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Integration of qualitative and quantitative risk assessment methods for gas refinery plants

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Abstract−This paper focuses on the development of an integrated risk assessment system to increase the safety of gas refinery plants. This integration is important in managing the design and the operation of chemical plants and it requires significant time, effort and specialized expertise. We propose a systematic procedure to apply quantitative and qualitative hazard identification methods before constructing the refinery plants. Advantages of this assessment procedure are 1) systematic method in identifying most of the important hazards, 2) complete and rigorous analysis, and 3) reducing budget and time. The qualitative and quantitative assessment method consists of two and three steps, respectively. After these five steps, the risk of the same plant was calculated and analyzed. This method can identify risk and potential hazards from local gas refinery plants effectively and systematically. We applied the integrated qualitative and quantitative risk assessment method to remove the risk of a local gas refinery plant with 160 P&ID sheets. The result of this work confirmed that no residential population existed within the bounds of the 1×10^{-6} per annum risk contour. Therefore, the risk to the public from the gas refinery plants should be considered acceptable.

Key words: Integrated Risk Assessment, Quantitative and Qualitative Methods, Potential Hazards, Annum Risk Contour

INTRODUCTION

1. Related Studies

The initial effort to design the plant includes the prediction of potential hazards. Numerous types of hazard analysis methods have been proposed and used. Hazard analysis methods are classified by qualitatively and quantitatively. The qualitative methods include checklist, what-if analysis, preliminary hazard identification (PHI), and hazard and operability (HAZOP) study. The quantitative methods include event tree analysis (ETA), fault tree analysis (FTA), failure modes and effects analysis (FMEA) and symbolic model verification (SMV) [1,2].

However, an efficient method of integration of various methods is required for more complete safety evaluation. This integration is important in managing the design and the operation of chemical plants and it requires significant time, effort and specialized expertise. So we proposed a systematic procedure to apply quantitative and qualitative hazard identification methods before constructing the refinery plants.

2. Composition of the Risk Assessment Methods

We used PHI and HAZOP as qualitative assessment methods. PHI can be used to rank hazard qualitatively based on potential consequence and likelihood before performing the HAZOP Study [3,4]. The PHI is used to recognize hazards. The HAZOP Study evaluates the adequacy of the mechanisms in order to control hazards in plants [5-7]. These mechanisms include administrative procedures as well as hardware controls. This study identifies the ways of improving the operating efficiency by the assessment of 160 Piping and Instrumentation Diagram (P&ID) sheets.

As the quantitative assessment, three methods were used: Hazard identification (HAZID), frequency analysis (FA) and consequence analysis (CA). HAZID is defined by the method that describes guides as hazard assessment. Quantitative modeling method calculates precisely by selecting a list of possible failure cases. After identifying the potential hazards, the FA step estimates the likelihood of accidents [8,9]. The frequency is usually obtained from the analysis of previous accident experience [10-12]. When the accidents occur, the effects of result reveal their impacts for personnel, equipment, structure and environment in the plants. The CA step describes the expected damage of the potential accident. The estimation of the consequences for each possible event is used for some form of computer modeling [13,14].

3. Scope and Objective of the Gas Refinery Plants

The target of this study is to evaluate the safety of a gas refinery plant including the following facilities:

- Gas condensate unit
- Flash gas compression unit
- Mercaptan removal unit
- HC dewpointing unit
- Sulfur recovery unit
- Storage unit
- Fuel gas system

The objective of this analysis is to quantify the individual risk per annum (IRPA) of fatality to personnel working in a gas refinery, to assess the risk to the surrounding populations, and to consider the adequacy of risk management measures as prevention, detection, ignition, control, mitigation, escape and evacuation. The scope of this study is the onshore refinery, covering the following operational facilities: gas receiving facilities, eight finger slug catchers, two

