*x^{0.92})^2(0.10*x^{0.65}) = 3.2654*10^6 m^3$$

$$\Rightarrow x = 13.8km$$

d). Width of the puff at a downward location of 2 km

$$\langle c\rangle(x,0,0,t) = \frac{G^*}{\sqrt{2}*\pi^{3/2}*\sigma_x\sigma_y\sigma_z}\exp\left[\left(\frac{-1}{2}\right)\left(\frac{x-ut}{\sigma_x}\right)^2\right]$$

$$\sigma_x = \sigma_y = 0.04*(2000)^{0.92} = 43.55m$$
$$\sigma_z = (0.10)*(4000)^{0.65} = 13.98m$$

$$14*10^{-6} = \frac{360}{\sqrt{2}*\pi^{3/2}*43.55^2*13.98}\exp\left[\left(\frac{-1}{2}\right)\left(\frac{x-ut}{43.55}\right)^2\right]$$

$$x-ut = 135m$$

The puff thickness=2*135=270 m

At wind speed of 2m/s, the puff will take (270/2)=135 sec to pass

Scenario 2: Failure of Naphtha tank

TNT equivalent mass of fuel,

$$W = \left[\frac{\eta M E_c}{E_{TNT}} \right]$$

$$= \frac{0.05 * 39300 * 46.6 * 10^3}{4652}$$

$$= 19683.8 kgTNT$$

$$Z = \frac{R}{W^{\frac{1}{3}}} = \frac{1000}{\sqrt[3]{19683.8}} = 3.7037 m / kg^{1/3}$$

$0.0674 \leq Z \leq 40$

From literature, a= -0.2144 and b=1.3503

$$a + b \log z = 0.5534$$

$$\log p^0 = \sum_{i=0}^{11} c_i (a + b \log z)^i$$

$$= \sum_{i=0}^{11} c_i (0.5534)^i$$

$$\log p^0 = 2.781 - 1.696(0.5534) - (0.1542)(0.5534)^2$$

$$\log p^0 = 1.8773$$

Scenario 3: BLEVE of LPG Horton sphere

Number of fragments $= -3.77+0.0096*[\text{vessel capacity, m}^3]$

Range of validity is 700-2500 m^3

$= -3.77 + 0.0096*2572$

$= 20.9212$

$\cong 21$ fragments

Maximum fire ball diameter

$D_{max} = 5.8*M^{1/3}$

$= 5.8*(12E^5)^{1/3}$

$= 616.342$ m

Fire ball combursion duration

$t_{BLEVE} = 0.45*M^{1/3}$ (M<30000 kg)

$t_{BLEVE} = 2.6*M^{1/6}$ (M>30000 kg)

$= 2.6*(12E^5)^{1/6}$

$= 26.8$ sec

Centre height of fire ball

$H_{BLEVE} = 0.75*D_{max}$

$= 0.75*(616.342)$

$=462.2565$ m

Initial ground level hemisphere diameter

$D_{intial} = 1.3\ D_{max}$

$=1.3*(616.342)$

$=801.2446$ m

The emitted flux at surface of the fire ball

$$E = \frac{RMH_c}{\pi * D^2{}_{\max} * t_{BLEVE}}$$

$$= \frac{0.3 * 12E^5 * 45840}{\pi * 616.342^2 * 26.8}$$

$$= 516 \text{ kW/m}^2$$

Path length

$$X_s = \sqrt{H_{BLEVE}{}^2 + L^2} - \frac{D_{max}}{2}$$

$$= \sqrt{462.5265^2 + 200^2} - 616.342/2$$

$$= 195.5 \text{ m}$$

The view factor assuming a vertically oriented target is determined from

$$F_{21} = \frac{L * (D/2)^2}{(L^2 + H^2{}_{BLEVE})^{3/2}} \quad (L > D_{max}/2)$$

$$F_{21} = \frac{H * (D/2)^2}{(L^2 + H^2{}_{BLEVE})^{3/2}}$$

$$= \frac{462.2565 * (616.342/2)^2}{(200^2 + 462.2565^2)^{3/2}}$$

$$= 0.3435$$

$$\tau_a = 2.02 * (P_w * X_s)^{-0.09}$$

$$= 2.02[2810 * (195.5)]^{-0.09}$$

$$= 0.615$$

$$E_r = \tau_a E F_{21}$$

$$=0.615*516*0.3435$$

$$=109 \text{ kW/m}^2$$

Scenario 4: BLEVE of Propylene Horton sphere

The calculations for this scenario were similar to scenario 3. Only the input values, in the calculation of emitted heat flux, like heat combustion were to be changed.

