Studies in support of a quantitative approach to hazardous area classification

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A study was made of the feasibility of putting hazardous area classification (HAC) on a more quantitative basis.

A review of current HAC practice showed that the widespread policy of setting fixed zone distances around sources of hazard was subjective and sometimes led to inconsistencies between different codes of practice when applied to the same situation. Fatality and injury statistics were used to show that there is a significant risk to workers from the ignition of flammable atmospheres, which should be reduced.

Data were researched and compiled to fit into a proposed framework for the quantification of HAC. These included information concerning leak source inventory; source leak frequency; and source leak size distribution. Mathematical models were collected which could be used to describe the emission and dispersion of flammable releases. Example calculations were performed for typical leak scenarios to illustrate the variation in hazard distances.

Estimates were made of the ignition and explosion probabilities of flammable leaks, which depended principally on emission size.

To compensate for uncertainties in the researched data, a fire and explosion model was devised to estimate the ignition frequency on a typical process plant. The model was applied to a "standard" plant which was formulated from researched data. By iteratively checking the estimated ignition frequencies against historical data it was concluded that reasonable agreement was achieved with some adjustment of the input data.

The special problems of HAC of indoor plants were also addressed.

It was concluded that the results of this study provided a basic framework for the quantification of HAC, although the quality of currently available data necessary for quantification is generally poor. The acquisition of better quality leak and ignition data should provide a platform from which the current work may progress. Further work should include the further refinement of the basic fire and explosion model to account for ignitions which HAC cannot protect against such as autoignitions. It was also noted that the behaviour of indoor releases requires clarification, together with the concept of a minimum flammable inventory below which there is negligible risk of ignition.
Studies in Support of
A Quantitative Approach to
Hazardous Area Classification

by

Andrew William Cox

A Doctoral Thesis
Submitted in Partial Fulfilment of the Requirements
for the Award of

Doctor of Philosophy

of the Loughborough University of Technology

1989

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To my mother Mrs Agnes Cox and to the memory of my father Mr William Thomas Cox

And to Patricia and William
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1. Introduction

Hazardous area classification (HAC) is a method used to control the movement of ignition sources in process areas where flammable atmospheres may occur. As the use of the method has grown more widespread the criteria applied in different industries for HAC are interpreted in a very subjective manner. This has resulted in significant inconsistencies in the overall approach adopted by various codes of practice, and thus the need for a more rational basis for HAC has been identified. This study set out to firstly review the current approach taken by the process industries to HAC, with a view to developing a rationale based on a quantitative approach. The main objective of this study was to develop and define an overall methodology within which the quantitative methods may be implemented.

This study approached the problem by firstly considering the current performance of the UK codes of practice in present use. This was achieved by analysing accident statistics involving the ignition of flammable substances. The study then proposed a modular approach to quantitative HAC and identified the individual basic units necessary for such an approach. These included data concerning:

- Leak source type
- Leak source inventory per plant
- Leak frequency
- Leak size
- Mathematical models for emission size and dispersion
- Ignition probability

Data for all these headings were researched and assembled to arrive at an outline procedure for assessing the risk presented by a particular release with regard to the ignition probability. To compensate for deficiencies in the researched data it was necessary to devise a model that could provide crosschecks on the input data selections made in terms of intermediate and final results. The model, termed here the fire and explosion model, provided estimated
overall ignition and explosion frequencies for a given plant leak source inventory. The overall results were further broken down into the contributions made by leak source type and fluid phase. The estimated frequencies were then checked against statistical data.

The special problems of HAC of indoor plants were also addressed.

Finally, a programme of future work was drawn up. This included the refinement and validation of the fire and explosion model. Also included was a detailed description of how a refined version of this model may be used for HAC.
2. Background

2.1 Development of philosophy and codes of practice

Historically, the bulk of the work concerning the control of hazardous areas has been done by the electrical engineering profession. This has resulted in an emphasis on the design of electrical equipment to be used in a hazardous area rather than on the methods used to classify the area itself. It has also resulted in an emphasis on the hazards posed by sparking electrical equipment as potential sources of ignition compared to other sources such as hot surfaces or naked flames. Note that although the term "hazardous area" implies a two-dimensional space it is used in HAC, and in this study, to describe the three-dimensional space around a leak source. The history of HAC has been described by Palles-Clarke [1], and is summarised below.

The first national code concerning the design and installation of electrical equipment in potentially flammable atmospheres was published in 1922 and was for use in coal mines [2]. It was realised that similar flammable hazards existed on oil drilling installations and in 1946 the Institute of Petroleum produced the first code which defined areas as either safe, dangerous or remotely dangerous. A safe area was one where a flammable atmosphere could not exist; a dangerous area was one where such an atmosphere was likely to occur. In a remotely dangerous area the occurrence of a flammable atmosphere could be prevented by taking proper precautions and so would only happen under abnormal circumstances. Remotely dangerous areas initially applied only to oil fields, although in 1956 the Institute of Petroleum enlarged the category to encompass all situations involving the handling of hydrocarbons. Note that this code was specific to the petroleum industry only. A general code encompassing such concepts for application in other industries confronted with similar problems was yet to be established.

The classification of remotely dangerous areas formed an important part of the Institute of Petroleum Model Codes published between 1964 and 1967. These codes are a landmark in HAC: due mainly to their scope and content they are still widely adopted by international
designers and contractors. The codes divided areas into either safe or dangerous and further sub-divided dangerous areas in terms of the likelihood of a flammable atmosphere being present as follows:

Division 0 - An area in which a dangerous atmosphere is continually present
Division 1 - An area in which a dangerous atmosphere is likely to occur under normal operating conditions
Division 2 - An area in which a dangerous atmosphere is likely to occur only under abnormal conditions.

The Institute of Petroleum Model Codes were the first to set standard division sizes and types around individual sources of release with the use of diagrams. This is a practice that has been widely adopted both nationally and internationally and is discussed further in Section 2.2.

The overall philosophy of the Institute of Petroleum Model Code was adopted nationally in the 1960's upon the publication of the revised British Standard CP 1003 in three parts. In the 1970's there was mounting criticism of the procedure whereby a standard division size was specified around a potential source of release irrespective of its type or the material being handled. In 1972 the International Electrotechnical Committee published report IEC 79-10 [3] which proposed that each source of release be graded according to the likelihood of a release occurring. This is commonly known as the source of hazard philosophy for HAC. Also, IEC 79-10 proposed that the hazardous area around the source of release be divided into zones, roughly equivalent to the aforementioned divisions. These were defined as follows:

Zone 0 - In which an explosive gas-air mixture is continuously present or present for long periods
Zone 1 - In which an explosive gas-air mixture is likely to occur in normal operation
Zone 2 - In which an explosive gas-air mixture is not likely to occur and if it occurs it will only exist for a short time.

IEC 79-10 also stipulated that the physical properties and release quantity of the flammable material in question be taken into account when setting zone distances around each source. However, there was no guidance as to how this may be achieved, and there were no definitions of what constituted normal or abnormal behaviour.
Despite its shortcomings IEC 79-10 proved very influential, providing the basis for the British Standard for electrical installations in hazardous locations BS 5345, the first part of which was published in 1976 [4]. Part 2, dealing explicitly with the classification of hazardous areas, was published in 1983 [5] and forms the basis of current UK practice. However, the way that BS 5345: Part 2 should be applied was left to the user. Qualitative guidance was given to how release sources may be graded according to the likelihood of release and thus the type and size of zone required for each. Examples were also given of how HAC tables describing each leak source may be drawn up, and how the final HAC diagrams may look with the calculated zones superimposed on the plant layout. However, there was a complete lack of guidance on which quantification methods may be used to set zone distances and types. It was generally felt that such a shortcoming on what was ostensibly a quantificational methodology seriously limited the use of BS 5345: Part 2. The formulation of BS 5345: Part 2 is discussed further in Section 2.4.1.

2.2 Current standards and codes

The current approach taken by national and international process industry operators to HAC can best be summarised by considering some standards and codes of practice used. A list of UK and overseas standards and codes is given in Appendix A. Table 2.1 outlines the approach taken by each UK code to setting zone distances and types, whilst Tables 2.2 and 2.3 consider respectively overseas codes and codes specific to offshore and onshore oil drilling operations. The summary shows that most of the standards and codes of practice in use at the present time either do not or only partly employ the source of hazard philosophy. Tables 2.1 - 2.3 show that the commonly adopted method of setting zone sizes is by using fixed distances suggested by the codes. Usually the codes include a list of diagrams showing plant items with the suggested zone, defined by geometry and distance, superimposed on each. The only wholly quantitative method for HAC is given by British Gas standard BGCPS/SHA1. However, this standard cannot be globally applied. It is only concerned with non-domestic methane applications, resulting in considerable simplification in the range of process systems to be considered and also the calculations necessary to model possible leaks.

The ICI/RoSPA code should also be highlighted as it has been a major reference source for HAC since its publication. It incorporates the guidelines laid down by IEC 79-10 by using the source of hazard philosophy and accounts for the difference in physical properties between
flammable substances by tabulating a range of fixed zone sizes for each source of release. However, it cannot be globally applied to all release sources (e.g., it cannot deal with flammable pools) or all flammable materials (e.g., hydrogen or solvents). It is understood that the ICI HAC code of practice is currently under revision, and is to embody the principles of BS 5345 part 2. Data from a draft of the revised code are presented in later sections. For the rest of the codes in Tables 2.1 - 2.3 where fixed zone distances are suggested these are either intended specifically for use with one substance (usually LPG) or for the typical range of substances encountered on process plant. In many cases the suggested distances are minimum values which are to be adapted according to individual situations. However, there is usually no indication of what the minimum case represents in terms of leaking fluid and leak size, and no indication as to how the suggested distances may be scaled for individual situations. These represent clear inadequacies in the approach which must be considered if the codes are to be applied in a more rigorous manner.

Table 2.4 lists a selection of codes from Tables 2.1 - 2.3 which use the suggested fixed distance approach. Under each code equipment items are listed which are regarded as sources of release and zoned using this method. The table is not exhaustive for each code but seeks to show that although the range of equipment covered is wide, there is an appreciable amount of overlap between codes. By considering the overlap between the codes in Table 2.4 the significant inconsistencies in the suggested fixed distance approach thus become clear. Figure 2.1 shows how some of the selected codes zone an LPG pump. There is a large difference between the largest and smallest zone (the codes which deal specifically with LPG are indicated). Similarly, Figures 2.2 and 2.3 show the result of applying the codes to an LPG flange and a general purpose fixed roof storage tank respectively. Figures 2.1 - 2.3 show that there is apparently a large discrepancy between the rationales used in each code to set zone sizes. It is suspected that in most cases the suggested distances are derived from subjective judgement on the part of the code developers rather than being based on theoretical considerations. It is therefore very difficult to rationally choose a code or standard to zone an installation when by all accounts the codes listed are supposedly equally valid, although their application may produce significantly different results. This has serious economic implications, since the imposition of very large zones is expensive. This may be expressed in terms of the added cost of the ignition-protected electrical equipment installed within the zone and also in terms of the change in working practices necessary within a zone e.g., if a zone overlaps a road. The operator’s interest is therefore served by limiting zone sizes without compromising the overall level of safety.
2.3 Discontent with current approaches

The current approach to hazardous area classification outlined in Section 2.2 is generally felt to be adequate as far as most code users are concerned. The suggested fixed distances provide a practical design method for zoning because they are quick and easy to apply. Engineers employed by large concerns with proprietary HAC codes that are adapted from national codes believe and are satisfied that the codes are applicable to their own activity and that the empirical nature of the distances reflects good practice. It is felt that although zone distances for a given situation may vary from code to code this only indicates varying degrees of conservatism which are deemed appropriate.

However, other engineers feel that the fixed distance approach is inadequate. They argue that although it is superficially attractive because of its simplicity in application it may not be appropriately applied unless some degree of theoretical consistency is achieved. It is also reasoned that the degree of conservatism supposedly incorporated into each zone is meaningless because the zone distances themselves vary enormously from code to code as illustrated in Figures 2.1-2.3. Sometimes the application of a code gives a result which engineering judgement suggests is wildly conservative, but which has to be implemented in spite of cost. This is particularly relevant to indoor locations, where even the presence of a small-bore pipe fitting may result in the flameproofing of the entire enclosure, regardless of its size. The codes may also give guidance which is inconsistent with the actual situation. An example is the need for protected electrical equipment in a boilerhouse containing both pipe fittings and naked flames. In this case it is obvious that protecting against the ignition of a pipe leak by using increased safety electrical equipment is pointless as the boiler flame represents a continuous ignition source. However this is often not accounted for by the relevant code of practice. Lastly, there is a special problem for the Health and Safety Executive (HSE) which is charged with administering zoning practice in the UK. With such a multiplicity of HAC codes available it is difficult to assess individual proposals, some of which are on the borderline of accepted practice.

The critics of the fixed distance approach felt that a more rigorous methodology for HAC was necessary to remove the above anomalies which would be based on quantitative evaluations such as emission and dispersion calculations. This would also make the methodology more adaptable and thus resulting in consistent application to different types of plant. The progress toward a quantitative method of HAC is described below.
2.4 Progress towards quantification

2.4.1 The BS 5345: Part 2 draft Appendix

The first major attempt toward the quantification of HAC was the formulation of the source of hazard philosophy as described in IEC 79-10 (see Section 2.1). This concept was further promulgated at a series of Institution of Electrical Engineering meetings in the 1970's and 1980's [6][7][8][9]. However, although there was general agreement that the source of hazard method was the way to progress there was no clear idea of how it should be applied. The type of approach usually envisaged is the procedure whereby firstly the leak frequency and leak size of a source is determined. The emission, vaporisation (in the case of a liquid), dispersion and ignition of the leak are then modelled and the results assessed against an absolute risk criterion such as the fatal accident rate (the FAR is the fatality rate for an activity per $10^8$ exposed hours [10]).

An attempt was made to formalise this procedure during the drafting of BS 5345: Part 2. An appendix was written containing a quantification procedure for sizing and classifying zones in a similar way to that outlined above. The appendix complemented the philosophy laid out in the main text and contained worked examples. However the draft appendix attracted heavy criticism [11], resulting in its deletion from the final document. Many of the criticisms were of a trivial nature e.g. errors in the worked examples. Other more valid criticisms were that the calculational models used were inadequate for the task or that assumptions made for the calculations were unjustified. More pointedly, it was argued that the results obtained were similar to the equivalent suggested fixed distances but involved considerable time and resource. All this seemed to be missing the point that the principal benefit of the source of hazard method is that it provides a deeper insight into potential hazards than can be gained by simply assigning zone distances according to a set of diagrams. Despite this, the arguments presented reflected a concern for ease of method application since some quantification process is required, rather than a rejection of the fundamental source-of-hazard concept.