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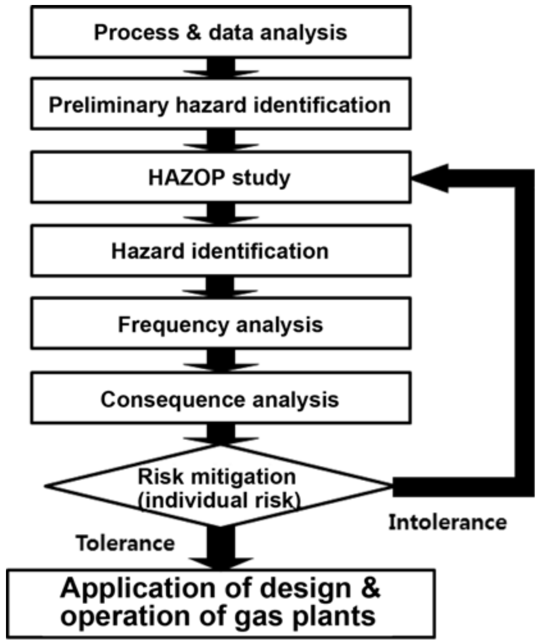

Fig. 1. Model for the integrated risk assessment procedur.

condensate processing trains, two gas treatment processing trains, two sulfur recovery units, import and export lines within the plant boundaries, gas export facilities and on-site product storage facilities.

This paper introduces an integrated risk assessment method to increase the safety of gas refinery plants. The integration method is efficient and systematic including both quantitative and qualitative risk assessment methods which are PHI, HAZOP Study, HAZID, CA and FA. Then we applied risk calculation and risk assessment to the same gas refinery plants for accident prevention.

PRELIMINARY HAZARD IDENTIFICATION

1. Function of PHI

This PHI step consists of two functions: 1) to enable recording the hazard controls which were already built into the design so that these are not removed at later stage without consideration to the safety impact, and 2) to promote identification of new and changed hazard controls which could be implemented in the design. This step considers pre-identified hazards such as potential undesired events, worst case potential consequences, safeguards and mitigation measures and many events. The scope of the PHI step is to identify hazards for using risk matrix with the normal operations of the gas refinery plants.

2. Hazard Identification Using Risk Matrix

Risk ranking of each hazard is expressed as a ranked matrix of frequency and severity in Fig. 2. The matrix is divided into three different types: 1) undefined region throughout the course of the

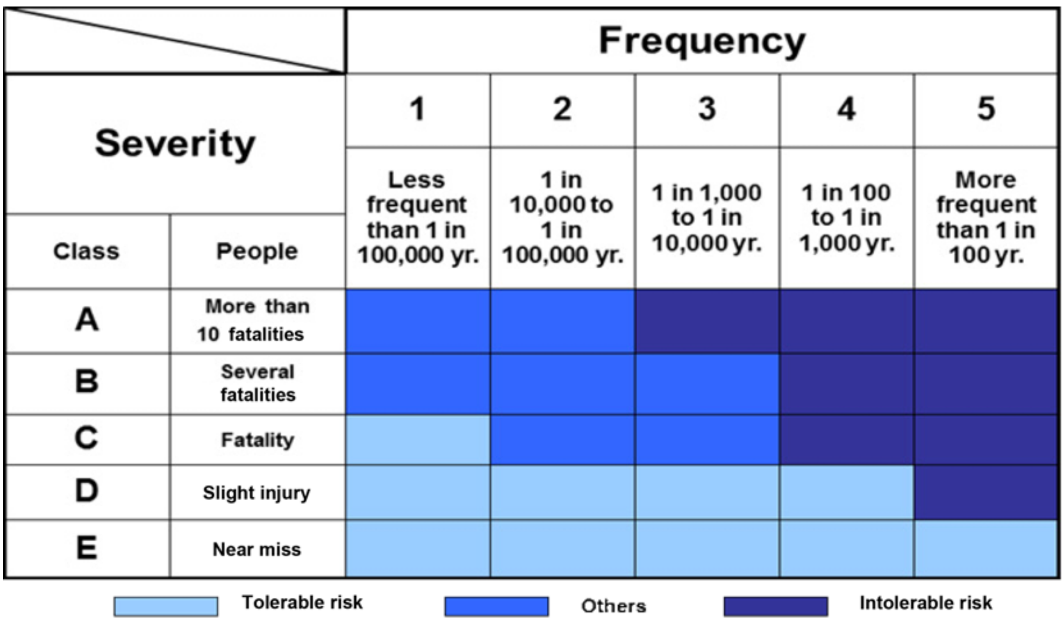

Fig. 2. PHI risk matrix of frequency and severity.