APPENDIX D

DATABASE DEVELOPMENT

D.1 Demographic Data

Population within 5-km belt around the refinery has been considered while preparing the grid. While preparing the population grid, it is assumed that the population is evenly distributed in the area occupied by it though this may not be strictly true. The consequences for various outcome cases such as thermal loads, vapour cloud explosion, toxic exposure depend also on whether people are indoor or outdoor. The following assumptions have been made:

Day time – 30% indoor, 70% outdoor
Night time – 70% indoor, 30% outdoor

Besides the difference in numbers present outdoors during the day and during the night, the total number of people present as well as in a grid may differ between day and night. Separate population grids are required to be filled in, one for daytime and the other for night-time. Here the day and night population in the surrounding villages and townships has been assumed to be same.

Table D.1: Population within 5km belt around HPCL-VR

(Source: District Census handbook, KLG-TNO, 2000)

	Locality	Distance & direction of village with ref. to HPCL	Area (sq. km.)	Population
1	Venkatapuram	5 Km/NW	4.14	9517
2	Malkapuram	2 Km/SE	--do--	1991
3	Gajuwaka	3 Km/W	22.08	42915
4	Peddagantyada	4 Km/S	--do--	51676
5	Gopalapatnam	3 Km/N	7.95	30600
6	Vepagunta		--do--	12578

Refinery was situated at about 1km (Northward) from the foot of Yarada hills. On the eastern side were Hindustan Shipyard, Eastern Naval Command, HPCL Terminal and IOC Terminal. Coromandel Fertilisers, Hindustan Zinc and Bharat Heavy Plates and Vessels (BHPV) were on the Western side. Residential colonies were located on the Southern side. On the Northern side was Additional Tankage Project (ATP) and the Andhra Petrochemical Plant was on the North-Eastern side. The Refinery had road access from the southern side (Old National Highway) and North-Eastern side.

In the assessment of the societal risk, demographic data on the offsite of the industrial activity was used. Only the employees working in the concerned industry were not included in the 'society'. With regard to the risk of the people employed in nearby industries with similar risk, it was debatable to consider them as population. Internationally consensus is lacking on this aspect. However, in the present study employees in the adjoining industries were not considered in the assessment of societal risk.

The population is mainly on the southern side. The detailed information on population distribution is not available. From the town data (area and total population) population density comes out to be 2159/sq.km., which is fairly representative of the populated area in the vicinity of the refinery. Therefore, population density of 2159/sq.km has been used in the calculation of societal risk.

D.2 Meteorological Data

Meteorological and climatological data collected by India Meteorological Department (IMD), (KLG-TNO, 2000) for Visakhapatnam has been used for the risk analysis computations. Based on this data the annual mean air temperature and mean % humidity have been calculated. The data on cloud cover and wind speed have been used to select appropriate atmospheric stability class and wind speed combinations. Distribution of wind direction frequency is a crucial parameter that is subject to significant variation from year to year and only averages over very long periods are relevant. Climatological tables prepared by IMD utilizing the wind frequencies data collected for a period of thirty years have been used in this study.

D.2.1 Annual mean air temperature

Annual mean air temperature is 27^0C.

D.2.2 Annual mean % humidity

Annual mean % humidity is 78%.

D.2.3 Wind speed

The annual wind speed is around 3m/s. During January to March and September to December (7 months in a year) the wind speed is less than 2.7m/s. During the rest of the year, from April to August, the average wind speed is between 4.0 to 4.6m/s.