2.4.2 The IIGCHL Loughborough study

Following the publication of BS 5345: Part 2 in 1983 a committee was set up which drew its members from the institutions interested in HAC to review and provide guidance in future
HAC strategy formulation, and which was consequently called the Inter-Institutional Group on the Classification of Hazardous Locations (IIGCHL). One of the objectives of the group was therefore to reassess the approach taken to HAC, in particular to consider the possible incorporation of quantification in an overall framework which is applicable to all situations where flammable atmospheres may arise. This led to the funding of a three year study at Loughborough University of Technology with the results providing the basis of this thesis.
Table 2.1 Approaches adopted by UK guidance and codes of practice to hazardous area classification

<table>
<thead>
<tr>
<th>Title</th>
<th>Publisher</th>
<th>Approach</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electrical Installations in Flammable Atmospheres. Vol. 1.5 (1975)*</td>
<td>ICI/RoSPA</td>
<td>Fixed distances which vary according to type of substance.</td>
</tr>
<tr>
<td>Area Classification (1986)*</td>
<td>BP Chemicals</td>
<td>Fixed distances supported by calculational method.</td>
</tr>
<tr>
<td>Hazardous Area Classification: A Guide for Fire and Explosion Inspectors (1986)</td>
<td>Health and Safety Executive</td>
<td>Based on ICI/RoSPA code and supported by calculational method</td>
</tr>
</tbody>
</table>

continued
<table>
<thead>
<tr>
<th>Table 2.1 continued</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>The Storage of Highly Flammable Liquids CS 2 (1978)</strong></td>
</tr>
<tr>
<td>Health and Safety Executive</td>
</tr>
<tr>
<td><strong>The Keeping of LPG in Cylinders or Small Containers CS 4 (1981)</strong></td>
</tr>
<tr>
<td>Health and Safety Executive</td>
</tr>
<tr>
<td><strong>The Storage of LPG at Fixed Installations CS 5 (1981)</strong></td>
</tr>
<tr>
<td>Health and Safety Executive</td>
</tr>
<tr>
<td><strong>Vehicle Finishing Units: Fire and Explosion Hazards PM 25 (1981)</strong></td>
</tr>
<tr>
<td>Health and Safety Executive</td>
</tr>
<tr>
<td><strong>Bulk Storage and Handling of High Strength Potable Alcohol (1986)</strong></td>
</tr>
<tr>
<td>Health and Safety Executive</td>
</tr>
<tr>
<td><strong>Highly Flammable Liquids in the Paint Industry (1978)</strong></td>
</tr>
<tr>
<td>Health and Safety Executive</td>
</tr>
</tbody>
</table>

*indicates that the document referred to is currently under revision. The revised document may employ a different approach to HAC than that given here.
Table 2.2 Approaches adopted by selected overseas codes of practice to hazardous area classification

<table>
<thead>
<tr>
<th>Title</th>
<th>Origin</th>
<th>Approach</th>
</tr>
</thead>
<tbody>
<tr>
<td>Guidelines for the Prevention of Danger in Explosives Atmospheres with Examples (1975)*</td>
<td>Berufsgenossenschaft der Chemischen Industrie (W. Germany)</td>
<td>Suggested fixed distances.</td>
</tr>
<tr>
<td>Code for Electrical Installations in Locations with Fire and Explosion Hazards 64-2</td>
<td>Comitato Elettrotecnico Italiano</td>
<td>Suggested fixed distances.</td>
</tr>
<tr>
<td>------------------------------------------------------</td>
<td>--------</td>
<td>---------------------------</td>
</tr>
<tr>
<td>Guidelines for the Classification of Hazardous Areas in zones in relation to Gas Explosion Hazards and to the Operation and Selection of Electrical Apparatus R No. 2 (1979)</td>
<td>Netherlands Directorate-General of Labour</td>
<td>Suggested fixed distances based on given leak rates.</td>
</tr>
</tbody>
</table>

* indicates that the document referred to is currently under revision. The revised document may employ a different approach to HAC than that given here.
Table 2.3 Approaches adopted by selected UK and overseas guidance to hazardous area classification of offshore installations

<table>
<thead>
<tr>
<th>Title</th>
<th>Origin</th>
<th>Approach</th>
</tr>
</thead>
<tbody>
<tr>
<td>Classification of Areas for Electrical Installations at Drilling Rigs and Production Facilities on Land and on Marine Fixed and Mobile platforms API 500B (1973)</td>
<td>American Petroleum Institute</td>
<td>Suggested fixed distances in association with API 500A.</td>
</tr>
</tbody>
</table>
Table 2.4 Plant items and equipment covered by hazardous area classification codes of practice and guidance

<table>
<thead>
<tr>
<th>Item</th>
<th>Code or guidance note (see below for key)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>1</td>
</tr>
<tr>
<td>Heavier than air source</td>
<td>x</td>
</tr>
<tr>
<td>Lighter than air source</td>
<td>x</td>
</tr>
<tr>
<td>As above - enclosed and ventilated</td>
<td>x</td>
</tr>
<tr>
<td>Storage tank - fixed roof and floating</td>
<td>x</td>
</tr>
<tr>
<td>Roof</td>
<td></td>
</tr>
<tr>
<td>Relief valve/vent</td>
<td>x</td>
</tr>
<tr>
<td>Sample point/drain</td>
<td>x</td>
</tr>
<tr>
<td>Flanges/connections</td>
<td>x</td>
</tr>
<tr>
<td>Tanker loading</td>
<td>x</td>
</tr>
<tr>
<td>Separators</td>
<td>x</td>
</tr>
<tr>
<td>Agitator gland</td>
<td></td>
</tr>
<tr>
<td>Drum or small container filling</td>
<td>x</td>
</tr>
<tr>
<td>Pumps/compressors</td>
<td>x</td>
</tr>
<tr>
<td>Aircraft refuelling</td>
<td>x</td>
</tr>
<tr>
<td>Paint spraybooth</td>
<td>x</td>
</tr>
<tr>
<td>Petrol store</td>
<td></td>
</tr>
<tr>
<td>Open liquid surfaces</td>
<td>x</td>
</tr>
<tr>
<td>LPG store</td>
<td>x</td>
</tr>
<tr>
<td>Garage/service station</td>
<td>x</td>
</tr>
</tbody>
</table>

continued
Table 2.4 continued

Drilling operations

<table>
<thead>
<tr>
<th>Item</th>
<th>Code or guidance note</th>
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<tbody>
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<td></td>
<td>1</td>
</tr>
<tr>
<td>Wellhead equipment</td>
<td>x</td>
</tr>
<tr>
<td>Base of drill</td>
<td>x</td>
</tr>
<tr>
<td>Mud tank</td>
<td>x</td>
</tr>
<tr>
<td>Shale shaker</td>
<td>x</td>
</tr>
<tr>
<td>Christmas tree</td>
<td>x</td>
</tr>
<tr>
<td>Pumped well</td>
<td>x</td>
</tr>
<tr>
<td>Pig launching/collecting point</td>
<td>x</td>
</tr>
</tbody>
</table>

Key to column labels

1. API 500A (US)
2. API 500B (US)
3. NFPA 30 and 70 (US)
4. MODU code (IMO)
5. TN B302 (Norway)
6. IP Code Part 1 (UK)
7. IP Code Part 8 (UK)
8. HSE Guidance notes (UK)
9. SS 421 08 20 (Sweden)
10. R No. 2 (Netherlands)
11. CCP Guidance notes (France)
12. BCG Guidance notes (W.Germany)
13. CEI 64-2 (Italy)
14. ICI RoSPA Vol 1.5 (UK)
15. ASA 2430.3 (Australia)
FIGURE 2.1 SUGGESTED FIXED ZONE DISTANCES AROUND AN LPG PUMP

SCALE 0 ———— 10 M
Figure 2.2  Suggested fixed Zone 2 distances around an LPG flanged joint
FIGURE 2.3 SUGGESTED FIXED ZONE DISTANCES AROUND A FIXED ROOF STORAGE TANK

SCALE 0 10 M

ZONE 1 ZONE 2
3. Objectives and strategy of study

The overall objectives of the study as laid down by the IIGCHL were as follows [12]:

"To assist the IIGCHL to suggest improvements to the UK approach to hazardous area classification and to provide background information for those involved in contributing to codes of practice in this area. The concept underlying the investigation is that the definition of hazardous areas may benefit from a quantitative approach."

Essentially this involved considering the components of the procedure outlined in Section 2.4.1 and reviewing the quantitative methods that may be applied to each of them.

It was recognised that the approach taken to quantification by the BS 5345: Part 2 draft appendix was fragmentary and not coherent. One of the prime objectives of the study was the construction of a comprehensive strategy for HAC into which data or model units could be fitted in a modular i.e. a building-block fashion. The advantage of this type of approach is that its components can be modified without affecting the overall strategy. For example, if a calculational model is found to be inadequate then it should be possible to substitute it directly without affecting the way that the models interact. Also, this approach enables the type of data needed to be precisely defined, so that if it is not available then at least the task of acquisition is clearly specified.

As a starting point, the flammable risk on outdoor plant was first considered. This was because outdoor plants are of similar construction i.e. they are comprised largely of pipework, vessels and pumps with the flammable inventory enclosed and probably pressurised. The leak sources are therefore relatively easy to define. Indoor plants are very much more diverse in nature and are less easy to typify. The problems of zoning indoor plant are addressed in Section 13. Notwithstanding these considerations, one prime motivation in formulating a methodology for HAC is the need for it to be applicable to as many situations as possible. In particular it is
important that the defined method is not dependent on the nature or scale of the process. Whilst the actual definition of a method for HAC was considered outside the scope of the study, this consideration underpinned the strategy adopted for the work described here. This was as follows:

1. The performance of current HAC practice was reviewed by way of examining the number of injuries and fatalities due to the ignition of flammable atmospheres. The selection of a suitable criterion for measuring the risk resulting from flammable releases was considered.

2. Items of equipment likely to leak were identified together with likely leak scenarios. Estimates were made of the numbers of these leak sources which may be found on typical process plant, to produce a plant "leak source profile". An estimate was then made of the number of plants at risk from fire and explosion, with a view to cross-checking incident data with the ignition model described below.

3. The equipment leak frequencies were investigated using data sources. The usefulness of confidence limits was considered, in view of the wide spread of values in the researched data. The need for a method of cross-checking frequency data was highlighted. Leak size distribution data were also researched.

4. Sets of leak emission and dispersion models were specified according to the likely leak scenarios. This was to investigate the effect of leak scenario and fluid physical properties in setting hazard distances using a representative selection of flammable fluids.

5. The probability of ignition of a given release was considered. Ignition frequency data were used to construct a basic model to predict the probability that a given release size will ignite or explode.

6. The equipment inventories, leak sizes, leak rates and ignition probabilities were condensed and incorporated to produce an approximation of a typical "standard plant". A fire and explosion model was then formulated which examines the contribution made by each equipment type and size to the overall risk of fire and explosion. The
model breaks down the overall ignition frequency for the plant into the contribution by fluid phase and equipment type. In this way some idea may be gained as to which plant items require the most protection against ignition.

7. The special problems attached to the application of the foregoing approach to indoor plants were considered, and related to the ventilation of such installations. A basis for an alternative approach was proposed to account for the different characteristics of indoor releases.

8. Lastly, further work was outlined. This included a rationale for the application of the fire and explosion model to HAC.

Steps 1 - 8 are summarised in Figure 3.1, which shows how the researched data "modules" fit into the overall strategy of the study. It is concluded in Section 15 that one of the principal aspects of future work leading on from this study should be the supplement and refinement of these modules.
Figure 3.1 Diagram of project strategy showing assembly of researched data.
4. Current performance

Before proposing improvements to the existing approach to hazardous area classification it was necessary to establish the current performance of existing standards and codes in preventing the ignition of flammable atmospheres. It was possible to measure this against two criteria: damage to people, and damage to property.

4.1 Fatalities and injuries

There are historical data available concerning the number of fatalities and injuries which are due to fires and explosions in industry. Between 1981 and 1986 such incidents were reported to the HSE under the Notification of Accidents and Dangerous Occurrence Regulations (NADOR). A short summary of these statistics is presented in the latter part of Appendix B. These regulations were superseded in 1986 by the Reporting of Injuries, Diseases and Dangerous Occurrence Regulations (RIDDOR). The advent of a new set of reporting regulations prompted the HSE to construct a computerised database containing this accident information. Detailed information from this database for the period April 1986 to March 1987 was made available to the study, from which it was possible to derive an estimate for the Fatal Accident Rate for leak ignitions. This analysis is described in Appendix B. Note that the derived FAR estimate is for leak ignitions only. Statistics for incidents involving open liquid surfaces are also given in the Appendix: however there were no calibrating figures readily available concerning the number of workers exposed to flammable risk indoors in order to produce an equivalent FAR estimate.

The overall target FAR for the British chemical industry is 3 - 4 fatalities per $10^8$ exposed hours, which is taken to be acceptable practice for such industrial-type activity. Kletz [13] argued that the fraction of this figure due to fire and explosion risk should be 0.1, producing a target of about 0.4 fatalities per $10^8$ exposed hours. The estimate using actual data given in Appendix B is 0.6, which appears acceptable when compared with the target figure. However, the figures for incidents involving open liquid surfaces indicate that if the uncalculated contribution from
this source is taken into account the overall estimate will be well above the target. By using a current fatality rate criterion it is therefore apparent that the combined performance of the existing standards and codes is inadequate, and that there is scope for a better method of HAC. The simple comparison drawn here is an indication of the suitability of the FAR as an individual risk criterion by which the effectiveness of HAC methods may be measured (i.e. the risk of death to an individual from a given type of incident). The perceived advantages of using the FAR as a risk criterion here are that it is derived directly from available statistical data and also that it is specifically applicable to process operators. It is also the one of the criteria currently in common use to define zone types (see Appendix H, referred to in Section 14). The use of the FAR resulting from fire and explosion incidents was convenient for this study as it avoided the need for such quantitative risk assessment (QRA) techniques as probit equations to predict fatality rates based on calculations such as likely exposure effects, whilst still demonstrating the need for an improved HAC methodology. Such techniques are usually applied to major incidents such as tank ruptures whose magnitude means that they will have an effect on the surrounding off-site population. Because the current emphasis in such assessments is on the risk to members of the public, these incidents are usually considered to be beyond the scope of HAC. There are also considerable uncertainties attached to using QRA techniques to model the individual risk from small scale incidents where the input data is of variable quality, as proved to be the case with this study. The use of QRA methods to model estimates of individual risk is discussed further in Section 14.

4.2 Property damage

At the outset of the study it was anticipated that the risk of fatality and injury due to ignition of leaks on outdoor plant would be negligible. This was especially felt to be the case for sparsely or unmanned installations. Instead it was thought the amount of damage to such installations would provide a criterion for assessing code performance which would be based on a measure of economic loss, such as insurance loss data. However, it proved difficult to visualise a coherent strategy that is based on both property and personal damage. It would be inevitable that at some future time a situation would arise that would require a choice as to which criterion to use. This may well involve such calculations as the value of a human life which were considered beyond the scope of this study. As outlined in Section 4.1 however, the general risk to plant personnel from leak ignition is not negligible. The notion of a property damage criterion has therefore not been expanded upon.
To summarise, the effectiveness of the HAC codes in current national use has been analysed in terms of the number of fatalities caused by fire and explosion. It was shown that the number of fatalities so caused was above the accepted target for such accidents. This led to the proposal of the fatal accident rate (FAR) as a suitable risk criterion for measuring the effectiveness of HAC methods.

Having considered risk criteria, the following Sections 5 - 9 describe how the data for the quantitative approach described in Figure 3.1 were researched and characterised.
5. Inventories of process plant leak sources

Table 5.1 lists equipment and activities that are commonly thought to require consideration in HAC together with examples of corresponding leak sources. Note that catastrophic events such as vessel ruptures or major pipework fractures have not been included. It is a commonly accepted viewpoint that such events are considered so unlikely as not to be worthwhile protecting against by HAC methods [5]. This section is concerned with methods of estimating the numbers of equipment leak sources listed in Table 5.1 which short-cut the tedious and time consuming exercise of counting each source individually. Such information can be used together with the leak frequency data presented in Section 7 to explore the probability of a flammable leak occurring on a particular plant. This approach is expanded upon in Section 5.3.

As a starting point it was necessary to obtain examples of leak source inventories of existing process plants. A literature search was conducted to obtain published data, and approaches were made to industrial concerns who are supporting the IIGCHL work for additional information.

5.1 Published data

A limited amount of published information was found concerning equipment inventories of process plant. Most sources located were connected with fugitive emission studies conducted by the US EPA (Environmental Protection Agency). Data from studies by Lipton and Lynch [14], Wallace [15] and Hughes et al. [16] are listed in Tables 5.2 - 5.4. Unfortunately the data do not cover all the leak sources listed in Table 5.1; neither does the proprietary (i.e. supplied by IIGCHL members) information presented in Section 5.2. Some of these other leak sources such as agitators and ventilation exhausts are present only on certain types of plant, whilst items such as hoses may be installed on a temporary basis. These leak sources are relatively few in number and may thus require separate classification. The main concern here was to obtain a leak source "profile": in other words the overall distribution of different types leak source on typical process plant. The consideration behind this was that by estimating the distribution of
leak sources on a process plant then eventually an overall level of flammable risk may be formulated. For the purposes of this study the following leak sources were given principal consideration:

- Pumps/compressors
- Flanged connections
- Valves
- Piping
- Small-bore connections (i.e. connections below 10 mm diameter)

It will be seen in Section 12 that the approach adopted for the use of the data pertaining to these leak sources may be expanded to include other leak sources from Table 5.1. The use of this data is considered further in Section 5.3.

Other inventory data of use to this study concerning the above leak sources are given by Hooper [17]. The objective of Hooper’s study was to enable the cost engineer of a new plant to estimate the numbers and therefore the cost of valves and fittings given the length of pipework of a known diameter. In his survey Hooper included pipe lengths and also numbers of valves and fittings aggregated for a number of plants. These are listed in Table 5.5 and are used in Section 5.2 to cross-check the proprietary data compiled in this study.

The data so far considered are concerned with overall plant inventory estimates. In another EPA fugitive emission study, Morgester et al. [18] grouped the number of valves inspected on 6 refineries into numbers contained in individual process units. These are listed in Table 5.6. The tabulated numbers represent the number of valves inspected and not the actual numbers in the various process units. However, taken as a whole they provide an indication of the relative numbers of emission sources for each unit.

### 5.2 Proprietary data

Three sets of data are described below which were made available to the study by industrial participants in this study.
5.2.1 Halogenated hydrocarbon plant

A summary estimate of likely leak sources on this plant was provided in the form of a hazard assessment report [19]. The information is in a similar format to the data lists given in Tables 5.2 - 5.6, and is given in Table 5.7.