Table 1. Example sheet of hazard identification

Hazard	Severity class	Frequency rank
Gas/condensate reception - loss of containment	A	3
Condensate stabilization - loss of containment	C	4
Flash gas compression - fire	C	3
Mercaptan removed by molecular sieve - explosion	B	2
Fuel gas system - explosion	C	2

hazard, 2) the regions which usually represent levels of risk acceptability and define a course of action following the assessment, 3) and others. This step assumed that D and E degree of severity class did not analyze hazards in the work of HAZOP study and other steps. But D degree of severity class and 1 degree of frequency considered severe hazard.

Table 1 shows an example sheet of hazard identification which presents the hazard identification. Fig. 2 shows their risk ranking based on the risk matrix.

HAZOP STUDY

1. Implementation

The fundamental purpose of HAZOP study is to identify possible hazards and operability problems and concerns. It is not an analysis method to remove risk and mitigate consequences. It is not the responsibility of the HAZOP team members identifying engineering and procedural solutions to potential hazards identified during the HAZOP Study. Procedural implements and engineering modifications identified in a HAZOP study to mitigate risk gives assurance that the plant is operated under some lower level of risk. Primary hazardous chemicals of the gas refinery plants process are as follows in Table 2.

Table 2. Hazards of chemicals used in the gas refinery plants

Chemical	Hazards
Propane (C_3H_8)	Flammability
Hydrogen sulfide (H_2S)	Toxicity/Flammability
Methanol (CH_3OH)	Combustibility
Amine	Toxicity/Flammability
C_5+	Flammability/Combustibility

2. Guide Words and Procedure

A HAZOP study is a simple structured methodology for hazard identification. It is an investigation technique designed to inspire imaginative thinking by a team of experts to identify hazards and operational problems while examining a process and system. A HAZOP study involves a systematic and methodical examination of design documents that describe the facility. A multidisciplinary team to identify safety hazards and operability problems by evaluating deviations from design intents performs the study. HAZOP information shows the deviations results from the combinations of guide words and process parameters and any other consequences, and suggested actions to prevent or mitigate the consequences. Each work sheet has a banner at the top that identifies the area, unit and date of the study session, the node and parameter reviewed, and the design intention of the process parameter. An experienced team facilitator systematically steers the team through the system components identified in the design documents by using a set of guide words to identify deviations from the design intent of key parameters.

These guide words are used on specific, defined areas of the plant design that encompass a common set of process parameters that are relevant to the goals of the HAZOP study. A systematic method of identifying the nodes on the P&ID is used to focus the team's attention and to provide sufficient documentation. A HAZOP study

Parameter	Deviation	Causes	Consequences	Safety guards	S	O	Recommendations
Maintenance	Maintenance	- Insufficient access for removing CP-01	- Maintenance impossible			O	General recommendation : Accessibility for CP-01 to be checked by
Flow	Less flow	- Turn-down case - PV-16A/B fully open - During pigging operation - Leakage	- Maloperation of the plant - Gas release	- Possible to turn-down operation - PV-16A/B fail close - Fire water monitor		O	General recommendation : to provide enough gas detectors all over plant
Pressure	Vacuum	- Shut in and cool-down during steam out	- High vacuum condition	- None	S	O	General recommendation : Operating manual should be specifically reference steam out
Others	Maintenance	- Actual manway I.D	- Impossible entrance and exit of tray	- None		O	General recommendation : Actual diameter of column manway and structure packing should be checked for tray remover
Others	Human error	- Two relief valve inlet line block valve simultaneously close	- Relief blocking	- Lock open and lock closed	S		General recommendation : Provide interlock or operation procedure for safety valve to isolation in accordance with API 520
Others	Site gradient	- Reverse flow from the burn pit (30m higher)	- Fill in the vessel with burn pit liquid	- None		O	General recommendation : All burn pit lines shall have a check valve
Pressure	Higher pressure	- External fire	- High pressure	- PSV-11/12	S		General recommendation : Relief valve tail pipe should have drain hole as per common industrial practice

Fig. 3. Example of HAZOP study.

is that the process does not have inherent hazards and operating problems when the unit is operating with process parameters within ranges specified by the unit design. These parameters are defined by the basic design documents for the unit such as the process flow diagrams, P&IDs, equipment specifications, and operating procedures. A related premise for a HAZOP study is that operating and maintenance personnel follow basic safety rules.