D.2.4 Stability Class

Dispersion of gases or vapours is influenced to a large extent by the atmospheric stability. The various Pasquill stability classes that are defined are:

A	Very Unstable
B	Unstable
C	Slightly Unstable
D	Neutral
E	Stable
F	Very Stable

The stability class at a particular location is generally dependent upon:

Time of the Day (Day or Night)
Cloud Cover
Season
Wind Speed

Six stability classes from A to F are defined while wind speed can take any one of numerous values. It may thus appear that a large number of outcome cases can be formulated by considering each one of very many resulting stability class-wind speed combinations. However in fact the number of stability class-wind speed combinations that needs to be considered for formulating outcome cases in any analysis is very limited. This is because, in nature, only certain combinations of stability class and wind speed occur. Thus, for instance, combinations such as A-3 m/s or B-5 m/s or F-4 m/s do not occur. As a result only one or two stability class-wind speed combinations need to be considered to ensure reasonable completeness of a QRA. Wind speed does not influence the consequences, as much as stability class and for a given stability class, the influence of wind speed is relatively less. On the other hand consequences vary considerably with stability class for the same speed.

The cloud cover data is as follows:

January-May (5 months) the cloud cover (all clouds) is 1.7-3.9 oktas
June-September (4 months) the cloud cover (all clouds) is 5.8-7 oktas
October-Dec. (3 months) the cloud cover (all clouds) is 2.5-5.2 oktas

Thin level of cloud cover along with the prevailing low wind velocities results in unstable conditions during the day (B) and highly stable conditions (E or F) at night. During the four months of monsoon (June-September) the cloud cover is about 6 oktas and this results in stability class B or C during the day and E or F during the night. The stability class distribution over the years works out as below:

B	33.3% (day other than monsoon)
C	16% (day monsoon)
E	50% (night)

The following wind stability class combinations and frequencies have been chosen for QRA:

B	3m/s	50%
E	2m/s	50%

D.3 GENERIC FAILURE FREQUENCIES AND DATA SOURCES

This section presents a compilation of failure frequency data from a variety of sources Data for the present QRA study has been chosen from this compilation

D.3.1 E&P Forum The Oil Industry International Exploration & Production Forum

The E & P Forum database ('Hydrocarbon leak and ignition data base', Report No 11.4/80 May 1992, The Oil Industry International Exploration & Production Forum (E&P Forum)) for leaks and ignition probabilities gives the cumulative frequencies for all hole sizes up to full bore for piping and other equipment. The objective of this database is to compile a database of existing leak frequency data and ignition probabilities for use in quantitative risk analysis. Experience data from operating companies in the petroleum, petrochemicals and nuclear industries have been used in this database. Apart from developing leak frequencies per equipment item, generation of a hole size distribution i.e., information about the size of releases, has been given priority The hole size distributions were often based on engineering judgement as very little detailed information is available in this area. The data taken from this database and used for this QRA were listed below.

Pressure Vessels

This category includes the pressure vessel itself but excludes all valves, piping, fittings beyond the first flange and the flange

Failure frequency

$1.5x10^{-4}$ per vessel year

Hole size distribution

Hole size (d_{hole})	$P(d_{eq}<d_{hole})$
25 mm	0.54
50 mm	0.89
150 mm	0.96
Full bore	1.00

(full bore is usually taken as the inner diameter of the largest connecting pipe)

Process piping excluding valves and flanges

Process piping excludes valves and flanges The frequencies for leaks from process piping are very low, typically around 10^{-6} per metre year This implies that a very large population is needed to predict reliable leak frequencies The most common method to obtain frequencies is therefore to extrapolate results from a smaller set of data, usually by making the frequency dimensional i.e., presenting the frequency in units [length of pipe/Diameter of the pipe]

Failure frequency

Since the data do not allow calculations of frequencies per diameter of pipe, pipe sizes are divided in three groups and the frequency calculated for each

Small bore $D \leq 3''$ $f = 7.0x10^{-5}$ /m year

Medium	$4" \leq D \leq 11"$	$f = 3.6x10^{-5}$ /m year
Large	$D \geq 12"$	$f = 2.7x10^{-5}$ /m year

Hole size distribution

Hole size (d_{hole}/D)	**$P(d_{eq}<d_{hole})$**
0.05	0.60
0.22	0.85
0.45	0.95
1.00	1.00

Valves

Failure of Valve body, stem and packer, excluding flanged joints. Full-bore ruptures are not considered credible. Failure frequency is $2.3x10^{-4}$ per valve year.

Hole size distribution

Hole size (d_{hole}/D)	**$P(d_{eq}<d_{hole})$**
0.05	0.65
0.10	0.88
0.20	0.94
1.00	1.00

Pump Centrifugal

For pump (double seal) excluding all valves, pipes and fittings beyond first flange and the flange itself, the hole size distribution has been generated from qualitative description of leaks.