5.2.2 Material take-off 1: Hydrocarbon plant

A set of inventory data was also provided concerning a medium-sized plant manufacturing a speciality hydrocarbon. This was in the form of a material take-off for the plant which had been recently completed and handed over. The material take-off was part of the hand-over documentation for the plant and contained details of materials ordered, and vessel and equipment specifications. The list of materials ordered included the following:

<table>
<thead>
<tr>
<th>Material</th>
<th>Specification</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipe length</td>
<td>Material of construction; nominal bore i.e. diameter</td>
</tr>
<tr>
<td>Valves</td>
<td>Material of construction; nominal bore; type</td>
</tr>
<tr>
<td>Flanges</td>
<td>Material of construction; nominal bore; ANSI class (see Section 8.2)</td>
</tr>
<tr>
<td>Gaskets</td>
<td>Material of construction; nominal bore; ANSI class</td>
</tr>
</tbody>
</table>

The equipment specifications included the numbers of pumps, heat exchangers and vessels; the process materials handled; and the temperature and pressure duties. The data from the material take-off were listed and condensed to give the tables shown. Table 5.8 lists details of valve numbers and pipe lengths for the pipe diameters used on the plant. It was necessary to assume negligible pipe and valve wastage i.e. that the amount of materials ordered on the take-off tallied with the content of the completed plant. The number of valves was cross-checked with the pipe lengths by compiling a valve-per-pipe length chart similar to the one synthesised by Hooper for the data shown in Table 5.5. This is shown in Figure 5.1. The diagonal lines are those used by Hooper to denote different piping complexities: for example, the line labelled 30/D corresponds to the case where the number of valves per metre is 30/(Pipe diameter in metres). Note that for different degrees of complexity the downward slope of the lines are the same. It was deduced that the distribution of valves and pipe diameters for Material take-off 1 as indicated by the points shown fell within the boundaries of the Hooper predictions, thus establishing the pipework...
on this plant as representative. There is however one anomaly on the graph i.e. the very high valve to pipe length ratio for small bore valves. This is a result of the take-off including drain valves on plant components and other valves which are not associated with pipework.

Table 5.9 lists numbers of standby and active pumps. These are mostly centrifugal pumps handling hydrocarbons, together with three reactor coolant pumps (one active, two standby).

Table 5.10 lists the number of pipe flanges and gaskets ordered, and includes an estimate of the number of gasketed connections. Note that this estimate cannot be made directly from the tabulated number of gaskets because there is considerable surplus included in the order to account for wastage during pressure testing and commissioning work. Instead, the numbers of nozzles and flanges on pumps, heat exchangers vessels and instruments were totalled from the take-off. The number of connections was then estimated using this figure together with the numbers of flanges and valves (where such items were not indicated as being welded).

Values for the gasketed connection/pipe length ratios are also given in Table 5.10. These calculations were carried out to investigate how the relationship varies with pipe diameter in a similar fashion to the Hooper valve analysis. With some exceptions the connections ratio mostly stays within the range 0.15-0.35, which provides a potentially useful scaling rule. The gasket type ratio given at the bottom of Table 5.10 represents the split of the overall number of gaskets between compressed asbestos fibre (CAF) gaskets, spiral wound joint (SWJ) gaskets, and ring-type joint (RTJ) gaskets. This is important in assessing likely flange leak sizes because reinforced gaskets of the SWJ type are much less likely to fragment and blow out under stress (see Section 8.2).

Finally, an estimate of the number of drain and sample points is given in Table 5.11 which was obtained from the line classification of the plant (line classification is the procedure whereby each piece of pipework is numbered and listed according to, amongst other things, its origin and destination).

5.2.3 Material take-off 2: Offshore production platform

The second material take-off made available to the study concerned an offshore oil production consisting of the following units:
Wellhead module
Separation module
Gas compression module
Utility module
Mud treatment module
Drilling derrick and superstructure
Drilling module
Living quarters
Module support frame
2 x Jet engine generator modules
Generator exhaust tower
Helideck structure
Flare

The take-off piping data were presented in a different form to the hydrocarbon plant data. Because of the large number of different operations being carried out on a platform the data were split into discrete systems for each function which are listed in Table 5.12. This meant that as well as total inventory figures sub-totals could also be derived for those piping systems handling flammable fluids. It was found that the bulk of the piping handling flammable substances was used for the process gas and process liquid systems. This analysis is therefore confined to these systems (here termed Material take-off 3) and the total inventory.

Valve and pipe inventories are given in Tables 5.13 and 5.14. Charts drawn up to cross-check that the valve distributions were in a similar form to the Hooper predictions for typical pipework are given in Figures 5.2 and 5.3. These show that both the total and process fluid piping system distributions are broadly consistent with the Hooper valve/pipe length distributions. If the inventory figures are compared it is apparent that the process fluid pipework is more complex for the smaller diameters than is the total piping. This extra complexity is reflected in the gasketted connections/pipe length ratios given in Tables 5.15 and 5.16. The implication of this is that process pipework has a higher leak source density than, say, service pipework. The different gasket type split for the two systems was also noted: this shows that the process fluid system has a higher proportion of reinforced gaskets than has the total inventory. The process fluid split is also approximately the same as for the hydrocarbon plant. Lastly, an estimated figure of the number of process system pumps is given in Table 5.17.
5.3 Analysis of process inventories by equipment type

Having obtained estimates of leak source distributions the objective of analysing these plant inventories was to build up a picture of a representative "standard" process plant i.e. a plant which has a typical number of pumps, valves and flanges depending on whether it is classed as large, medium or small. The formulation of this plant is described in Section 12: it is sufficient to note here only that it was based principally on data from Material take-off 1. In Section 12 the inventory data are combined with leak frequency and size data from Sections 7 and 8 to provide a basis for an ignition analysis. For example, it becomes possible to estimate the relative frequencies of fire and explosion from a pump leak, valve leak and so on. For this approach it was therefore important that the component ratios of valves/pumps etc. on the standard plant corresponded to those found in the inventory information given in Sections 5.1 and 5.2. An investigation was conducted to determine whether there was any degree of consistency in these component ratios between the various data sources.

Component/pipe length ratios were used in the last section to check take-off data for consistency with literature data. However, pipe length is not a good calibrating measure for general use as piping details are seldom given in literature sources and are difficult to estimate on an existing plant without the aid of take-off data. The number of pumps was considered a better basis for a component ratio analysis as figures for these are usually available whether from literature sources or in actuality. This was the approach adopted for Table 5.18 which gives leak source/pump ratios for the data sources in the foregoing tables. These figures show that there is surprisingly good agreement for quite different types of process plant. The variation in valve and flange ratios is apparently due to plant size rather than type, with smaller plants having fewer of these components per pump than large plants. The high valve per pump ratio for Material take-off 1 is again due to the large amount of extraneous small-bore valves included in the total number. The consistency of the figures suggested they could confidently be used in the "standard plant" approach to ignition analysis described further in Section 12.

5.4 Analysis of process inventories by fluid phase

It is shown in Section 11 that for a given area of process plant the probability of leak ignition is different for gas or vapour leaks and liquid leaks. To enable this distinction to be made for a particular leak it is therefore necessary to know whether the fluid in containment is
gas, flashing or non-flashing liquid. For chemical plants this will of course vary according to
the process. The basis for such a categorisation was constructed by deriving example data from
the material take-offs described above. This was then used as a starting point in the formulation
of the "standard plant". The hydrocarbon plant data were regarded as obviously relevant for this
purpose although the derived data were likely to be specific for this process. For this reason it
was also a useful exercise to perform for the platform take-off data, which could be regarded
as typical for all offshore platforms.

The material take-off for the hydrocarbon plant does not explicitly specify the phase of
the fluid in each length of piping. However, a rough estimate of the overall phase split was made
by considering the temperature and pressure regimes of the principal process items. The result
of this is given in Table 5.19, which lists the numbers of these items and the types of fluid
handled in each (or more specifically the type of fluid that may leak). Note that some items
handle more than one fluid type. It was felt that the spread of fluid types thus derived constituted
a reasonable overall approximation although it does not include details on whether the split is
different over the range of pipe diameters e.g. it is not certain whether large pipes mainly carry
gas or liquid. This consideration becomes significant when considering the emission from such
sources, since it is clear that a flashing liquid leak from a large pipe split presents a far greater
hazard than the equivalent vapour leak due to the higher mass flow.

It was possible to be more specific for the production platform data with regard to the
process fluid system. Table 5.20 shows the pipe lengths handling process liquid and gas for each
pipe diameter, together with the percentage of the total. However, no information was available
on the percentage of process liquid piping handling flashing liquid, which restricts the usefulness
of the information. Also, from the table it became clear that there was little benefit to be gained
from investigating how the gas-liquid split varied with pipe diameter due to the randomness of
the distribution.

In conclusion, this section has investigated the formulation of a leak source profile for a
typical process plant. Firstly, the leak sources of principal interest have been identified. Literature
data concerning numbers of these leak sources have been presented where available. A detailed
analysis of two plants handling flammable materials has been performed: for a medium-sized
hydrocarbon plant; and for an offshore production platform. From these data sources it has been
concluded that there is a reasonably good correlation between the ratios of different leak source
types to the number of pumps for each plant considered. A crude estimate was also made of the
distribution of fluid phases being handled on the hydrocarbon plant. These estimates are utilised in Section 12 in the formulation of a model to estimate fire and explosion frequencies for typical process plant.
<table>
<thead>
<tr>
<th>Equipment</th>
<th>Example of leak sources</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flanges: Pipe or vessel</td>
<td>Gasket</td>
</tr>
<tr>
<td>Valves</td>
<td>Body rupture</td>
</tr>
<tr>
<td></td>
<td>Gland/seal</td>
</tr>
<tr>
<td>Pumps:</td>
<td></td>
</tr>
<tr>
<td>centrifugal</td>
<td>Seal</td>
</tr>
<tr>
<td>reciprocating</td>
<td>Seal</td>
</tr>
<tr>
<td></td>
<td>Valve chamber</td>
</tr>
<tr>
<td>Compressors:</td>
<td></td>
</tr>
<tr>
<td>centrifugal</td>
<td>Seal</td>
</tr>
<tr>
<td>reciprocating</td>
<td>Seal</td>
</tr>
<tr>
<td></td>
<td>Valve chamber</td>
</tr>
<tr>
<td>Agitators</td>
<td>Gland/seal</td>
</tr>
<tr>
<td>Drain points</td>
<td>Full bore or partial release</td>
</tr>
<tr>
<td>Drains</td>
<td>Rising vapours</td>
</tr>
<tr>
<td>Sample points</td>
<td>Full bore or partial release</td>
</tr>
<tr>
<td>Open ended lines</td>
<td>Full bore release</td>
</tr>
<tr>
<td>Atmospheric vents</td>
<td>Full bore release</td>
</tr>
<tr>
<td>Pressure relief vents</td>
<td>Full bore release</td>
</tr>
<tr>
<td>Hoses</td>
<td>Guillotine fracture</td>
</tr>
<tr>
<td></td>
<td>Connecting fittings</td>
</tr>
<tr>
<td>Small bore connections</td>
<td>Guillotine fracture</td>
</tr>
<tr>
<td></td>
<td>Fitting leak</td>
</tr>
<tr>
<td>Ventilation exhausts</td>
<td>Diluted vapour release</td>
</tr>
</tbody>
</table>

Activities:

<table>
<thead>
<tr>
<th>Activities</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Open pools or containers</td>
<td>Rising vapours</td>
</tr>
<tr>
<td>Tanker/vessel filling</td>
<td>Overfilling, liquid release</td>
</tr>
<tr>
<td>Maintenance</td>
<td>Dismantling of equipment and pipework</td>
</tr>
<tr>
<td>Container transfer and handling</td>
<td>Puncture of container e.g. barrel, leakage through mis-handling</td>
</tr>
</tbody>
</table>
Table 5.2 Estimated number of leak sources in a large refinery [Lipton and Lynch, 1987]

<table>
<thead>
<tr>
<th>Leak source</th>
<th>Number of items</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flanges</td>
<td>46,500</td>
</tr>
<tr>
<td>Valves</td>
<td>11,500</td>
</tr>
<tr>
<td>Pump seals</td>
<td>350</td>
</tr>
<tr>
<td>Compressors</td>
<td>70</td>
</tr>
<tr>
<td>Relief valves</td>
<td>100</td>
</tr>
<tr>
<td>Drains</td>
<td>650</td>
</tr>
</tbody>
</table>

Table 5.3 Estimated number of leak sources on a medium sized plant [Wallace, 1979]

<table>
<thead>
<tr>
<th>Leak source</th>
<th>Number of items</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flanges</td>
<td>2,410</td>
</tr>
<tr>
<td>Valves;</td>
<td></td>
</tr>
<tr>
<td>- In-line gas</td>
<td>365</td>
</tr>
<tr>
<td>- In-line liquid</td>
<td>670</td>
</tr>
<tr>
<td>- Open-ended</td>
<td>415</td>
</tr>
<tr>
<td>Pump seals;</td>
<td></td>
</tr>
<tr>
<td>- Packed</td>
<td>6</td>
</tr>
<tr>
<td>- Mechanical, single</td>
<td>43</td>
</tr>
<tr>
<td>- Mechanical, double</td>
<td>10</td>
</tr>
<tr>
<td>Compressor seals</td>
<td>2</td>
</tr>
<tr>
<td>Safety relief valves</td>
<td>5</td>
</tr>
</tbody>
</table>
Table 5.4 Estimated number of sources on four petrochemical plants [Hughes, Tierney and Khan, 1979]

<table>
<thead>
<tr>
<th>Leak Source</th>
<th>Monochlorobutadiene plant</th>
<th>Butadiene plant</th>
<th>Ethylene Oxide plant</th>
<th>Dimethyl tetradecylate plant</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flanges</td>
<td>1,500</td>
<td>26,000</td>
<td>n/a</td>
<td>n/a</td>
</tr>
<tr>
<td>Valves</td>
<td>640</td>
<td>6,700</td>
<td>n/a</td>
<td>n/a</td>
</tr>
<tr>
<td>Pumps</td>
<td>25</td>
<td>174</td>
<td>69</td>
<td>67</td>
</tr>
</tbody>
</table>
Table 5.5 Data base of pipework fittings and valves [Hooper, 1982]

<table>
<thead>
<tr>
<th>Pipe diameter (mm)</th>
<th>Total pipe length (m)</th>
<th>Number of flanges</th>
<th>Number of valves</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>11 152</td>
<td>1 818</td>
<td>11 859</td>
</tr>
<tr>
<td>20</td>
<td>10 867</td>
<td>2 973</td>
<td>7 551</td>
</tr>
<tr>
<td>25</td>
<td>40 851</td>
<td>12 552</td>
<td>10 363</td>
</tr>
<tr>
<td>40</td>
<td>39 768</td>
<td>7 299</td>
<td>3 313</td>
</tr>
<tr>
<td>50</td>
<td>46 880</td>
<td>11 727</td>
<td>4 199</td>
</tr>
<tr>
<td>75</td>
<td>41 191</td>
<td>10 427</td>
<td>2 441</td>
</tr>
<tr>
<td>100</td>
<td>27 790</td>
<td>6 608</td>
<td>1 346</td>
</tr>
<tr>
<td>150</td>
<td>25 498</td>
<td>4 578</td>
<td>898</td>
</tr>
<tr>
<td>200</td>
<td>22 200</td>
<td>3 592</td>
<td>466</td>
</tr>
<tr>
<td>250</td>
<td>12 869</td>
<td>1 613</td>
<td>301</td>
</tr>
<tr>
<td>300</td>
<td>5 395</td>
<td>762</td>
<td>162</td>
</tr>
<tr>
<td>350</td>
<td>1 311</td>
<td>342</td>
<td>72</td>
</tr>
<tr>
<td>400</td>
<td>3 377</td>
<td>506</td>
<td>90</td>
</tr>
<tr>
<td>450</td>
<td>1 158</td>
<td>362</td>
<td>41</td>
</tr>
<tr>
<td>500</td>
<td>1 869</td>
<td>804</td>
<td>34</td>
</tr>
<tr>
<td>600</td>
<td>1 963</td>
<td>357</td>
<td>40</td>
</tr>
<tr>
<td>900</td>
<td>528</td>
<td>66</td>
<td>12</td>
</tr>
</tbody>
</table>
Table 5.6 Number of valves inspected in 6 refineries categorised by unit type [Morgester et al., 1979]

<table>
<thead>
<tr>
<th>Unit</th>
<th>Number of valves inspected</th>
</tr>
</thead>
<tbody>
<tr>
<td>Storage</td>
<td>2398</td>
</tr>
<tr>
<td>Refiner</td>
<td>1803</td>
</tr>
<tr>
<td>FCC</td>
<td>1737</td>
</tr>
<tr>
<td>Fractionation</td>
<td>1540</td>
</tr>
<tr>
<td>Alkylation</td>
<td>1522</td>
</tr>
<tr>
<td>Crude unit</td>
<td>1243</td>
</tr>
<tr>
<td>Coker</td>
<td>870</td>
</tr>
<tr>
<td>LSFO</td>
<td>726</td>
</tr>
<tr>
<td>Hydrotreating</td>
<td>637</td>
</tr>
<tr>
<td>Hydrocracker</td>
<td>609</td>
</tr>
<tr>
<td>Other units</td>
<td>60</td>
</tr>
</tbody>
</table>

Table 5.7 Summary equipment inventory for halogenated hydrocarbon manufacturing facility (process streams only)

<table>
<thead>
<tr>
<th>Item</th>
<th>Number</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flanges (not including vessel manholes)</td>
<td>860</td>
</tr>
<tr>
<td>Valves</td>
<td>266</td>
</tr>
<tr>
<td>Pumps</td>
<td>7</td>
</tr>
<tr>
<td>Sample points</td>
<td>5</td>
</tr>
<tr>
<td>Vents</td>
<td>2</td>
</tr>
</tbody>
</table>
Table 5.8 Material take-off 1: Hydrocarbon valve and pipe length inventory