This work has accomplished the general method and procedure of HAZOP study such as five steps with the results of the PHI step: 1) selecting nodes, 2) identifying hazard using guide words, 3) identifying possible causes and consequences, 4) evaluating existing safeguards (i.e., relief valves, anticipatory alarms, emergency shutdown (ESD) system, written procedures, conservative design, preventive maintenance, training, routine monitoring, redundant controls, environmental sensors and emergency response), and 5) recording the results of each node studied in the worksheets.

HAZARD IDENTIFICATION

HAZOP study generally deals with process related hazards, HAZID deals with the primary process, but also non-process, hazards. In this study, HAZOP study was carried out on the entire process to propose improvements. And then hazard management such as prevention, detection, and ignition, regarding the proposed improvements that correspond to the intolerable risk, are recommended in Fig. 4.

The identification of the hazards associated with the gas refinery plants is generally grouped into two main categories: process hazards and non-process hazards.

1. Process Hazards

Process hazards relate to accidental releases of toxic or flammable material, which can expose onsite and offsite persons, and result in a single or multiple fatalities. The method for identification of hazardous events essentially involves the systematic definition of failure cases for all items of equipment within those systems identified as having a major hazard potential. It covers essentially all the main processing and storage systems including associated pipe work.

The failure cases are defined by using a specific set of conditions to represent a range of possible conditions of failure. It is not practicable or necessary to consider every possible permutation of size and location of hole, exact inventory at time of failure, temperature and pressure and so on, since all of these can in practice vary continuously between certain limits. Thus, representative values of each parameter, which needs to be defined to model the failure case, are selected in such a way as to cover the spectrum of possible values.

Process systems will typically consist of many linked items of equipment and pipework, containing hazardous material. The process flow schemes and piping and instrumentation diagrams are used to identify normal inventories of hazardous materials for each process section. This is based on actual liquid hold-up data which is available or calculated from equipment size and function. Typically, three failure sizes are modeled for each subsystem:

- Catastrophic failures
- Large leaks: 100 mm hole diameter
- Medium leaks: 25 mm hole diameter

2. Non-process Hazards

Non-process hazards, such as occupational and transportation risk, relate to specific activities such as falling down a staircase, burns suffered from contact with hot equipment, being exposed to dropped objects ,and road traffic accidents.

The quantification of the risk to personnel from non-process hazards has been addressed in the analysis of occupation and transportation hazards.

FREQUENCY ANALYSIS

1. Classification of Failure Cases

Failure cases are developed by following the hazard identification. Failure cases are all possible discrete failures resulting from the operation of process systems. These are determined by breaking the plant up into sections that would show similar behavior upon release as the terms of material, temperature, pressure, discharge