Failure frequency: 1 7x 10^{-2} /yr

Hole size distribution (full bore is usually taken as inner diameter of largest pipe connected to the pump)

Hole size (d_m)	**$P(d<d_m)$**
10 mm	0.82
50 mm	0.96
Full bore	1.00

D.3.2 Failure Frequency data from TNO database

Leaks and rupture of piping

Several studies from the past have led to a distribution in failure probabilities for two categories of pipelines. Process piping (i.e. inside a plant) and long range transport pipelines. The figures accounts for the following assumptions:

- Regarding the more fluctuating process conditions in and more frequent activities around process pipelines, it is expected that failure probabilities in this group are higher than for transport pipelines.
- It is assumed that larger diameter pipeline will have a larger integrity and consequently a lower frequency of being damaged than small ones.
- For pipelines with a d > 6" (150m) a guillotine rupture is not credible.

Generic failure case frequencies per year per meter for transfer and process pipelines

	TRANSFER			PROCESS		
Diameter	**Rupture**	**Hole 10%**	**Hole 1%**	**Rupture**	**Hole 10%**	**Hole 1%**
d < 3"	1×10^{-6}	3×10^{-6}	1×10^{-5}	3×10^{-6}	1×10^{-5}	3×10^{-5}
3 < d < 6"	3×10^{-7}	1×10^{-6}	3×10^{-6}	1×10^{-6}	3×10^{-6}	1×10^{-5}
d > 6"	-	3×10^{-7}	1×10^{-6}	-	1×10^{-6}	3×10^{-6}

Catastrophic failure

Pressure Vessels	1×10^{-5} /yr
Atmospheric Storage Tanks	1×10^{-6} /yr

D.3.1 Failure Frequency data from the Rijnmond Report (COVO study)

Base failure rate data for equipment

Piping, d < 50 mm	Rupture	8.77E-6/m/yr
	Significant Leak	8.77E-5/m/yr
Piping, 50 mm < d < 150 mm	Rupture	2.62E-8/m/yr
	Significant Leak	5.20E-6/m/yr
Piping, d > 150 mm	Rupture	8.70E-8/m/yr
	Significant Leak	2.62E-6/m/yr

Pressure Vessels	Serious leakage	1E-5/yr
	Catastrophic failure	1E-6/yr
Atmospheric Storage tank leak	Serious leakage	1E-4/yr
	Catastrophic failure	6E-6/yr

D.3.4 BLEVE Likelihood

The likelihood of catastrophic failure of liquefied pressurised gas storage with occurrence of a BLEVE is dependent on the design of the storage installation, and in particular the protective features present. As a result, this likelihood is usually not obtained from generic failure data but is instead modelled using a fault tree.

D.3.5 Ignition Probability

It was necessary in Quantitative Risk Assessment to estimate the probability of ignition and of explosion if a leak occurs. This was an area where there is relatively little guidance available.

Ignition of a leak may occur either at the point of leak or at some distance from it. The cause of ignition may be the leak event itself or an ignition source. Probability of ignition is a function of material released, release location, release conditions, release rate and duration, number and strength of ignition sources.

Ignition probabilities for leaks of flammable fluids taken from ignition model developed by Cox et al (1990) were given below:

Leak (kg/s)	**Probability of given ignition**	
	Gas	**Liquid**
Minor (<1)	0.01	0.01
Major (1-50)	0.07	0.03
Massive (>50)	0.3	0.08

D.4 THERMAL EFFECTS

The thermal effects were simple and are based on extensive experimental data. Duration of exposure was also considered. People exposed to a thermal radiation from a fire will tend to take escape actions. Table D.2 describes the damage caused to process equipment with variation in the received flux.

Table D.2 Damage Caused at Various Incident Levels of Thermal Radiation (Theodore et. al., 1989, AIChE/CCPS, 1997)

Incident Flux (kW/m^2)	**Type of Damage Caused**[a]
37.5	Sufficient to cause damage to process equipment: 100% lethality
25.0	Minimum energy required to ignite wood at infinitely long exposures (non-piloted): 100% lethality
12.5	Minimum energy required for piloted ignition of wood, melting plastic tubing: 100% lethality
4.0	Sufficient to cause pain to personnel if unable to reach cover within 20 seconds; blistering of skin (first degree burns are likely): 0% lethality
1.6	Will cause no discomfort over long exposure

At the lower levels, where time is required to cause serious injury to people, there is often the possibility to escape or take shelter. The accuracy of the incident flux damage relationship is considered to be adequate for initial hazard assessments and within the estimation of hazardous incidents. The correlation of thermally induced damage or injury may be applied to hazard assessment.