<table>
<thead>
<tr>
<th>Pipe diam. (mm)</th>
<th>Carbon steel valves</th>
<th>Pipe length (m)</th>
<th>Valve/ Pipe length (m³)</th>
<th>Stainless steel valves</th>
<th>Pipe length (m)</th>
<th>Valve/ Pipe length (m³)</th>
<th>Overall Valve/Pipe Length (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>525</td>
<td>272</td>
<td>1.93</td>
<td>12</td>
<td>18</td>
<td>0.67</td>
<td>1.852</td>
</tr>
<tr>
<td>20</td>
<td>197</td>
<td>84</td>
<td>2.345</td>
<td>6</td>
<td>24</td>
<td>0.25</td>
<td>1.880</td>
</tr>
<tr>
<td>25</td>
<td>716</td>
<td>2020</td>
<td>0.354</td>
<td>71</td>
<td>384</td>
<td>0.185</td>
<td>0.327</td>
</tr>
<tr>
<td>40</td>
<td>112</td>
<td>2046</td>
<td>0.055</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0.055</td>
</tr>
<tr>
<td>50</td>
<td>223</td>
<td>1812</td>
<td>0.123</td>
<td>2</td>
<td>48</td>
<td>0.042</td>
<td>0.121</td>
</tr>
<tr>
<td>80</td>
<td>180</td>
<td>4422</td>
<td>0.041</td>
<td>0</td>
<td>12</td>
<td>0</td>
<td>0.041</td>
</tr>
<tr>
<td>100</td>
<td>64</td>
<td>762</td>
<td>0.084</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.084</td>
</tr>
<tr>
<td>150</td>
<td>33</td>
<td>750</td>
<td>0.044</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.044</td>
</tr>
<tr>
<td>200</td>
<td>15</td>
<td>552</td>
<td>0.027</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.027</td>
</tr>
<tr>
<td>250</td>
<td>9</td>
<td>322</td>
<td>0.028</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.028</td>
</tr>
<tr>
<td>300</td>
<td>5</td>
<td>66</td>
<td>0.076</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.076</td>
</tr>
<tr>
<td>350</td>
<td>0</td>
<td>24</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>400</td>
<td>0</td>
<td>672</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>450</td>
<td>2</td>
<td>30</td>
<td>0.067</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.067</td>
</tr>
</tbody>
</table>

Table 5.9 Material take-off 1: Hydrocarbon plant pumps

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Active pumps</td>
<td>13</td>
</tr>
<tr>
<td>Standby pumps</td>
<td>10</td>
</tr>
<tr>
<td>Loading pumps</td>
<td>2</td>
</tr>
</tbody>
</table>

36
<table>
<thead>
<tr>
<th>Pipe diam. (mm)</th>
<th>Pipe length (m)</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
<th>8</th>
</tr>
</thead>
<tbody>
<tr>
<td>15^1</td>
<td>290</td>
<td>132</td>
<td>173</td>
<td>6</td>
<td>0</td>
<td>84</td>
<td>90</td>
<td>42</td>
<td>0.310</td>
</tr>
<tr>
<td>20^2</td>
<td>108</td>
<td>250</td>
<td>394</td>
<td>8</td>
<td>0</td>
<td>93</td>
<td>201</td>
<td>49</td>
<td>1.861^6</td>
</tr>
<tr>
<td>25^3</td>
<td>2404</td>
<td>821</td>
<td>911</td>
<td>127</td>
<td>393</td>
<td>195</td>
<td>715</td>
<td>106</td>
<td>0.297</td>
</tr>
<tr>
<td>40^4</td>
<td>2046</td>
<td>309</td>
<td>367</td>
<td>58</td>
<td>112</td>
<td>99</td>
<td>268</td>
<td>40</td>
<td>0.131</td>
</tr>
<tr>
<td>50^5</td>
<td>1858</td>
<td>701</td>
<td>703</td>
<td>103</td>
<td>450</td>
<td>75</td>
<td>628</td>
<td>73</td>
<td>0.338</td>
</tr>
<tr>
<td>80</td>
<td>4434</td>
<td>748</td>
<td>538</td>
<td>43</td>
<td>360</td>
<td>68</td>
<td>470</td>
<td>277</td>
<td>0.106</td>
</tr>
<tr>
<td>100</td>
<td>762</td>
<td>239</td>
<td>252</td>
<td>31</td>
<td>128</td>
<td>47</td>
<td>206</td>
<td>33</td>
<td>0.270</td>
</tr>
<tr>
<td>150</td>
<td>750</td>
<td>187</td>
<td>181</td>
<td>17</td>
<td>66</td>
<td>49</td>
<td>132</td>
<td>55</td>
<td>0.176</td>
</tr>
<tr>
<td>200</td>
<td>552</td>
<td>132</td>
<td>127</td>
<td>17</td>
<td>30</td>
<td>40</td>
<td>87</td>
<td>45</td>
<td>0.158</td>
</tr>
<tr>
<td>250</td>
<td>322</td>
<td>67</td>
<td>72</td>
<td>8</td>
<td>18</td>
<td>23</td>
<td>49</td>
<td>18</td>
<td>0.152</td>
</tr>
<tr>
<td>300</td>
<td>66</td>
<td>38</td>
<td>34</td>
<td>9</td>
<td>10</td>
<td>8</td>
<td>27</td>
<td>11</td>
<td>0.409</td>
</tr>
<tr>
<td>350</td>
<td>24</td>
<td>6</td>
<td>8</td>
<td>2</td>
<td>0</td>
<td>3</td>
<td>5</td>
<td>1</td>
<td>0.208</td>
</tr>
<tr>
<td>400</td>
<td>672</td>
<td>25</td>
<td>21</td>
<td>4</td>
<td>0</td>
<td>9</td>
<td>13</td>
<td>12</td>
<td>0.019</td>
</tr>
<tr>
<td>450</td>
<td>30</td>
<td>16</td>
<td>18</td>
<td>10</td>
<td>0</td>
<td>2</td>
<td>12</td>
<td>4</td>
<td>0.400</td>
</tr>
<tr>
<td>500</td>
<td>90</td>
<td>8</td>
<td>8</td>
<td>8</td>
<td>0</td>
<td>0</td>
<td>8</td>
<td>0</td>
<td>0.089</td>
</tr>
<tr>
<td>600</td>
<td>12</td>
<td>6</td>
<td>6</td>
<td>2</td>
<td>0</td>
<td>2</td>
<td>4</td>
<td>2</td>
<td>0.333</td>
</tr>
</tbody>
</table>

Gasket type ratio CAF:SWJ:RTJ = 35:55:10

Column headings:

1. Total gaskets ordered
2. Total flanges ordered
3. Pipe-instrument/nozzle flange connections
4. Pipe-valve flange connections
5. Pipe-pipe flange connections
6. Total flanged connections (i.e. 3 + 4 + 5)
7. Residual gaskets for pressure testing, spares etc.
8. Flanged connection/pipe length ratio (m⁻¹)

continued
Footnotes for Table 5.10:

1. From the material-take off it was possible to assume that all valve-pipe connections here were welded or had compression fittings. Not enough gaskets were included in the inventory to justify the assumption that some were flanged.

2. The same assumption was made here as for 1. regarding valve-pipe connections.

3. An assumption of 25% flanged and 75% welded valve connections was made to arrive at a reasonable estimate for the number of gaskets used.

4. As for 3. assuming half the valves are welded and half are flanged.

5. All valve-pipe connections here and below are assumed to be flanged.

6. This figure is high because of the large amount of instrumentation with connections this size.

Table 5.11 Material take-off 1: Hydrocarbon plant drain and sample points

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Oily water drains</td>
<td>15</td>
</tr>
<tr>
<td>Flammable material</td>
<td>5</td>
</tr>
<tr>
<td>Drains and sample points</td>
<td>8</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
</tbody>
</table>

It is uncertain how many drains may give rise to flammable releases since some are piped to a safe location.
<table>
<thead>
<tr>
<th>Process gas</th>
<th>Instrument air</th>
</tr>
</thead>
<tbody>
<tr>
<td>Process liquid</td>
<td>Process air</td>
</tr>
<tr>
<td>(i.e. Material take-off 3)</td>
<td>Bulk air</td>
</tr>
<tr>
<td>High pressure mud</td>
<td>Plant air</td>
</tr>
<tr>
<td>Low pressure mud</td>
<td>Engine exhaust</td>
</tr>
<tr>
<td>Kill mud</td>
<td>Atmospheric vent (Hazardous)</td>
</tr>
<tr>
<td>Deluge mud</td>
<td>Atmospheric vent (Non-hazardous)</td>
</tr>
<tr>
<td>Oily mud drains</td>
<td>Bulk vent</td>
</tr>
<tr>
<td>Fire fighting foam</td>
<td>Warm flare header</td>
</tr>
<tr>
<td>Untreated sea water</td>
<td>Open drains (Hazardous)</td>
</tr>
<tr>
<td>Fire water</td>
<td>Open drains (Non-hazardous)</td>
</tr>
<tr>
<td>Drill water</td>
<td>Deck drains</td>
</tr>
<tr>
<td>Injection water</td>
<td>Closed drains</td>
</tr>
<tr>
<td>Produced water</td>
<td>Black sewage</td>
</tr>
<tr>
<td>Water drains</td>
<td>Waste treatment (Grey water)</td>
</tr>
<tr>
<td>Hydraulic oil</td>
<td>High pressure cement</td>
</tr>
<tr>
<td>Seal oil</td>
<td>Low pressure cement</td>
</tr>
<tr>
<td>Low-toxicity oil</td>
<td>Bulk cement</td>
</tr>
<tr>
<td>Lube oil</td>
<td>Cement dump</td>
</tr>
<tr>
<td>Diesel fuel</td>
<td>Bulk barytes/bentonite</td>
</tr>
<tr>
<td>Jet fuel</td>
<td>Halon</td>
</tr>
<tr>
<td>Fuel gas</td>
<td>Well fluid</td>
</tr>
<tr>
<td>Methanol</td>
<td>Sodium Hypochlorite</td>
</tr>
<tr>
<td>Cooling medium</td>
<td>Glycol</td>
</tr>
<tr>
<td>Heating medium</td>
<td>Chemicals - General</td>
</tr>
<tr>
<td>Wax inhibitor</td>
<td></td>
</tr>
<tr>
<td>Corrosion inhibitor</td>
<td></td>
</tr>
<tr>
<td>Inert gas</td>
<td></td>
</tr>
<tr>
<td>De-emulsifier</td>
<td></td>
</tr>
<tr>
<td>Oxygen scavenger</td>
<td></td>
</tr>
</tbody>
</table>
Table 5.13 Material take-off 2: Production platform valve/pipe length inventory (total)

<table>
<thead>
<tr>
<th>Pipe diam. (mm)</th>
<th>Carbon steel pipe length (m)</th>
<th>Valve/ Pipe steel valves length (m)</th>
<th>Overall pipe Valve/ Pipe length (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>284</td>
<td>41</td>
<td>6.89</td>
</tr>
<tr>
<td>20</td>
<td>443</td>
<td>55</td>
<td>8.10</td>
</tr>
<tr>
<td>25</td>
<td>604</td>
<td>3489</td>
<td>0.17</td>
</tr>
<tr>
<td>40</td>
<td>320</td>
<td>2730</td>
<td>0.12</td>
</tr>
<tr>
<td>50</td>
<td>473</td>
<td>3385</td>
<td>0.14</td>
</tr>
<tr>
<td>80</td>
<td>190</td>
<td>3149</td>
<td>0.06</td>
</tr>
<tr>
<td>100</td>
<td>174</td>
<td>2309</td>
<td>0.08</td>
</tr>
<tr>
<td>125</td>
<td>7</td>
<td>433</td>
<td>0.016</td>
</tr>
<tr>
<td>150</td>
<td>246</td>
<td>2410</td>
<td>0.102</td>
</tr>
<tr>
<td>200</td>
<td>105</td>
<td>1333</td>
<td>0.079</td>
</tr>
<tr>
<td>250</td>
<td>65</td>
<td>869</td>
<td>0.075</td>
</tr>
<tr>
<td>300</td>
<td>17</td>
<td>367</td>
<td>0.046</td>
</tr>
<tr>
<td>350</td>
<td>6</td>
<td>133</td>
<td>0.045</td>
</tr>
<tr>
<td>400</td>
<td>7</td>
<td>180</td>
<td>0.039</td>
</tr>
<tr>
<td>450</td>
<td>0</td>
<td>140</td>
<td>0</td>
</tr>
<tr>
<td>500</td>
<td>7</td>
<td>174</td>
<td>0.040</td>
</tr>
</tbody>
</table>

40
Table 5.14 Material take-off 3: Production platform valve/pipe length inventory for process fluid streams

<table>
<thead>
<tr>
<th>Pipe diam. (mm)</th>
<th>Carbon steel valves</th>
<th>Pipe length (m)</th>
<th>Valve/ Pipe length (m⁻¹)</th>
<th>Stainless steel valves</th>
<th>Pipe length (m)</th>
<th>Valve/ Pipe length (m⁻¹)</th>
<th>Overall Valve/ Pipe length (m⁻¹)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>135</td>
<td>7.6</td>
<td>17.8</td>
<td>15</td>
<td>0</td>
<td>0</td>
<td>19.7</td>
</tr>
<tr>
<td>20</td>
<td>137</td>
<td>3.4</td>
<td>40.3</td>
<td>14</td>
<td>0</td>
<td>0</td>
<td>44.4</td>
</tr>
<tr>
<td>25</td>
<td>280</td>
<td>45.0</td>
<td>6.22</td>
<td>33</td>
<td>5.2</td>
<td>6.35</td>
<td>6.24</td>
</tr>
<tr>
<td>40</td>
<td>65</td>
<td>228.4</td>
<td>0.285</td>
<td>25</td>
<td>127.2</td>
<td>0.197</td>
<td>0.253</td>
</tr>
<tr>
<td>50</td>
<td>138</td>
<td>495.1</td>
<td>0.279</td>
<td>21</td>
<td>46.2</td>
<td>0.455</td>
<td>0.294</td>
</tr>
<tr>
<td>80</td>
<td>51</td>
<td>467.3</td>
<td>0.109</td>
<td>4</td>
<td>49.4</td>
<td>0.081</td>
<td>0.106</td>
</tr>
<tr>
<td>100</td>
<td>75</td>
<td>309.0</td>
<td>0.243</td>
<td>2</td>
<td>87.7</td>
<td>0.023</td>
<td>0.194</td>
</tr>
<tr>
<td>150</td>
<td>36</td>
<td>242.4</td>
<td>0.149</td>
<td>3</td>
<td>53.6</td>
<td>0.056</td>
<td>0.132</td>
</tr>
<tr>
<td>200</td>
<td>17</td>
<td>205.3</td>
<td>0.083</td>
<td>2</td>
<td>33.5</td>
<td>0.060</td>
<td>0.080</td>
</tr>
<tr>
<td>250</td>
<td>28</td>
<td>398.4</td>
<td>0.070</td>
<td>2</td>
<td>25.5</td>
<td>0.078</td>
<td>0.071</td>
</tr>
<tr>
<td>300</td>
<td>9</td>
<td>60.9</td>
<td>0.148</td>
<td>1</td>
<td>0</td>
<td>0</td>
<td>0.164</td>
</tr>
<tr>
<td>350</td>
<td>3</td>
<td>31.1</td>
<td>0.096</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0.096</td>
</tr>
<tr>
<td>400</td>
<td>6</td>
<td>165.3</td>
<td>0.036</td>
<td>2</td>
<td>6.6</td>
<td>0.303</td>
<td>0.047</td>
</tr>
<tr>
<td>450</td>
<td>0</td>
<td>44.1</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>500</td>
<td>2</td>
<td>25.6</td>
<td>0.078</td>
<td>0</td>
<td>33.9</td>
<td>0</td>
<td>0.034</td>
</tr>
</tbody>
</table>
Table 5.15 Material take-off 2: Production platform gasket and flange inventory (total)