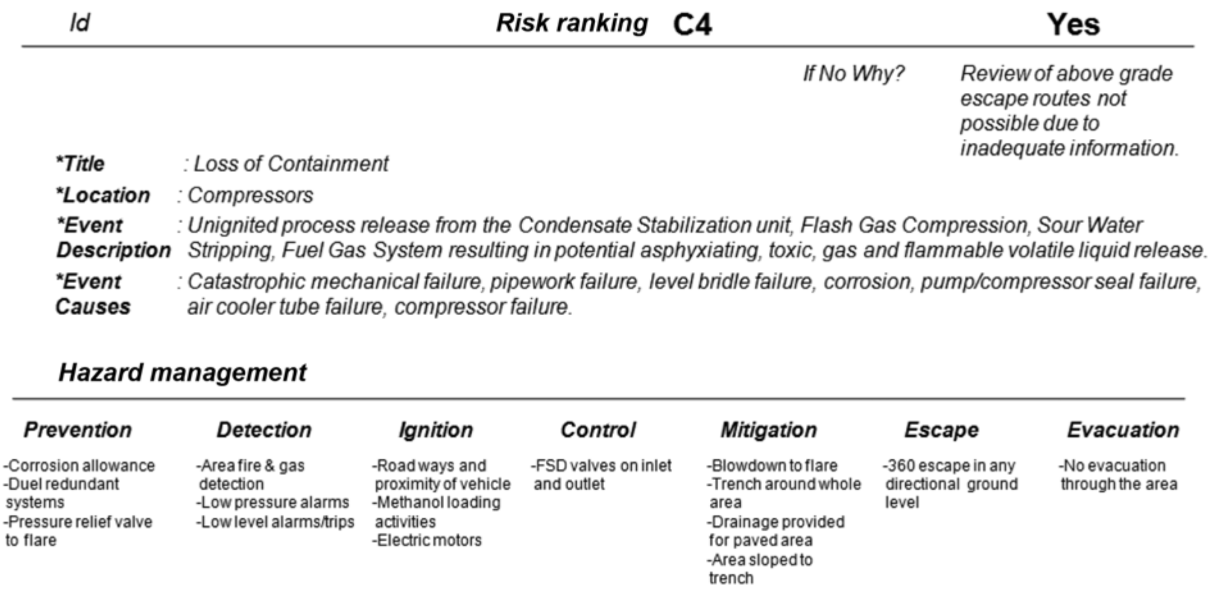

Fig. 4. Example of hazard identification.

rate and dispersion characteristics.

The dispersion characteristics of the release determine the nature and frequency of the final end event such as jet fire, explosion. flash fire, or toxic cloud. The failure cases are required to model the consequences and frequency of the release. Each failure case is assigned by unique identification number, pressure and temperature data, leak size, leak flow rate, equipment list and contributing towards the leak frequency. This step calculates the overall leak frequency of the equipment contributing for each size of hole. The combination of the release frequency with the event selection and probabilities of the frequencies are calculated.

2. Failure Case Definition

The overall failure frequency for each failure case has been calculated by counting the type and number of equipment items contributing to each failure case. Inspection of the tables reveals the technique applied in the estimation of the leak frequencies for a failure case. By combining the process equipment count with the failure rate per equipment item, the overall failure case frequency has been estimated. This study has some definition of the failure case as temperature, pressure, material, phase, mass, and duration. For example, temperature is defined by the conditions of release and process in centigrade degree. Pressure is defined by the conditions of release and process in bar gauge. Phase is defined by the material where it is in the process and storage such as vapor, two phase and liquid. Duration is usually defined based on a review of the time to failure, detection, response, isolation and blow down of the release source.

CONSEQUENCE ANALYSIS

1. Introduction to the Modeling

An analysis of the consequences of releases forms the precursor to the risk analysis. Such an analysis does not consider the frequency of events; it merely determines the scale of flammable, explosive or toxic releases. The consequence analysis has used PHAST program as consequence modeling software.

Before flammable and toxic effects are modeled, it is necessary to establish how a material would behave upon release. These are taken various forms ranging from a liquid spill to an aerosol or a vapor release. The release behavior and dispersion characteristics are dependent on:

- Physical properties of the material involved
- Hole size or release rate
- Release pressure and temperature
- Release phase
- Ambient conditions
- Nature of the surroundings

Here, ambient conditions include humidity, weather class, wind speed and temperature of ground and air. The surrounding land includes tall buildings, trees, increase air turbulence, thus encouraging more rapid air entrainment and dilution of a dispersing vapor cloud.

2. Ignition Modeling

The likelihood and consequences of a release igniting are central to the evaluation process undergone in this study.

There are two key elements to the approach taken:

- Immediate ignition: for each release a probability of immediate ignition is assigned, the appropriate consequence model is then employed, for example jet fire, pool fire, BLEVE, etc.
- Delayed ignition: all delayed ignition sources show time dependent behavior, where the probability of ignition increases with the duration that a flammable vapor cloud between the upper and lower flammable limits is in contact.

The immediate and delayed ignition probability has a direct impact on the frequency at which flammable consequences manifest themselves as the ignition probability directly affects the risk results. The time to ignition also has an impact in terms of the type, size and location of potential consequences.

FA was performed by focusing on a set of primary risk factors and equipment obtained from HAZID analysis of an entire process of interest. CA was performed with an emphasis on primary risk factors and equipment of high risk ranking. A set of scenarios was built on the basis on HAZID analysis (locations, events, etc.) when performing FA and CA.