Effective exposure duration of 10 seconds was assumed for calculation. In case of BLEVE the exposure duration was taken as BLEVE duration. The limit for 1% of the exposed people to be killed due to heat radiation was presented in Table D.3.

Table D.3 Radiation Consequences

Exposure duration, s	Radiation level for 1% lethality, kW/m^2	
	With protection	Without protection
10	21.2	16.7
20	9.3	7.3

Since in practical situations people exposed to heat radiation in the event of a fire will have sufficient clothing, protection by clothing had been assumed. For these reasons, Q = 21.2 kW/m^2 had been selected as the critical heat radiation for 1% lethality. This assumes that people would be able to find a cover or a shield against thermal radiation in 10 seconds time. Furthermore, 100% lethality may be assumed for all people suffering from direct contact with flames.

D.4 OVERPRESSURE EFFECTS

The overpressure generated from the explosion effect is sufficient to devastate the surroundings. The following Tables D.4a, b illustrate the damage estimates for common structures and process equipment based on overpressure.

Table D.4a. Damage Estimates for Common Structures Based on Overpressure (Clancey, 1972, AIChE/CCPS, 1999.).

Pressure		Damage
psig	kPa	
0.02	0.14	Annoying noise (137 dB if of low frequency 10-15 Hz)
0.03	0.21	Occasional breaking of large glass windows already under strain
0.04	0.28	Loud noise (143 dB), sonic boom, glass failure
0.1	0.69	Breakage of small windows under strain
0.15	1.03	Typical pressure for glass breakage
0.3	2.07	"Safe distance" (probability 0.95 of no serious damage below this value); projectile limit; some damage to house ceilings; 10% window glass broken
0.4	2.76	Limited minor structural damage
0.5-1.0	3.4-6.9	Large and small windows usually shattered; occasional damage to window frames
0.7	4.8	Minor damage to house structures
1.0	6.9	Partial demolition of houses, made uninhabitable
1-2	6.9-13.8	Corrugated asbestos shattered; corrugated steel or aluminum panels, fastenings fail, followed by buckling; wood panels (standard housing) fastenings fail, panels blown in
1.3	9.0	Steel frame of clad building slightly distorted
2	13.8	Partial collapse of walls and roofs of houses
2-3	13.8-20.7	Concrete or cinder block walls, not reinforced, shattered
2.3	15.8	Lower limit of serious structural damage
2.5	17.2	50% destruction of brickwork of houses
3	20.7	Heavy machines (3000 lb) in industrial building suffered little damage; steel frame building distorted and pulled away from

		foundations
3-4	20.7-27.6	Frameless, self-framing steel panel building demolished; rupture of oil storage tanks
4	27.6	Cladding of light industrial buildings ruptured
5	34.5	Wooden utility poles snapped; tall hydraulic press (40,000 lb) in building slightly damaged
5-7	34.5-48.2	Nearly complete destruction of houses
7	48.2	Loaded train wagons overturned
7-8	48.2-55.1	Brick panels, 8-12 inches thick, not reinforced, fail by shearing or flexure
9	62.0	Loaded train boxcars completely demolished
10	68.9	Probable total destruction of buildings; heavy machine tools (7000 lb) moved and badly damaged; very heavy machine tools (12,000 lb) survive
300	2068	Limit of crater lip