<table>
<thead>
<tr>
<th>Pipe diam. (mm)</th>
<th>Pipe length (m)</th>
<th>Total gaskets ordered</th>
<th>Total flanges&lt;sup&gt;1&lt;/sup&gt;</th>
<th>Estimated gaskets used&lt;sup&gt;1&lt;/sup&gt;</th>
<th>Gasketed joints/pipe length</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>2292</td>
<td>1517</td>
<td>1737</td>
<td>1213</td>
<td>0.529</td>
</tr>
<tr>
<td>20</td>
<td>253</td>
<td>1689</td>
<td>1056</td>
<td>1351</td>
<td>5.340</td>
</tr>
<tr>
<td>25</td>
<td>10317</td>
<td>2945</td>
<td>2564</td>
<td>2356</td>
<td>0.228</td>
</tr>
<tr>
<td>40</td>
<td>6956</td>
<td>1638</td>
<td>1462</td>
<td>1310</td>
<td>0.188</td>
</tr>
<tr>
<td>50</td>
<td>7334</td>
<td>3014</td>
<td>2037</td>
<td>2037</td>
<td>0.329</td>
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<tr>
<td>75</td>
<td>5226</td>
<td>1478</td>
<td>1431</td>
<td>1182</td>
<td>0.226</td>
</tr>
<tr>
<td>100</td>
<td>4256</td>
<td>1491</td>
<td>1595</td>
<td>1193</td>
<td>0.280</td>
</tr>
<tr>
<td>125</td>
<td>576</td>
<td>163</td>
<td>152</td>
<td>130</td>
<td>2.264</td>
</tr>
<tr>
<td>150</td>
<td>3830</td>
<td>1341</td>
<td>1266</td>
<td>1073</td>
<td>0.280</td>
</tr>
<tr>
<td>200</td>
<td>1990</td>
<td>766</td>
<td>725</td>
<td>613</td>
<td>0.308</td>
</tr>
<tr>
<td>250</td>
<td>1108</td>
<td>463</td>
<td>409</td>
<td>370</td>
<td>0.334</td>
</tr>
<tr>
<td>300</td>
<td>765</td>
<td>193</td>
<td>189</td>
<td>154</td>
<td>0.201</td>
</tr>
<tr>
<td>350</td>
<td>194</td>
<td>70</td>
<td>60</td>
<td>56</td>
<td>0.289</td>
</tr>
<tr>
<td>400</td>
<td>824</td>
<td>212</td>
<td>212</td>
<td>170</td>
<td>0.206</td>
</tr>
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<td>450</td>
<td>210</td>
<td>78</td>
<td>80</td>
<td>62</td>
<td>0.295</td>
</tr>
<tr>
<td>500</td>
<td>457</td>
<td>72</td>
<td>83</td>
<td>58</td>
<td>0.127</td>
</tr>
<tr>
<td>Other sizes</td>
<td></td>
<td>99</td>
<td>604</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Gasket type ratio CAF:SWJ:RTJ/Rubber = 66:33:1

1. This is the number of pipe flanges only. The total number of flanges on vessels, pumps and so on was not known. Since the total numbers of flanges and gaskets is roughly the same for each size it was considered likely that the majority of flanges are used to connect pipe to pieces of equipment as opposed to other lengths of pipe. However, gaskets are over-ordered to allow for wastage during pressure testing and commission. The assumption therefore made was that the number of gasketted joints is approximately 80% of the number of gaskets given above.
Table 5.16 Material take-off 3: Production platform gasket and flange inventory (process fluids)

<table>
<thead>
<tr>
<th>Pipe diam. (mm)</th>
<th>Pipe length (m)</th>
<th>Total gaskets ordered</th>
<th>Total flanges¹</th>
<th>Estimated gaskets used¹</th>
<th>Gasketted joints/ pipe length</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>7.6</td>
<td>375</td>
<td>305</td>
<td>300</td>
<td>39.5</td>
</tr>
<tr>
<td>20</td>
<td>3.4</td>
<td>307</td>
<td>170</td>
<td>246</td>
<td>72.4</td>
</tr>
<tr>
<td>25</td>
<td>50.2</td>
<td>715</td>
<td>648</td>
<td>512</td>
<td>11.4</td>
</tr>
<tr>
<td>40</td>
<td>355.6</td>
<td>405</td>
<td>341</td>
<td>324</td>
<td>0.911</td>
</tr>
<tr>
<td>50</td>
<td>541.3</td>
<td>551</td>
<td>420</td>
<td>441</td>
<td>0.814</td>
</tr>
<tr>
<td>75</td>
<td>516.7</td>
<td>251</td>
<td>182</td>
<td>201</td>
<td>0.389</td>
</tr>
<tr>
<td>100</td>
<td>396.7</td>
<td>279</td>
<td>223</td>
<td>223</td>
<td>0.563</td>
</tr>
<tr>
<td>150</td>
<td>296.0</td>
<td>194</td>
<td>145</td>
<td>155</td>
<td>0.524</td>
</tr>
<tr>
<td>200</td>
<td>238.8</td>
<td>124</td>
<td>107</td>
<td>99</td>
<td>0.415</td>
</tr>
<tr>
<td>250</td>
<td>423.9</td>
<td>122</td>
<td>186</td>
<td>98</td>
<td>0.230</td>
</tr>
<tr>
<td>300</td>
<td>60.9</td>
<td>41</td>
<td>33</td>
<td>33</td>
<td>0.539</td>
</tr>
<tr>
<td>350</td>
<td>31.1</td>
<td>20</td>
<td>13</td>
<td>16</td>
<td>0.514</td>
</tr>
<tr>
<td>400</td>
<td>171.9</td>
<td>61</td>
<td>60</td>
<td>49</td>
<td>0.284</td>
</tr>
<tr>
<td>450</td>
<td>44.1</td>
<td>16</td>
<td>10</td>
<td>13</td>
<td>0.290</td>
</tr>
<tr>
<td>500</td>
<td>59.5</td>
<td>22</td>
<td>30</td>
<td>18</td>
<td>0.296</td>
</tr>
</tbody>
</table>

Gasket type ratio CAF:SWJ:RTJ = 36:64:0.001 (for footnote see previous page)

Table 5.17 Production platform compressors and pumps handling flammable process and non-process materials

<p>| | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Process pumps (pumps concerned with drilling and processing operations)</td>
<td>39</td>
<td></td>
</tr>
<tr>
<td>Non-process pumps (e.g. diesel fuel pumps)</td>
<td>28</td>
<td></td>
</tr>
<tr>
<td>Compressors</td>
<td>6</td>
<td></td>
</tr>
</tbody>
</table>

The split between active and standby pumps and compressors was not known in this case.
Table 5.18 Plant component ratios per pump for data sources

<table>
<thead>
<tr>
<th>Ratio of item/pump</th>
<th>Data source (see below)</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flanges</td>
<td>133</td>
<td>41</td>
<td>60</td>
<td>150</td>
<td>123</td>
<td>117</td>
<td>72</td>
<td></td>
</tr>
<tr>
<td>Valves</td>
<td>33</td>
<td>25</td>
<td>26</td>
<td>39</td>
<td>38</td>
<td>87</td>
<td>29</td>
<td></td>
</tr>
<tr>
<td>Compressors</td>
<td>0.2</td>
<td>0.03</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0.15</td>
<td></td>
</tr>
<tr>
<td>Pressure relief vents</td>
<td>0.29</td>
<td>0.08</td>
<td>-</td>
<td>-</td>
<td>0.29</td>
<td>0.44</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Sample points</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0.71</td>
<td>*0.32</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Drain points</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Drains</td>
<td>1.86</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0.8</td>
<td>-</td>
<td></td>
</tr>
</tbody>
</table>

*this represents the total for both drain and sample points combined

Columns:

1. Lipton and Lynch  
2. Wallace  
3. Hughes et al. (Monochlorobenzene)  
4. Hughes et al. (Butadiene)  
5. Chlorinated hydrocarbon plant  
6. Hydrocarbon plant  
7. Production platform
Table 5.19 Material take-off 1: Spread of fluid phases within hydrocarbon plant components

<table>
<thead>
<tr>
<th>Item</th>
<th>Number</th>
<th>Gas/vapour</th>
<th>Flashing liquid</th>
<th>Non-flashing liquid</th>
<th>Steam/water</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pumps (active)</td>
<td>15</td>
<td>8</td>
<td>7</td>
<td>7</td>
<td></td>
</tr>
<tr>
<td>Towers</td>
<td>4</td>
<td>4</td>
<td>4</td>
<td>4</td>
<td></td>
</tr>
<tr>
<td>Tanks</td>
<td>7</td>
<td>7</td>
<td>7</td>
<td>7</td>
<td></td>
</tr>
<tr>
<td>Drums</td>
<td>7</td>
<td>5</td>
<td>3</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Reactors</td>
<td>2</td>
<td>2</td>
<td></td>
<td>2</td>
<td></td>
</tr>
<tr>
<td>Heat exchangers (active)</td>
<td>12</td>
<td>9</td>
<td>12</td>
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<td>11</td>
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<tr>
<td>Total</td>
<td>27</td>
<td>27</td>
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</tr>
<tr>
<td>Percent</td>
<td>30</td>
<td>30</td>
<td>23</td>
<td>17</td>
<td></td>
</tr>
<tr>
<td>Percent (excluding steam/water)</td>
<td>36</td>
<td>36</td>
<td>28</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
### Table 5.20 Material take-off 3: Spread of fluid phases within production platform process fluid pipework

<table>
<thead>
<tr>
<th>Pipe diameter (mm)</th>
<th>Total pipework length (m)</th>
<th>%Liquid</th>
<th>%Gas</th>
<th>Total length as percentage of total on platform</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>7.6</td>
<td>53</td>
<td>47</td>
<td>0</td>
</tr>
<tr>
<td>20</td>
<td>3.4</td>
<td>59</td>
<td>41</td>
<td>0.1</td>
</tr>
<tr>
<td>25</td>
<td>50.2</td>
<td>48</td>
<td>52</td>
<td>0.5</td>
</tr>
<tr>
<td>40</td>
<td>356</td>
<td>94</td>
<td>6</td>
<td>5</td>
</tr>
<tr>
<td>50</td>
<td>541</td>
<td>45</td>
<td>55</td>
<td>7</td>
</tr>
<tr>
<td>75</td>
<td>517</td>
<td>85</td>
<td>15</td>
<td>10</td>
</tr>
<tr>
<td>100</td>
<td>397</td>
<td>38</td>
<td>62</td>
<td>9</td>
</tr>
<tr>
<td>150</td>
<td>296</td>
<td>4</td>
<td>96</td>
<td>7</td>
</tr>
<tr>
<td>200</td>
<td>239</td>
<td>68</td>
<td>32</td>
<td>12</td>
</tr>
<tr>
<td>250</td>
<td>424</td>
<td>69</td>
<td>31</td>
<td>38</td>
</tr>
<tr>
<td>300</td>
<td>61</td>
<td>42</td>
<td>58</td>
<td>8</td>
</tr>
<tr>
<td>350</td>
<td>31</td>
<td>0</td>
<td>100</td>
<td>16</td>
</tr>
<tr>
<td>400</td>
<td>172</td>
<td>2</td>
<td>98</td>
<td>21</td>
</tr>
<tr>
<td>450</td>
<td>44</td>
<td>0</td>
<td>100</td>
<td>21</td>
</tr>
<tr>
<td>500</td>
<td>60</td>
<td>0</td>
<td>100</td>
<td>13</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>3187</strong></td>
<td><strong>56</strong></td>
<td><strong>44</strong></td>
<td><strong>7</strong></td>
</tr>
</tbody>
</table>
Figure 5.1 Material Take-Off 1: Distribution of valve numbers with respect to pipe diameter. The lines indicate differing degrees of piping complexity as defined by Hooper [17], where the most complex is shown as $30/D$, i.e. number of valves $= 30$/pipe diameter.
Figure 5.2 Material Take-Off 2: Distribution of valve numbers with respect to pipe diameter for all piping systems
Figure 5.3 Material Take-Off 3: Distribution of valve numbers with respect to pipe diameter for process gas and liquid piping systems
6. Inventory of major plants

As outlined in the project strategy in Section 3 a prime objective of the study was the modelling of ignition frequency for a representative process plant. In order to cross-check the modelled estimates of ignition frequency against statistical fire and explosion data it was necessary to estimate the number of plants where there is a risk of major fire and explosion. Data were derived from literature sources concerning the following installation types:

- Petrochemical/chemical plants
- Refinery units
- LPG installations
- Natural gas installations

For chemical and petrochemical processes the global number of plants was estimated using the Directory of Chemical Producers [20]. For the years 1986-1988 the number of plants in the United States, Western Europe and United Kingdom was estimated as:

<table>
<thead>
<tr>
<th></th>
<th>Number of plants</th>
</tr>
</thead>
<tbody>
<tr>
<td>United States</td>
<td>1420</td>
</tr>
<tr>
<td>Western Europe</td>
<td>1455</td>
</tr>
<tr>
<td>United Kingdom</td>
<td>185</td>
</tr>
</tbody>
</table>

These represent the numbers of plants handling some 130 different flammable materials (the Western Europe figure includes the UK figure).

A yearly update of the global number of refineries is provided by the International Petroleum Encyclopaedia [21]. For 1988 the number of refineries in the countries in the above categories were as follows:
Figures for the years 1982, 1984 and 1985 are also presented in Table 12.4 for US refineries only. A refinery normally consists of a series of plants, or processing units. For this study it was assumed that each refinery consists of an average of five individual processing units, each of which is equivalent to one chemical plant. For the purposes of this Section these are termed refinery unit-equivalent plants. The total number of refinery unit-equivalent plants (i.e. refinery units and chemical process plants) from these two plant categories was therefore estimated as:

<table>
<thead>
<tr>
<th></th>
<th>United States</th>
<th>Western Europe</th>
<th>United Kingdom</th>
</tr>
</thead>
<tbody>
<tr>
<td>Number of refineries</td>
<td>187</td>
<td>109</td>
<td>15</td>
</tr>
</tbody>
</table>

United States  $1420 + (5 \times 187) = 2355$
Western Europe $1455 + (5 \times 109) = 2000$
United Kingdom $185 + (5 \times 15) = 260$

The numbers of LPG and natural gas installations in the UK were estimated using data of Pape and Nussey [22]. These listed the following numbers of installations notifiable to the HSE under the Notification of Installations Handling Hazardous Substances (NIHHS) regulations (1982):

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>LPG installations</td>
<td>600</td>
</tr>
<tr>
<td>Natural gas installations</td>
<td>400</td>
</tr>
</tbody>
</table>

For this study, it was necessary to adjust these figures to account for the number of installations where it was judged that there was no significant fire risk. For example, the number of LPG installations was reduced to 450 by disregarding those with a storage capacity of less than 25 tonnes. However, the modified figure still includes a significant proportion of installations which by themselves do not constitute "plant", such as cylinder stores. The number of natural gas installations also consists mainly of low pressure storage holder sites. Notwithstanding these considerations the overall number of installations in the UK constituting a major fire and explosion hazard was estimated as:
<table>
<thead>
<tr>
<th>Installation Type</th>
<th>Quantity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chemical/refinery units</td>
<td>260</td>
</tr>
<tr>
<td>LPG installations</td>
<td>450</td>
</tr>
<tr>
<td>Natural Gas installations</td>
<td>400</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>1110</strong></td>
</tr>
</tbody>
</table>

The ratio of total installations: chemical/refinery installations was determined and used as a scaling factor to estimate the total numbers of installations at risk in the US and Western Europe:

- **Total installations: Chemical/refinery units (UK):** $\frac{1110}{260} = 4.3$
- **Number of major plants (US):** $4.3 \times 2000 = 10127$
- **Number of major plants (Western Europe):** $4.3 \times 2355 = 8600$
- **Total number in Western Europe and US:** $10127 + 8600 = 18727$

It proved necessary to correct the above estimates for the numbers of LPG and natural gas installations. From the fire and explosion incident analysis carried out in Section 12 it became apparent that very few, if any, incidents occur on gas storage sites. It was also judged from the explosion data given in Section 12 that some of the chemical/petrochemical plants seemed to have a low explosion risk. The number of refinery unit-equivalent plants in the UK was accordingly reduced by 30% to a final estimate of 180 (the standard plant formulated in the forthcoming Sections was designed to be equivalent to a refinery unit). The revised figures for Western Europe and the US were then derived as follows:

- **Refinery unit-equivalent plants: Total installations (UK):** $\frac{180}{1110} \approx 0.16$
- **Number of refinery unit-equivalent plants (US):** $0.16 \times 10127 = 1620$
- **Number of refinery unit-equivalent plants (Western Europe):** $0.16 \times 8600 = 1376$
- **Total number in Western Europe and US:** $1620 + 1376 = 2996$

The above estimates involved several generalised assumptions such as the consistency of ratios for different types of plant. These may not be justified for individual plant types: however as an overall estimation method the accuracy was considered sufficient to use the above inventory figures as a basis for a preliminary investigation of ignition frequencies per plant in Section 12.
7. Failure and leak frequency of equipment

Section 5 described a method of evaluating the relative quantities of different principal emission sources on areas of process plant. To estimate the risk of fire and explosion presented by these potential sources of hazard it is necessary to know how often these leaks occur. This section deals with how the leak frequencies from various sources were assessed for this study.

7.1 Fixed frequency emission sources

The information in Table 5.1 shows that whilst some sources are constantly emitting and are of fixed size, others present a hazard at some set interval. Constant emission sources include:

Ventilation exhausts
Open drains
Open pools e.g. degreasing baths

Sources emitting at set intervals include:

Sample points
Drain points
Atmospheric vents

and also activities such as:

Tanker/vessel filling
Maintenance activities

The likely size of such emissions can readily be calculated and hazard distances estimated using the calculation models given in Section 9. The frequency and duration of set interval emission
sources can also be established by observation. The risk from such emissions could and should be minimised by the implementation of adequate operating procedures and permit-to-work systems within the defined hazard area for the period of emission. The use of hazardous area classification techniques is usually confined to situations where there is a constant risk of fire and explosion e.g. the risk of random leakage from pressurised process plant.