RISK RESULTS

1. Introduction of LSIR and IRPA

The risks are presented results of gas refinery plants in the form of location specific individual risk (LSIR) contours and individual risk per annum (IRPA) for an average onsite operator.

The location specific individual risk is defined as the level of risk an individual would be exposed to present in a particular outdoor location for a whole year. Most commonly, an LSIR criterion level of 1×10^{-6} per year is used which is not exceeded in residential population areas and land zoned for residential development, whereas 1×10^{-5} per year is commonly used as the criterion level to exceed at neighboring industrial developments.

The individual risk per year is normally used to assess onsite risk to individual workers. IRPA is defined as the risk to an individual per year as a result of activity. IRPA for process related hazard takes into account the amount of time a person actually spends on-site in the various areas of the gas refinery plants. This study calculates that individual risk is the number of people at risk over number of fatalities per year.

2. Result of LSIR

The LSIR contours in Fig. 5 are a graphical representation of the risk to people from the process hazards associated with the gas refinery plants.

The highest risk levels on the refinery are centered over the gas/condensate process areas as evident from the red one-in-ten-thousand (1×10^{-4}) contours. The higher risk levels are a reflection of the higher concentration for process equipment presenting potential leak sources. This study is combined with the higher concentration of potential ignition sources causes to the location with the highest risk.

Within the gas treatment units, this area, the gas treatment area, where hydrogen sulfide is removed from the gas stream before being directed to the sulfur recovery units. The gas regeneration column and the line present a toxic vapor risk. This is particularly evident from the 1×10^{-5} per year risk contour extending to the west.

The 1×10^{-6} per year contour is centered on the gas/condensate processing area, indicating that these areas most contribute towards this risk contour.