TABLE D.4b. Damage Estimates Based on Overpressure for Process Equipment

Equipment	Overpressure, psi																								
	0.5	1.0	1.5	2.0	2.5	3.0	3.5	4.0	4.5	5.0	5.5	6.0	6.5	7.0	7.5	8.0	8.5	9.0	9.5	10	12	14	16	18	20
Control house steel roof	A	C	D				N																		
Control house concrete roof	A	E	P	D			N																		
Cooling tower	B			F			O																		
Tank: cone roof		D				K							U												
Instrument cubicle			A			LM						T													
Fire heater				G	I					T															
Reactor: chemical				A				I					P					T							
Filter				H					F										V		T				
Regenerator						I				IP					T										
Tank: floating roof						K							U												D
Reactor: cracking							I							I							T				
Pine supports							P					SO													
Utilities: gas meter									Q																
Utilities: electronic transformer									H						I					T					
Electric motor										H								I							V
Blower										Q										T					
Fractionation column											R			T											
Pressure vessel: horizontal												PI						T							
Utilities: gas regulator												I								MQ					
Extraction column													I							V	T				
Steam turbine															I						M	S			V
Heat exchanger															I			T							
Tank sphere																I						I	T		
Pressure vessel: vertical																					I	T			
Pump																					I		V		

Key to Table D.4b

A. Windows and gauges broken

B. Louvers fall at 0.2- 0.5 psi

C. Switchgear is damaged from roof collapse

D. Roof collapses

E. Instruments are damaged

F. Inner parts are damaged

G. Brick cracks

H. Debris—missile damage occurs

I. Unit moves and pipes break

J. Bracing falls

K. Unit uplifts (half tilted)

L. Power lines are severed

M. Controls are damaged

N. Block walls fall

O. Frame collapses

P. Frame deforms

Q. Case is damaged

R. Frame cracks

S. Piping breaks

T. Unit overturns or is destroyed

U. Unit uplifts (0.9 tilted)

V. Unit moves on foundation

D.6 SUMMARY OF DATA USED FOR ANALYSIS

Population

Demography details including the population of various villages and towns in the southern side of the refinery showed that they are densely populated.

Meteorological Data

1. Ambient conditions were taken as average Relative humidity of 78%, and average Temperature of 27°C.
2. Release consequence calculations had been carried out for B-3 m/s and E-2 m/s stability class-wind speed combinations.

Failure Frequency Data

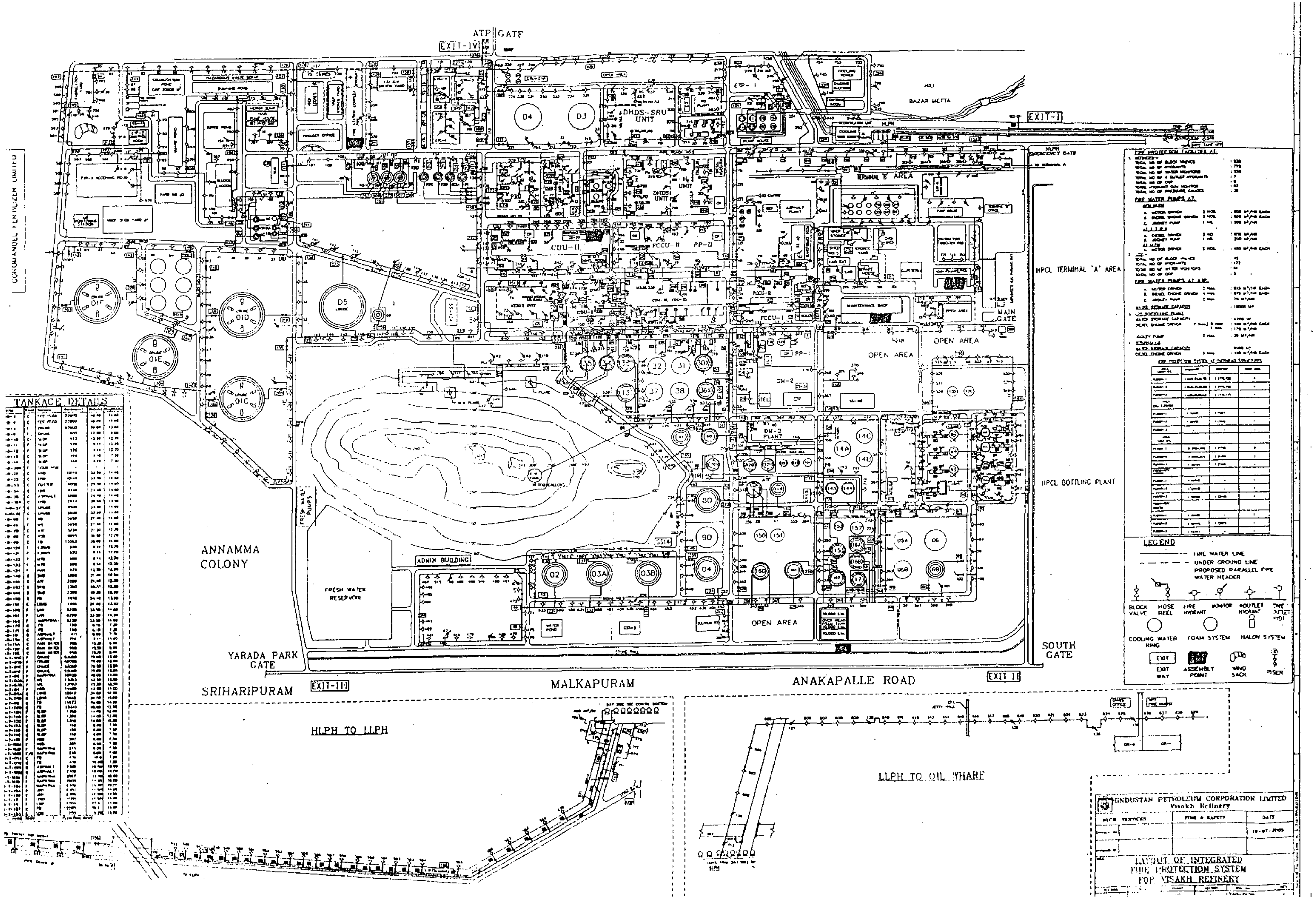
3. For piping failure with large hole size (d_{hole}/D = 0.45 to 1), the failure frequency has been taken as follows:

Size	**Small leak** **(d_{hole}/D = 0.22 to 0.45)**	**Large leak** **(d_{hole}/D = 0.45 to 1)**
$D \leq 3"$	7E-6 /m year	3.5E-6 /m year
$3" \leq D \leq 11"$	3.6E-6 /m year	1.8E-6 /m year

For Valves: For large release (full bore or maximum 6") failure frequency taken is 1 38E-5/valve yr.

4. For catastrophic failure of Atmospheric Storage Tanks the failure frequency has been taken as 1E-6/yr.
5. BLEVE likelihood is estimated by developing a Fault Tree.
6. Ignition probabilities for leaks of flammable fluids developed by Cox, Lees and Ang. had been used in the present study.
7. Radiation intensity is estimated from the above presented Tables D.2,3
8. The damage estimates common structures and process equipment based on overpressure was presented in Tables D.4a, b.

APPENDIX - E


ATP GATE
EXIT-IV
EXIT-I
BAZAR METTA
DHDS-SRU UNIT
CDU-II
FCCU-II
PP-II
TERMINAL 'B' AREA
HPCL TERMINAL 'A' AREA
MAIN GATE
OPEN AREA
FCCU-I
MAINTENANCE SHOP
PP-I
DM-2
DM-3 PLANT
HPCL BOTTLING PLANT
D4
D3
D5
01F
01D
01E
01C
ADMIN BUILDING
FRESH WATER RESERVOIR
FRESH WATER PUMPS
ANNAMMA COLONY
YARADA PARK GATE
SRIHARIPURAM
EXIT-III
MALKAPURAM
ANAKAPALLE ROAD
EXIT II
SOUTH GATE
COROMANDEL FERTILIZER LIMITED
TANKAGE DETAILS
HLPH TO LLPH
LLPH TO OIL WHARF
LEGEND
FIRE WATER LINE
UNDER GROUND LINE
PROPOSED PARALLEL FIRE WATER HEADER
BLOCK VALVE
HOSE REEL
FIRE HYDRANT
MONITOR
4 OUTLET HYDRANT
ONE OUTLET HYDT
COOLING WATER RING
FOAM SYSTEM
HALON SYSTEM
EXIT WAY
ASSEMBLY POINT
WIND SACK
RISER
FIRE PROTECTION FACILITIES AT
FIRE WATER PUMPS AT
FIRE PROTECTION SYSTEM AT OVERHEAD STRUCTURES
HINDUSTAN PETROLEUM CORPORATION LIMITED
Visakh Refinery
FIRE & SAFETY
DATE
LAYOUT OF INTEGRATED
FIRE PROTECTION SYSTEM
FOR VISAKH REFINERY