It should be noted that there is also some additional risk with the above emission sources if the probability of an extraordinary event occurring is included. Such events may be the over-filling of a storage vessel or the partial closing of a sample point. The probability of these events is hard to estimate because they are dependent on human error rather than actual equipment failure and consequently the frequency estimates are based on engineering judgement rather than actual incident data. However, it was felt that the frequency of such events could also be controlled by the use of adequate procedures and by designing the installation so that opportunities for accidents would be minimal. Whilst it may therefore be possible to model the consequences of such accidents, the frequency of these events is dependent on individual contributing circumstances which may have to be evaluated separately. The control of emissions which occur as a result of some activity rather than an equipment leak is discussed further in Section 13 concerning indoor plant. This study was primarily concerned with classifying areas where unplanned emissions may occur in the form of leaks from equipment such as valves and pumps.

7.2 Leak frequency of common process items

A literature search was conducted for published data concerning equipment leak frequency. Data were found for the following items:

- Pipework
- Flanges/gaskets
- Valves
- Pumps
- Compressors
- Pressure relief valves
- Small-bore connections
For this study vessel failure was not considered beyond the failure of gaskets on manholes and nozzles.

The data given for the above items comprised a mixture of overall observed failure frequency data with a breakdown of failure modes and "best-guess" estimates used for hazard assessment work. These were both collected and presented so that overall best-guesses could be made for each equipment type. It was not intended to produce absolute failure frequencies for each component, since it was felt that such detailed analysis was beyond the scope of the project. Instead, the data in the following sections were used to identify the order of magnitude in which each leak frequency may be expected to lie. Together with the equipment inventories and leak size information the derived estimates were used as described in Section 12 to model the relative ignition risk presented by each equipment type. The individual risks may then be consolidated to produce an overall ignition risk estimate for a collection of emission sources such as may be found in an area of process plant. The use to which the surveyed data were put is described further in Section 7.3.

To allow for variations in sampling by the different authors, and to show the degree of uncertainty that is involved with the failure frequency data presented here, the experimentally observed failure frequencies were confidence tested using the $\chi^2$ distribution. The 90% limits are shown where applicable. The $\chi^2$ confidence test is described in Appendix C.

Some information has also been published concerning leak size distributions for some of the leak sources considered. This is also listed in the forthcoming section, although the overall strategy for assigning probable leak hole sizes is laid out in Section 8.

7.2.1 Pipework

It was found that published pipework failure data were rather awkward to use. Much of it is given as estimated pipework failures per plant or per section of pipework, which gave difficulty in interpretation because the length of pipework involved in each case is seldom specified. It was also not clear in some instances whether the failure rate given for piping included pipework flanges and fittings. To avoid confusion, attention was focused on data sources which
gave failure data as failures per unit length. It was assumed that the pipework failure rates referred to here applied to failures of pipe only. Note that in this instance pipework failure frequency is synonymous with pipework leak frequency.

First Canvey Report [23]

Failure frequency for 3 km LPG pipework ≥ 150 mm nominal bore

\[
\begin{align*}
5 \times 10^{-3}.\text{year}^{-1} &= 0.167 \times 10^{-5}.\text{m}^{-1}.\text{year}^{-1}
\end{align*}
\]

Bush [24]

The following estimate was based on eight pipe leak incidents on nuclear reactors and averaged failure rates from the American Atomic Energy Commission Reactor Safety Study (WASH 1400) [25]

- Pressurised Water Reactor pipework = \(4.6 \times 10^{-5}.\text{m}^{-1}.\text{year}^{-1}\)
- Boiling Water Reactor pipework = \(4.3 \times 10^{-5}.\text{m}^{-1}.\text{year}^{-1}\)
- Combined = \(4.3 \times 10^{-5}.\text{m}^{-1}.\text{year}^{-1}\)

Welker and Schoor [26]

- Maximum failure frequency for flanged and welded LNG pipework = \(6.6 \times 10^{-10}.\text{ft}^{-1}.\text{hour}^{-1}\) = \(1.9 \times 10^{-5}.\text{m}^{-1}.\text{year}^{-1}\)

Welker and Johnson [27]

- Failure frequency for welded LNG pipe = \(17 \times 10^{-10}.\text{ft}^{-1}.\text{hour}^{-1}\) = \(4.9 \times 10^{-5}.\text{m}^{-1}.\text{year}^{-1}\)
Pape and Nussey [22]

Guillotine fracture frequency for 25 mm chlorine pipe

As above; pipe split frequency

Rijnmond report [28]

Pipework leak frequencies were given per metre.hour. These are converted below into metre.years (one year consists of 8760 hours).

<table>
<thead>
<tr>
<th>Pipe diameter (mm)</th>
<th>Leak frequency</th>
<th>Leak size</th>
<th>Description of leak</th>
<th>Frequency (m⁻¹.year⁻¹)</th>
<th>Percent of total</th>
</tr>
</thead>
<tbody>
<tr>
<td>≥ 50</td>
<td>8.76 × 10⁻⁷</td>
<td>A</td>
<td>Rupture</td>
<td>2.2 × 10⁻⁴/D</td>
<td>4</td>
</tr>
<tr>
<td>50 - 150</td>
<td>2.63 × 10⁻⁷</td>
<td>0.2.A</td>
<td>Catastrophic</td>
<td>7.5 × 10⁻⁴/D</td>
<td>15</td>
</tr>
<tr>
<td>≤ 150</td>
<td>8.76 × 10⁻⁸</td>
<td>0.05.A</td>
<td>Big</td>
<td>12 × 10⁻⁴/D</td>
<td>25</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.01.A</td>
<td>Small</td>
<td>28 × 10⁻⁴/D</td>
<td>56</td>
</tr>
</tbody>
</table>

The industrial comment on the Rijnmond report also quoted a leak size distribution for pipework. This was based on Gulf failure data and was also quoted by Hawksley [29]:

This is in broad agreement with work by Blything and Parry [30] who analysed pipe failure data for one particular chemical plant and found that 5% of failures were guillotine fractures.
7.2.2 Flange/gasket failures

Data collected under this heading cover failure of both flange and connecting gasket, although most estimates given below relate to gasket deterioration or bad fitting. Some information was also available concerning likely leak size, from which it was occasionally possible to infer whether the gasket being referred to was of a reinforced type or not. The difference in failure modes between gasket types is discussed further in Section 8.

Pape and Nussey [22]

Frequency of a gasket section between two restraining bolts being blown out:

\[
\begin{align*}
0.6 \text{ mm thick gasket} & \quad = \quad 3 \times 10^{-6}\text{gasket}^{-1}\text{year}^{-1} \\
3.0 \text{ mm thick gasket} & \quad = \quad 5 \times 10^{-6}\text{gasket}^{-1}\text{year}^{-1}
\end{align*}
\]

Batstone and Tomi [31]

The authors conducted a literature survey from which the following failure frequencies for full bore pipe connection ruptures (i.e. where the total cross-sectional area of the pipe became the leak hole) were compiled:

<table>
<thead>
<tr>
<th>Pipe diameter</th>
<th>Failure frequency $\times 10^6$ connection$^{-1}$ year$^{-1}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\leq 25 \text{ mm}$</td>
<td>30</td>
</tr>
<tr>
<td>40 mm</td>
<td>10</td>
</tr>
<tr>
<td>50 mm</td>
<td>7.5</td>
</tr>
<tr>
<td>80 mm</td>
<td>5</td>
</tr>
<tr>
<td>100 mm</td>
<td>4</td>
</tr>
<tr>
<td>$\geq 150 \text{ mm}$</td>
<td>3</td>
</tr>
</tbody>
</table>

These figures also include ruptures of the piping itself.
Pipe joint/gasket failure frequency estimate

given by UKAEA

\[= 0.5 \times 10^{-6} \text{gasket}^{-1} \text{hour}^{-1}\]

\[= 4.4 \times 10^{-3} \text{gasket}^{-1} \text{year}^{-1}\]

American Atomic Energy Commission Reactor Safety Study (WASH 1400) [25]

Average gasket failure frequency in operational mode

within range

\[= 0.03 \text{gasket}^{-1} \text{year}^{-1}\]

0.001 - 0.1

Smith [33]

The average gasket failure frequency is assumed to lie between the limits:

\[1.8 - 44 \times 10^{-4} \text{gasket}^{-1} \text{year}^{-1}\]

British Gas Engineering Standard PS/SHA1 [34]

An estimate of \(5 \times 10^{-4}\) failures per year was suggested for a maximum credible leak hole size of 0.25 mm\(^2\) for all types of gasket handling natural gas.

Halogenated hydrocarbon plant [19]

Another estimate of typical gasket leak frequency concerned the halogenated hydrocarbon plant first considered in Section 5.2.1. The leak frequency was based on data provided by a reliability data bank and modified according to the experience of the operator.

\[= 5 \times 10^{-3} \text{gasket}^{-1} \text{year}^{-1}\]
7.2.3 Valves

Available information on valve type and size is given below. The data sources that only give overall failure frequency have been omitted as this study was concerned with available leak frequencies only.

OREDA-84 databook [35]

The Offshore Reliability Databook gives calculated failure frequencies for various types of remotely actuated valves used in the North Sea. These are listed in Table 7.1.

Aupied, LeCouiec and Procaccia [36]

The authors investigated the valve and pump failures occurring on French nuclear power plant. The valve failures are summarised below whilst the pump failures are given in Section 7.2.4.

Type: Control, shut off, non-return and relief valves (manual and remotely actuated)
Handling: Water and condensate
Size: various

Failure frequency in operation = \( 29 \times 10^{-2} \text{. valve}^{-1} \text{. year}^{-1} \)
90% confidence limits = \( 26 - 32 \times 10^{-2} \text{. valve}^{-1} \text{. year}^{-1} \)

Leak frequency (estimated) = \( 8.7 \times 10^{-2} \text{. valve}^{-1} \text{. year}^{-1} \)
90% confidence limits = \( 8.2 - 11 \times 10^{-2} \text{. valve}^{-1} \text{. year}^{-1} \)

British Gas plant leakage studies [37],[38]

Two LNG plants were monitored for component leakages. The following leak frequencies were obtained for a mixture of line valves:
<table>
<thead>
<tr>
<th></th>
<th>Plant 1 [37]</th>
<th>Plant 2 [38]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cryogenic</td>
<td>0.70</td>
<td>0.47</td>
</tr>
<tr>
<td>Non-cryogenic</td>
<td>1.4</td>
<td>0.55</td>
</tr>
<tr>
<td>Combined</td>
<td>1.1</td>
<td>0.49</td>
</tr>
</tbody>
</table>

**Systems Reliability Service databank [39]**

A selection of typical valve failure frequencies from the SRS databank is given in Table 7.2 as examples of typical information available from this source. As with the OREDA-84 data there is a wide spread of failure frequencies within the confidence limits given, which meant that it was difficult to identify the order of magnitude within which an average failure frequency may lie. However, although the failure frequencies show little consistency the leak frequencies for the various valve types not handling steam or water are of the same order of magnitude. Also, the leak frequencies for actuated valves do not seem significantly different to those of manual valves. This may enable considerable simplification when considering the average leak frequency for a variety of valve types.

Some examples of valve leak frequency estimates were also obtained where industrial operators made their own "in-house" estimates of average valve leak frequency for hazard assessment use. These estimates, listed below, were usually based on SRS data and subsequently adjusted according to the operators' own experience with the valve type in question. It was found that this approach was also employed for other items such as pumps where there are a relatively large amount of observed failure data available.

**Smith [33]**

Smith listed average valve leak frequencies per $10^6$ hours assuming that external leaks form 15% of the overall failures, and that the valves are not at high temperature or handling highly corrosive substances. The figures have been converted below into leaks per year by dividing by 8760. Where two figures are given these represent the probable upper and lower limits, whilst where three figures are given the middle figure represents the most widely used figure in the source data banks used.
<table>
<thead>
<tr>
<th>Type</th>
<th>Leak frequency $\times 10^2$ valve$^{-1}$ year$^{-1}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ball</td>
<td>0.13 - 0.46</td>
</tr>
<tr>
<td>Butterfly</td>
<td>0.13 - 2.6 - 3.9</td>
</tr>
<tr>
<td>Gate</td>
<td>0.19 - 1.9</td>
</tr>
<tr>
<td>Needle</td>
<td>0.19</td>
</tr>
<tr>
<td>Plug</td>
<td>0.13 - 1.1</td>
</tr>
</tbody>
</table>

**British Gas Engineering Standard PS/SHA1 [34]**

British Gas suggested the following leak frequency for all types of valve gland resulting in a leak hole size of 0.25 mm$^2$:

Valve gland leak frequency $= 0.1 \times 10^2$ valve$^{-1}$ year$^{-1}$

This estimate applies to gland leaks only. On large valves there may be other leak sites such as grease nipples and drain valves which are not included.

**Rijnmond Report [28]**

The Rijnmond report used leak frequency data principally derived from WASH 1400 which dealt with complete valve rupture.

For:
- Manual valves
- Air - fluid operated valves
- Motorised valves
- Solenoid operated valves

Rupture frequency $= 0.0088 \times 10^2$ valve$^{-1}$ year$^{-1}$

within range $= 0.00088 - 0.088 \times 10^2$ valve$^{-1}$ year$^{-1}$
Halogenated hydrocarbon plant [19]

A valve leak frequency was estimated for the halogenated hydrocarbon plant mentioned in Section 7.2.2. As with the gasket leak frequency estimate, this estimate was also based on published data augmented by operational experience.

\[
\text{Valve failure frequency} = 1.8 - 10.8 \times 10^2 \text{.valve}^{-1} \text{.year}^{-1}
\]

assuming 50% are leaks

\[
\text{Leak frequency adopted} = 3 \times 10^2 \text{.valve}^{-1} \text{.year}^{-1}
\]

Wallace [15]

In a study of fugitive (i.e. minor) emissions from process plant, Wallace estimated the leak frequency of a mixture of valve types to be $6 \times 10^2 \text{.valve}^{-1} \text{.year}^{-1}$.

7.2.4. Pumps

The equipment failure data survey also involved the collection of failure frequency data for centrifugal and reciprocating pumps. It was usually possible to differentiate between centrifugal pump seal types from examination of the literature and failure information; however it was tentatively assumed that pumps listed below as handling hazardous material have mechanical seals whilst the rest use packed gland seals.

Aupied, LeCouiec and Procaccia [36]

The authors used a French databank to produce a calculated failure frequency for nuclear reactor feedwater pumps. These were centrifugal pumps with mechanical tungsten carbide/graphite seals.

\[
\text{Pump failure frequency in operation} = 4.9 \text{.pump}^{-1} \text{.year}^{-1}
\]

\[
\text{Pump leak frequency} = 1.4 \text{.pump}^{-1} \text{.year}^{-1}
\]
These data were from a reliability study of 85 centrifugal pumps handling ethylene on a petrochemical plant over a 19 month period.

Pump failure frequency
90% confidence limits

Pump leak frequency
90% confidence limits

Halogenated hydrocarbon plant [19]

The leak frequencies for three banks of pumps were calculated using maintenance data and listed as follows:

Bank 1 (Column reflux - 3 pumps) = 0.27 .pump\(^{-1}\).year\(^{-1}\)
Bank 2 (System feed - 2 pumps) = 2.9 .pump\(^{-1}\).year\(^{-1}\)
Bank 3 (Column feed - 2 pumps) = 6.3 .pump\(^{-1}\).year\(^{-1}\)

OREDA-84 databook [35]

Table 7.3 contains pump failure data from this source.

Systems Reliability Service databank [39]

A selection of pump failure data are presented in Table 7.4 which were felt to be typical of the large amount available from this source.

British Gas plant leak studies [37],[38]

The leak frequency of the LNG pump on two LNG storage sites was calculated to be as follows:
Pump leak frequency

<table>
<thead>
<tr>
<th>Plant 1 [37]</th>
<th>Plant 2 [38]</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.022</td>
<td>0.035</td>
</tr>
</tbody>
</table>

90% confidence limits

<table>
<thead>
<tr>
<th>Plant 1 [37]</th>
<th>Plant 2 [38]</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.004 - 0.070</td>
<td>0.012 - 0.080</td>
</tr>
</tbody>
</table>

Smith [33]

The author presented estimates of average leak frequency in a similar form to those given above for valves. Smith estimated that 50% of the number of failures result in leaks.

Centrifugal pumps

Leak frequency (pump\(^{-1}\).year\(^{-1}\))

0.044 - 0.13 - 0.35

First Canvey Report [23] and Rijnmond Report [28]

The following estimate of pump rupture (i.e. the rupture of a joint) was made from SRS data and used for both reports:

Catastrophic failure frequency

\[ = 10^4 \text{pump}^{-1}\cdot\text{year}^{-1} \]

Wallace [15]

Fugitive emission frequency of pump seals

\[ = 0.12 \text{pump}^{-1}\cdot\text{year}^{-1} \]

7.2.5 Compressors

It proved difficult to derive meaningful data for compressor leak frequency from published data sources. In the plant fugitive emission study mentioned above, Wallace [15] estimated the fugitive emission frequency of gas compressors to be 0.07 per compressor per year, whilst Kletz [97] estimated the probability of a major leak on a reciprocating compressor as 0.1 per plant per year for typical petrochemical installations. Johnson and Welker [27] reported three compressor crankcase explosions from a study of 27 LNG installations covering a total period of 717 operating years. The overall minor failure frequency was given as 2.0 per compression system per year (Johnson and Welker classified failures as major or minor according to the time taken
for repair). Whilst no actual leak frequencies were provided it was inferred here that this at least indicated that compressors present a formidable explosion risk. Another indication that compressor failure frequencies are relatively high compared to other leak sources was obtained from the SRS data listed in Table 7.5, where the leak frequencies are of the same order of magnitude as the Kletz estimate.