The 1×10^{-7} outer risk contour is characterized by process related

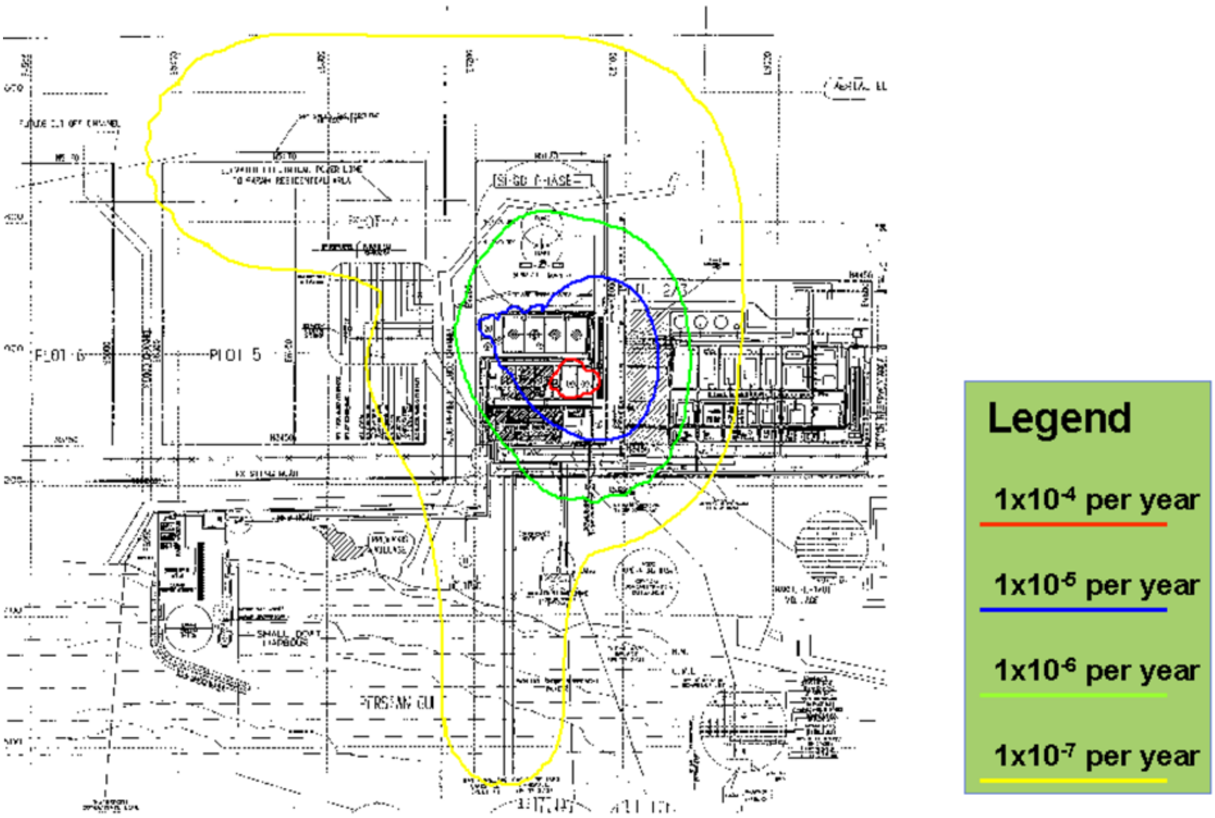

Fig. 5. Location specific individual risk per annum contours.

Table 3. Main process risk by unit

Process area/Unit	Percent of total risk (%)
Gas inlet pipeline and slugcatcher	42.0
Gas/condensate unit	18.2
Tank farm	14.6
Sulphur recovery unit	8.0
Condensate stabilizer unit	3.2
Mercaptan removal unit	1.8
Total	100

hazards in the raw gas import pipeline and the gas export pipeline. The condensate export pipeline generates risk levels below 1×10^{-7} per year and hence does not feature. For the gas import and export pipelines the risk levels are less than one-in-a-million (1×10^{-6}) per year, and hence the one-in-a-million per year risk contour does not feature for these lines.

3. Results of Main Process Risk

Table 3 presents the main risk of the various operational areas makes towards process related fatality risks. The risk ranking results confirm that the slug catchers and the gas/condensate process unit are the main contributors toward risk due to toxic vapor risks associated with hydrogen sulfide.

4. Results of IRPA

The overall individual risk per year for each work group shows the result of occupational, transport and process related risks as Fig. 6. The individual risk is highest for the maintenance personnel due to their higher exposure to process and occupational hazards.

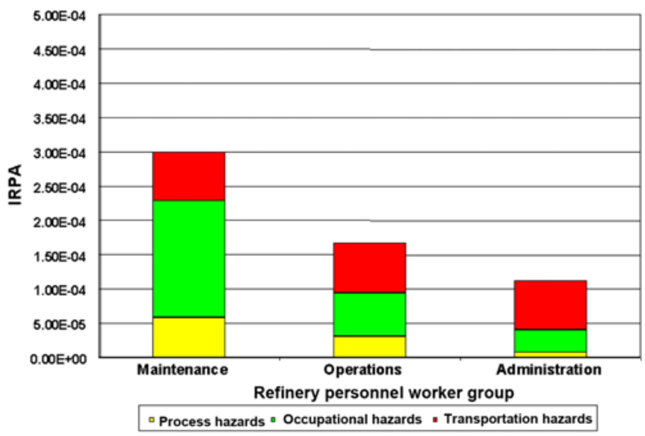

Fig. 6. Result of individual risk per annum for each work group.

The risk to individuals of each worker group has been quantified for the following hazards:

- Process Hazards
- Occupational Hazards
- Transportation Hazards

In the absence of any existing risk criteria for the gas refinery plant criteria for individual risk per year for on-site workers that the maximum tolerable risks of 1×10^{-3} per year for on-site workers, whereas the risk may be regarded as negligible if less than 1×10^{-5} per year.

The IRPA results are between a maximum of 3.15×10^{-4} per year for maintenance personnel and 9.34×10^{-5} per year for office staff.

The IRPA results fall within the region where risk reduction measures are desirable if economically viable and practicable.

CONCLUSION

The aim of this work is to manage risks properly and to propose a systematic procedure of applying quantitative and qualitative hazard identification methods to remove risk and applied to local gas refinery plants before construction. Results of this research provide the input form to the design and safety management of the installation. This work provides an assessment if there is no residential population within the bounds of the 1×10^{-6} per annum risk contour. Therefore, the risk to the public from the gas refinery plants should be considered acceptable. Lastly, the proposed integrated system of qualitative and quantitative risk assessment methods for gas refinery plants may be helpful for reducing budget and time, identifying most of important hazard using a systematic method and performing a complete and rigorous analysis.

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