7.2.6. Pressure relief valves

Pressure relief valves become a source of unplanned emissions if they lift before the set pressure is reached or if they fail to seat correctly. Estimated frequencies for these occurrences were obtained and are listed below:

British Gas Engineering Standard PS/SHA1 [34]

For clean applications, British Gas suggested a relief valve leak frequency of 0.02.valve^{-1}.year^{-1}.

Lawley and Kletz [41]

Pressure relief valve lift light frequency was estimated to be 0.02.valve^{-1}.year^{-1}.

Rijnmond Report [28]

\[
\text{Lift light frequency within range} = 0.06.\text{valve}^{-1}.\text{year}^{-1} \\
0.02 - 0.09.\text{valve}^{-1}.\text{year}^{-1}
\]

Wallace [15]

Pressure relief valve fugitive emission frequency was estimated to be 0.07.valve^{-1}.year^{-1}

Halogenated hydrocarbon plant [19]

This estimate of relief valve failure frequency was made using maintenance data for plant handling natural gas:
Relief valve passing               =  0.109.valve\(^{-1}\).year\(^{-1}\)
Relief valve lifting light        =  0.03.valve\(^{-1}\).year\(^{-1}\)

**Systems Reliability Service databank [39]**

Representative data from the SRS databank for spring-loaded pressure relief valves are given in Table 7.6. This lists failure frequencies only; however, the Rijnmond Report estimated that about 90% of relief valve failures consist of the valve lifting light or passing. If this factor is applied to the data in Table 7.6 it becomes apparent that these frequencies also lie within one order of magnitude i.e. 0.01 - 0.1.valve\(^{-1}\).year\(^{-1}\).

### 7.2.7. Small-bore connections

It was felt that the deterioration and fracture of small-bore pipework and connections may contribute significantly to the overall risk of fire and explosion. This is reflected for example in the current practice of specifying the use of flow restrictors in impulse lines on petrochemical plant under construction and modification. Other small-bore leak sources include hoses and hose fittings, compression fittings and screwed connections on the ends of lines.

It proved difficult to obtain leak frequency data for these sources. Such information is scarce because it is often compounded with other items e.g. failures of connections on transmitters may be listed under transmitter failures. Leak frequencies were however inferred from similar information concerning air lines. For example, the Rijnmond Report [28] used estimated fracture frequencies of the same order of magnitude for instrument impulse lines and air lines:

<table>
<thead>
<tr>
<th>Frequency</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Frequency of air line fracture</td>
<td>0.01.line(^{-1}).year(^{-1})</td>
</tr>
<tr>
<td>Frequency of impulse line fracture</td>
<td>0.03.line(^{-1}).year(^{-1})</td>
</tr>
<tr>
<td>Frequency of impulse line blockage</td>
<td>0.06.line(^{-1}).year(^{-1})</td>
</tr>
<tr>
<td>Range for overall frequency of impulse line failure</td>
<td>0.09 - 0.26 .line(^{-1}).year(^{-1})</td>
</tr>
</tbody>
</table>

Further data from the SRS data bank of a higher order of magnitude for overall impulse line failure frequency are given below for three samples:
It was necessary to contrast the failure frequency of the small-bore pipe used for impulse lines with the failure frequency of the connections used since they were quite different. The SRS databank also contains information on compression fittings, some of which is given below:

<table>
<thead>
<tr>
<th>Sample</th>
<th>1</th>
<th>2</th>
<th>3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Failure rate .line⁻¹.year⁻¹</td>
<td>0.92</td>
<td>0.98</td>
<td>0.086</td>
</tr>
<tr>
<td>90% upper confidence limit</td>
<td>1.0</td>
<td>1.3</td>
<td>0.16</td>
</tr>
<tr>
<td>90% lower confidence limit</td>
<td>0.84</td>
<td>0.75</td>
<td>0.043</td>
</tr>
</tbody>
</table>

The above data were selected because they cover a wide range of values and so illustrate the danger of selecting any one value as "average". However, it was noted that most of the data for compression fittings in the databank were of the same order of magnitude as Sample 3. Other comparable data on small-bore fittings may be found in the British Gas code of practice PS/SHA1 which suggested leak frequencies of 0.00005 and 0.0008.connection⁻¹.year⁻¹ for 0.25 mm² leak holes in compression fittings and screwed connections respectively. Also, the Batstone and Tomi estimates given in Section 7.2.2 implied a comparable joint rupture frequency for pipework less than 25 mm diameter of 0.00003 .connection⁻¹ .year⁻¹.

It was therefore concluded that the difference in failure frequencies between small-bore line rupture and small-bore connection rupture indicated that the main hazard lies in fractures of the line itself rather than of the connections. The failure rate of flexible hoses is commonly estimated to be of the same order of magnitude as for impulse lines. For example, the Rijnmond Report adopted the figures of Green and Bourne, who estimated that the probable failure frequency for hoses lay between 0.035 and 0.35 .hose⁻¹ .year⁻¹.

7.3 Application of surveyed data

The foregoing data were assembled with a view to providing estimates of typical leak frequency for items of process equipment. However, the frequency values given show that there
is a considerable spread from which to select individual working estimates. Furthermore, the wide confidence limits associated with the experimentally derived values illustrate the somewhat poor quality of the available data. During the data survey it became apparent that for a given equipment item the leak frequency may vary according to:

1. **Type:** Some valve and pump seal types are inherently more prone to leakage than others due to differences in design. A simple example is that packed gland pump seals are more likely to leak than mechanical seals. This is enlarged upon in Section 8. It is also suspected that valve leak frequency may vary according to manufacturer.

2. **Duty:** It is generally felt that equipment in a hostile environment is more prone to leaks than equipment sited, say, indoors. This broad statement is supported by the high leak frequencies of off-shore equipment indicated by the OREDA data in Tables 7.1 and 7.3. Also, it is felt that equipment such as valves which are operated constantly and regularly are more likely to leak than other equipment which is on standby or not otherwise regularly operated.

Both these factors are well-known; however there are few current data available with which to quantify the effects of these factors on leak frequency. The results of a study by Anyakora et al. concerning the effect of environment on process plant instrument failures was published in 1971 [103] and drew speculative conclusions concerning the quantification of these factors. However it was apparent from the researched data that this subject has not been progressed.

Considering the scarcity of relevant leak frequency data it was therefore decided to adopt values for further analysis based largely on the judgemental estimates given here. The approach adopted for the investigation of the effect of leak frequency on the flammable risk presented by a leak source is described in detail in Section 12. Suffice here to say that it was primarily based on applying leak frequency information to a formulated "standard" plant derived from the inventory information given in Material Take-Off 1 in Section 5. Equipment leak frequencies were assigned according to the size of the equipment and the magnitude of the leak. The rationale for adopting initial frequency values was as follows.

As detailed in Section 12 the standard plant was assumed to consist of pipework of nominal bore 25 mm, 50 mm, 150 mm and 300 mm. For each of these sizes a basic leak frequency was established according to equipment type and which corresponded to a *minor* non-catastrophic leak. Two additional leak levels were then specified for each basic value for *major*
non-catastrophic leaks and for catastrophic leaks (implying complete fracture or *rupture* of the item concerned). Note that although HAC is not principally concerned with catastrophic failures such as vessel ruptures it was necessary in this study to consider the frequency of such events where the main risk lay in the subsequent ignition of the release. Such consideration was necessary in order to assess the ignition frequency of such events compared to minor releases.

The value of the leak *severity fraction*, or the ratio of the frequencies of major leaks to minor leaks, was assumed to be 1:10:100 for rupture:major:minor leaks. This is broadly similar to the Pape and Nussey and Rijnmond estimates given in Section 7.2.1 for pipe, but which was also assumed here for all equipment types here as a preliminary estimate. The basic leak frequency estimates adopted for each source type were as follows:

**Pipework**

Pipe leak frequency (leaks/metre/year \times 10^4)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.01</td>
<td>0.01</td>
<td>0.003</td>
<td>0.001</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.10</td>
<td>0.10</td>
<td>0.030</td>
<td>0.010</td>
</tr>
<tr>
<td>Minor leak</td>
<td>1.00</td>
<td>1.00</td>
<td>0.300</td>
<td>0.100</td>
</tr>
</tbody>
</table>

The initial estimates of leak frequency for each item were not sophisticated. For example, the above values are essentially the same as those given in the Rijnmond Report. The refinement of these and the proceeding initial values is described in Section 12.

**Valves**

After consultation with the various industrial supporters of the IIGCHL, it was felt that the high leak frequencies indicated in Section 7.2.2 were unrealistic of typical process conditions. For example, Table 7.2. indicates that steam and water valves are particularly prone to leakage, a conclusion which may also be drawn from the work of Aupied et al. concerning steam and water valves in French nuclear plant. However, this study was not particularly concerned with steam leaks, and so the basic values adopted were based more on the estimates of Smith, British Gas and the Rijnmond Report:
Valve leak frequency (leaks/valve/year \( \times 10^4 \))

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.1</td>
<td>0.1</td>
<td>0.1</td>
<td>0.1</td>
</tr>
<tr>
<td>Major leak</td>
<td>1.0</td>
<td>1.0</td>
<td>1.0</td>
<td>1.0</td>
</tr>
<tr>
<td>Minor leak</td>
<td>10.0</td>
<td>10.0</td>
<td>10.0</td>
<td>10.0</td>
</tr>
</tbody>
</table>

Flanges/Gaskets

It was similarly felt that the gasket leak frequencies given in Section 7.2.2 were not typical of gasket performance on current process plant. Consequently the leak frequencies adopted for further analysis were based on the more conservative estimates of Smith, British Gas, and Pape and Nussey (the absence of a catastrophic gasket failure mode is explained in Section 8.2).

<table>
<thead>
<tr>
<th>Flange leak frequency (leaks/flange/year ( \times 10^4 ))</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipe diameter (m)</td>
</tr>
<tr>
<td>Major leak</td>
</tr>
<tr>
<td>Minor leak</td>
</tr>
</tbody>
</table>

Pumps

The pump leak frequencies given in Section 7.2.4 are the highest of all the equipment items listed. Again this was felt to be unrepresentative of actual conditions since the estimate given by Wallace indicated that such frequent leaks are very small fugitive emissions where the main concern is the pollution effect rather than the risk of ignition. Fugitive emissions are considered further in Section 8.7. It was thus considered that more serious leaks such as may result in ignition were much less frequent, and were of the order of the British Gas plant leakage studies:

<table>
<thead>
<tr>
<th>Pump leak frequency (leaks/pump/year ( \times 10^4 ))</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipe diameter (m)</td>
</tr>
<tr>
<td>Rupture leak</td>
</tr>
<tr>
<td>Major leak</td>
</tr>
<tr>
<td>Minor leak</td>
</tr>
</tbody>
</table>
The particular value of three was adopted in order to relate pump leak frequency to other equipment leaks.

**Small-bore connections**

The frequency of rupture (termed here major) leaks and minor leaks was considered using the data in Section 7.2.7. Here the initial basic leak frequency value was derived from data for the more serious leak instead of (as in the above cases) the minor leak, as the majority of available data related to this case. The major leak frequency was therefore taken to be similar to the air-line rupture frequencies indicated earlier:

<table>
<thead>
<tr>
<th>Small bore connection leak frequency (leaks/connection/year × 10⁴)</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>(diameter = 0.01 m)</td>
<td></td>
</tr>
<tr>
<td>Major leak</td>
<td>5.0</td>
</tr>
<tr>
<td>Minor leak</td>
<td>50.0</td>
</tr>
</tbody>
</table>

Having thus identified a crude band of leak frequencies with which to commence investigation of the overall leak frequency on a typical process plant it was not felt worthwhile to progress with the residual leak frequency information presented in this section i.e. relief valve and compressor leaks. This was for two main reasons. Firstly, it was not felt that relief valve "lift-light" failures were directly relevant to this study. Although they are classified as unplanned emissions they are almost always vented through a stack where possible emissions should already be analysed for dispersion effects since the stack may be expected to emit at set intervals as part of normal procedure. Secondly, regarding compressor failures, there were very little data available on which to base the selection of even a crude initial estimate. It should however be noted that the procedure for investigating the ignition frequency of a collection of emission sources described in Section 12 may be expanded at some future date to include other emission sources to those considered here.

To summarise, a literature survey was carried out for leak frequencies of typical process items. This indicated a wide range of values which may apply for a given equipment item, which made it difficult to justify the choice of a single value as a universal working estimate. Provisional estimates were therefore made of leak frequency according to equipment size and leak severity for selected equipment items. Section 8 relates leak severity to leak hole size.
Table 7.1 OREDA-84 Remotely actuated valve failure frequencies [35].

Confidence limits are 90%  
Leak and failure frequencies $\times 10^3 \text{valve}^{-1} \text{year}^{-1}$

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit</th>
<th>Confidence limit</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Lower</td>
<td>Upper</td>
</tr>
<tr>
<td>Type: Shut off; hydraulically operated globe/gate/ball valve and actuator</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Size: $&lt; 600$ mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrocarbon gas</td>
<td>83</td>
<td>60</td>
<td>105</td>
</tr>
<tr>
<td>Type: as above</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Size: $&gt; 600$ mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrocarbon gas</td>
<td>69</td>
<td>39</td>
<td>96</td>
</tr>
<tr>
<td>Type: Pneumatically operated ball valve and actuator</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Size: 150 mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Crude oil, gas</td>
<td>7.9</td>
<td>4.3</td>
<td>11</td>
</tr>
<tr>
<td>Type: Shut off; pneumatically operated globe/gate/ball valve and actuator</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Size: $&lt; 600$ mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Liquid</td>
<td>52</td>
<td>3.3</td>
<td>105</td>
</tr>
<tr>
<td>Type: Control; hydraulically operated globe/gate valve and actuator</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Size: $&lt; 600$ mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrocarbon gas</td>
<td>46</td>
<td>39</td>
<td>53</td>
</tr>
<tr>
<td>Liquid</td>
<td>88</td>
<td>44</td>
<td>140</td>
</tr>
<tr>
<td>Type: Control; pneumatically operated globe valve and actuator</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Size: $&lt; 600$ mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrocarbon gas</td>
<td>53</td>
<td>43</td>
<td>64</td>
</tr>
</tbody>
</table>
Table 7.2 Systems Reliability Service failure frequency data for various valve types

Confidence limits are 90%
Leak and failure frequencies $\times 10^2$ valve$^{-1}$ year$^{-1}$

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit Lower</th>
<th>Confidence limit Upper</th>
<th>Leak frequency Lower</th>
<th>Leak frequency Upper</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Type: Gate; manual</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>3.0</td>
<td>1.3</td>
<td>5.9</td>
<td>1.0</td>
<td>0.2</td>
</tr>
<tr>
<td>Water</td>
<td>49</td>
<td>29</td>
<td>76</td>
<td>11</td>
<td>2.8</td>
</tr>
<tr>
<td>Water</td>
<td>42</td>
<td>24</td>
<td>68</td>
<td>18</td>
<td>2.0</td>
</tr>
<tr>
<td>Type: Gate; motorised with actuator</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Steam</td>
<td>18</td>
<td>9.0</td>
<td>20</td>
<td>6.6</td>
<td>1.6</td>
</tr>
<tr>
<td>Water</td>
<td>59</td>
<td>38</td>
<td>89</td>
<td>17</td>
<td>5.0</td>
</tr>
<tr>
<td>Steam</td>
<td>107</td>
<td>78</td>
<td>145</td>
<td>28</td>
<td>14</td>
</tr>
<tr>
<td>Type: Globe; manual</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Chlorine</td>
<td>54</td>
<td>36</td>
<td>78</td>
<td>5.4</td>
<td>1.0</td>
</tr>
<tr>
<td>Type: Globe; pneumatic with actuator</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Argon</td>
<td>5.6</td>
<td>2.5</td>
<td>11</td>
<td>1.9</td>
<td>0.3</td>
</tr>
<tr>
<td>Chlorine</td>
<td>43</td>
<td>27</td>
<td>65</td>
<td>2.7</td>
<td>0.1</td>
</tr>
<tr>
<td>Type: Globe; motorised with actuator</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Liquid nitrogen</td>
<td>10</td>
<td>6.4</td>
<td>15</td>
<td>2.4</td>
<td>0.8</td>
</tr>
<tr>
<td>Helium</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>2.5</td>
<td>0.4</td>
</tr>
</tbody>
</table>

continued
Confidence limits are 90%
Leak and failure frequencies $\times 10^2 \text{valve}^{-1} \text{year}^{-1}$

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit</th>
<th>Leak frequency</th>
<th>Confidence limit</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Lower</td>
<td>Upper</td>
<td>Lower</td>
</tr>
<tr>
<td>Type: Tapered plug; manual</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>HCl</td>
<td>70</td>
<td>50</td>
<td>95</td>
<td>7.0</td>
</tr>
<tr>
<td>HCl</td>
<td>12</td>
<td>2.1</td>
<td>38</td>
<td>6.0</td>
</tr>
<tr>
<td>Type: Ball; remotely controlled without actuator</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>3.8</td>
</tr>
<tr>
<td>Type: Mixture; remotely actuated without actuator</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>2.3</td>
</tr>
</tbody>
</table>
Table 7.3  OREDA-84 Pump failure frequencies [35]

Confidence limits are 90%
Leak and failure frequencies pump^{-1}. year^{-1}

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit</th>
<th>Leak frequency</th>
<th>Confidence limit</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Lower</td>
<td>Upper</td>
<td></td>
</tr>
<tr>
<td>Type: Centrifugal; single stage</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>2.3</td>
<td>1.4</td>
<td>3.4</td>
<td>0.14</td>
</tr>
<tr>
<td>Type: Centrifugal; multi-stage</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>11</td>
<td>1.8</td>
<td>19</td>
<td>2.1</td>
</tr>
<tr>
<td>Crude oil</td>
<td>23</td>
<td>16</td>
<td>32</td>
<td>5.0</td>
</tr>
<tr>
<td>Crude oil</td>
<td>9.7</td>
<td>6.5</td>
<td>15</td>
<td>0.55</td>
</tr>
<tr>
<td>Type: Reciprocating; single stage</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Condensate</td>
<td>219</td>
<td>175</td>
<td>272</td>
<td>75</td>
</tr>
<tr>
<td>Glycol</td>
<td>31</td>
<td>24</td>
<td>39</td>
<td>7.9</td>
</tr>
</tbody>
</table>
Table 7.4 Systems Reliability Service failure frequency data for various pump types

Confidence limits are 90%
Leak and failure frequencies pump\(^{-1}\) year\(^{-1}\)

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit</th>
<th>Leak frequency</th>
<th>Confidence limit</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Lower</td>
<td>Upper</td>
<td>Lower</td>
<td>Upper</td>
</tr>
<tr>
<td><strong>Type: Centrifugal</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Sodium metasilicate</td>
<td>5.8</td>
<td>4.1</td>
<td>8.1</td>
<td>3.0</td>
</tr>
<tr>
<td>Sodium metasilicate</td>
<td>3.7</td>
<td>2.3</td>
<td>5.7</td>
<td>1.4</td>
</tr>
<tr>
<td>HCl</td>
<td>6.2</td>
<td>4.1</td>
<td>8.9</td>
<td>4.1</td>
</tr>
<tr>
<td>Water</td>
<td>4.0</td>
<td>2.8</td>
<td>5.4</td>
<td>1.8</td>
</tr>
<tr>
<td>Carbonate/Salt solution</td>
<td>11</td>
<td>9.1</td>
<td>13</td>
<td>6.3</td>
</tr>
<tr>
<td><strong>Type: Reciprocating</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Methanol</td>
<td>1.5</td>
<td>1.2</td>
<td>1.9</td>
<td>0.99</td>
</tr>
<tr>
<td>Sulphonated hydrocarbon</td>
<td>1.2</td>
<td>1.0</td>
<td>1.4</td>
<td>0.43</td>
</tr>
</tbody>
</table>
Table 7.5 Systems Reliability Service failure frequency data for compressors

Confidence limits are 90%
Leak and failure frequencies. \( \text{compressor}^{-1} \cdot \text{year}^{-1} \)

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit Lower</th>
<th>Confidence limit Upper</th>
<th>Leak frequency</th>
<th>Confidence limit Lower</th>
<th>Confidence limit Upper</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Type: Reciprocating; multi-stage</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Ammonia</td>
<td>2.5</td>
<td>0.99</td>
<td>5.3</td>
<td>0.50</td>
<td>0.026</td>
<td>2.4</td>
</tr>
<tr>
<td>Ammonia</td>
<td>5.7</td>
<td>4.2</td>
<td>7.5</td>
<td>0.50</td>
<td>0.059</td>
<td>1.3</td>
</tr>
</tbody>
</table>

Table 7.6 Systems Reliability Service failure frequency data for spring-loaded relief valves

Confidence limits are 90%
Leak and failure frequencies \( \times 10^2 \cdot \text{valve}^{-1} \cdot \text{year}^{-1} \)

<table>
<thead>
<tr>
<th>Handling</th>
<th>Failure frequency</th>
<th>Confidence limit Lower</th>
<th>Confidence limit Upper</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ammonia</td>
<td>16</td>
<td>15</td>
<td>17</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>13</td>
<td>6.1</td>
<td>24</td>
</tr>
<tr>
<td>Aromatic</td>
<td>6.5</td>
<td>4.1</td>
<td>11</td>
</tr>
<tr>
<td>Aromatic</td>
<td>12</td>
<td>9.1</td>
<td>17</td>
</tr>
<tr>
<td>Ethylene</td>
<td>12</td>
<td>4.1</td>
<td>23</td>
</tr>
<tr>
<td>Ethylene</td>
<td>5.0</td>
<td>2.5</td>
<td>9.0</td>
</tr>
<tr>
<td>Light hydrocarbons</td>
<td>8.7</td>
<td>3.0</td>
<td>20</td>
</tr>
<tr>
<td>Organic solvent</td>
<td>14</td>
<td>9.2</td>
<td>14</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>9.0</td>
<td>2.4</td>
<td>16</td>
</tr>
</tbody>
</table>
8. Component leak hole size

The determination of the most likely leak sizes resulting from pipe, seal or gasket failure proved a very difficult problem to resolve. For the purposes of this study absolute leak size values for a component type were impossible to establish because the leak size may depend on any number of unknown factors. Instead, a similar exercise to that carried out for component leak frequency was carried out for leak size. A literature survey was conducted for component leak size estimates. These were then used to identify an order of magnitude within which a typical leak size may lie for a given item of equipment. It was also possible to compare leak sizes for a given component against other component leak sizes to assess relative severity. Wherever possible leak size distribution data were also included from which it was hoped to identify the frequency of particularly large leaks within the overall leak frequency (some of the leak size distribution data have already been referred to in Section 7).

The leak size information given below comprises estimated hole sizes and emission rates. In the literature, gas leak sizes were usually expressed as hole sizes whilst liquid leaks were expressed as emission rates (typically drops per second). Each may be used in the calculation models in Section 9 to estimate a hazard distance based on the dispersion of the leak source emission.

8.1 Pipework

Pape and Nussey [22]

For 25 mm chlorine pipe:
Ratio of guillotine fractures: pipe splits = 1:10

No information was given concerning the expected size of the pipe split.
For general pipework consisting of pipes, flanges and valves the results of two studies by the authors are given in Table 8.1 for pipe and component ruptures and leaks. These figures supplement the pipe rupture data given in Section 7.2.1. As with the Pape and Nussey data these are useful mainly in estimating the fraction of leaks that are full-bore ruptures.

Rijnmond Report [28]

Two estimates of leak size distributions expected for pipework failure are given in Section 7.2.1. The Gulf distribution is more readily usable to estimate emission sizes as it gives relative leak areas. However, Figure 8.1 shows the original figure given by Hawksley [29] from which the Gulf leak size distribution was evidently derived by the respondent to the Rijnmond Report. This shows that this distribution is very approximate and so it was considered misleading to immediately accept these results.

8.2 Flanges/Gaskets

When discussing leakages from gaskets a distinction should be drawn between gasket types. In older plants the use of compressed asbestos fibre (CAF) gaskets is widespread. These gaskets are unreinforced and it has been customary to assume a worst-case failure scenario whereby a section of gasket between two bolts is blown out. The size of the resulting hole depends upon the flange bolting arrangement, which in turn depends on flange size and pressure duty. Tables 8.2 and 8.3 list flange dimensions according to the standard ANSI B16.5 [42] from which the leak slot area can be calculated. (The flange dimensions for larger pipe sizes may be found in the standard MSS-SP44 [43].)

For example, for an 8 inch (i.e. 200 mm) nominal bore ANSI class 300 flange the CAF gasket leak hole size may be estimated as follows:

\[
\text{Bolt centre circumference} = \pi d_p
\]
\[
= 1037 \text{ mm}
\]

where \(d_p\) is the flange diametric distance i.e. the distance between opposite bolt centres.
Bolt thickness = 22 mm
Number of bolts = 12
therefore leak section length = \((1037/12) - 22\) = 64 mm
Assuming a 1 mm thick gasket, leak area = 64 mm²

CAF gaskets have however been largely superseded for joints operating at high temperatures and pressures by spirally wound joint (SWJ) gaskets. These consist of a strip of gasket sealing material backed by metal reinforcement which is then spirally wound into a ring. Another gasket which is in common use for joints experiencing extreme wound into a ring. Another gasket which is in common use for joints experiencing extreme temperatures is the solid metal ring-type joint gasket (RTJ). These reinforced and solid gaskets cannot blow out in the manner of a CAF gasket and usually only leak if they are improperly assembled or if the flange faces themselves are pitted or scored. The leak sizes are therefore much smaller than for CAF gaskets and are commonly estimated to be of the order of 1 mm². Gasket leak sizes assumed in industrial hazard assessment work are given below.

British Gas Engineering Standard PS/SHA 1 [34]

Assumed leak area for all types of gasket handling natural gas = 0.25 mm²

The figure was adopted for all gasket types because it was thought that this represents the maximum leak size which may go undetected for any period of time. It does not imply that more severe leaks may never occur.

ICI/RoSPA Electrical Code (Vol. 1.5) [44]

Suggested leak hole dimensions for a CAF gasket section blow-out = 1.6 mm × 25 mm = 40 mm²
SWJ leak hole size dimensions = 0.05 mm × 50 mm = 2.5 mm²

As a general rule the guide suggested that SWJ gasket leak sizes are about 0.1 × the leak size of an equivalent CAF gasket. As previously mentioned in Section 2.2, the ICI/RoSPA guide is
currently under review. A recent (i.e. 1987) draft of the new guide [45] recommends keeping these hole sizes the same. The draft also suggests that a leak size of 0.5 mm² be assumed for a failure of a metal-to-metal joint without a resilient gasket. The resulting leak size distribution between joint types was thus taken to be 40:2.5:0.5, or approximately 10:1:0.1 for CAF:SWJ:RTJ gaskets.

Halogenated hydrocarbon plant [19]

The following estimates of leak size were used in the hazardous area classification procedure for the plant first mentioned in Section 5.2.1:

Small leak in liquid ethylene chloride flange gasket  = 5 drops/second  
= 0.22 \times 10^3 \text{ kg.s}^{-1} \text{ approx.} 

Large leak hole diameter  = 2.5 mm  
Large leak hole area  = 4.9 mm²  

Ratio of the frequency of large to small leaks (i.e. the severity fraction)  = 0.05  

For equipment handling natural gas, the following leak hole sizes were assumed:

Small hole size  = 7 mm²  
Large hole size  = 38 mm²  
Severity fraction  = 0.02  

The natural gas hole sizes were also applied to leaks from similar equipment handling hydrogen.

8.3 Valve glands

It was not possible to differentiate between valve type when extracting information from the open literature concerning valve leakages. The following estimates were therefore taken as relating to an aggregate of types.
Valve leak sizes were estimated using the same criterion as gasket leaks i.e. the largest leak size that would go undetected for any length of time was adopted.

Valve gland stem seal
(packed gland) leak area = 0.25 mm²

ICI draft procedure [45]

Typical valve leaks were not quantified by the ICI/RoSPA code. However the draft new procedure suggests a maximum valve gland liquid leak rate of 2 drops per second, or approximately $2 \times 10^{-4}$ kg.s⁻¹.

Halogenated hydrocarbon plant [19]

Valve gland leaks were taken to be of comparable size to flange leaks:

Ethylene chloride valve gland leak rate
= 5 drops/second
= $0.22 \times 10^{-3}$ kg.s⁻¹

Large hole size
= 5.1 mm²

Severity fraction
= 0.05

8.4 Pump seals

When discussing leaks from centrifugal pump seals it is necessary to differentiate between packed gland seals and mechanical seals.

The characteristics of pump seals have been described fully by Ramsden [46] and Hoyle [47]. Packed gland seals control rather than prevent leakage and are therefore seldom used on centrifugal pumps handling flammable liquids, although they may be found on reciprocating pumps. Mechanical seals attempt to prevent visible leakages completely, although there is always a very small emission resulting from the need to lubricate the seal faces. Because of this...
use of mechanical seals is widespread on systems handling flammable liquids and consequently most of the emission data below relate to this seal type. However, mechanical seals may fail completely, thus resulting in serious leakage. Throttle bushes are therefore sometimes fitted to limit the leak rate. The seal area may also be enclosed and purged or drained; in these circumstances it is then the mouth of the drain that is the emission source. The use of double or tandem mechanical seals is also common on pumps handling flammable liquids, which again limits the leak size.

ICI draft procedure [45]

Draft hole size estimates made for a variety of seal arrangements are listed below. They were derived from expert judgement on the relative leak sizes for each seal arrangement.

Single mechanical seal with
no auxiliary sealing device
- Continuous release in operation
- Continuous release on standby
- Failure leak release
(described as "discontinuous dribble")

Single mechanical seal plus
auxiliary sealing device (e.g. lip seal or soft packing) and local drain
Drain - continuous release
(operation/standby)
- failure leak release
Bush - continuous release
- failure leak release
Seal - failure leak release
(approx. 2 drops per second)

\[
\begin{align*}
\text{Continuous release in operation} & : 0.02 \times 10^3 \text{ kg.s}^{-1} \\
\text{Continuous release on standby} & : 0.1 \times 10^3 \text{ kg.s}^{-1} \\
\text{Failure leak release} & : 1.0 \times 10^3 \text{ kg.s}^{-1} \\
\text{Drain - continuous release} & : 0.02/0.1 \times 10^3 \text{ kg.s}^{-1} \\
\text{Bush - continuous release} & : 0.02 \times 10^3 \text{ kg.s}^{-1} \\
\text{Seal - failure leak release} & : 0.2 \times 10^3 \text{ kg.s}^{-1}
\end{align*}
\]
As above, but with innocuous flushing liquid between seal and auxiliary device in which leakage is soluble

Seal - failure leak release
(leakage into flushing liquid)
Drain - failure leak release
Bush - failure leak release
(both as a result of flush loss)

Tandem mechanical seals where the space between seals is at atmospheric pressure during normal operation and leakage is readily apparent

Vent - failure leak release

Packed gland
- Continuous release
- Failure leak release

Summers-Smith [48]

Estimated constant leak rate for a single mechanical seal on a centrifugal pump

\[ = 0.003 - 0.03 \times 10^{-3} \text{ kg.s}^{-1} \]

8.5 Compressor seals

ICI draft procedure [45]

Estimates of the leak sizes occurring as a result of various compressor seal failures are presented below.
Reciprocating compressors

Closed vented valve chamber
- Continuous release 1.0 mm²
- Failure leak release 5.0 mm²

Closed vented valve chamber with liquid injected gland
- Failure leak release 5.0 mm²

Purged gland
- Failure leak release 1.0 mm²

Hand unloader valves, spindles sealed by lip seals and O-rings
- Failure leak release 1.0 mm²

Centrifugal compressors

The sealing arrangements for centrifugal compressors usually employ either a mechanical face-contact seal or a floating close-contact bush. High pressure oil is injected into the seal area as a lubricant and purge which is then drained. The equivalent leak orifice size for a compressor seal is estimated using the formula $x \pi d_s^2$ where $d_s$ is the shaft diameter and $x$ is a factor based on:

1. A "base-case" shaft diameter $d_F$
2. The clearance between the seal and the shaft $l$
3. The orifice discharge coefficient $C_D$

For example, for a purged labyrinth seal the leak hole size was based on a 150 mm diameter shaft ($d_F$) with a seal clearance ($l$) of 1 mm. Therefore;
Equivalent orifice (m²)  

\[ = C_D \pi d_s l \]

\[ = 0.001 C_D \pi d_s \]

\[ = 0.001 C_D \pi d_s \cdot \left( \frac{d_s}{d_F} \right) \]

(for the base case \( d_s = d_F \), for all other shaft diameters \( d_s/d_F \) represents a scaling factor)

\[ = \frac{0.001}{0.15} C_D \pi d_s^2 \]

\[ = 0.0067 C_D \pi d_s^2 \]

for \( C_D = 0.6 \), equivalent orifice area

\[ = 0.004 \pi d_s^2 \]

For a range of compressor seals the following failure leak sizes were suggested:

Purged labyrinth with purge flow into compressor
- Seal area \( 0.004 \pi d_s^2 \)

Purged labyrinth with purge flow out of compressor
- Vent \( 0.004 \pi d_s^2 \)
- Gland (assuming sealant loss) \( 0.004 \pi d_s^2 \)

Floating ring seal with purged seal liquid reservoir
- Seal trap drain (for 3 mm diameter drain) \( 7 \text{ mm}^2 \)
- Seal reservoir vent \( 0.0006 \pi d_s^2 \)
- Seal area \( 0.0006 \pi d_s^2 \)

This is based on \( d_F = 0.1 \text{ m} \) and \( l = 0.1 \text{ mm} \).
Loss of sealant is assumed.
Halogenated hydrocarbon plant

In a HAC study of another plant on the same site as the plant first considered in Section 5.2.1 [49], an estimate was made of the continuous leakage from a hydrogen booster gland which surmised that the overall leak rate corresponded to an orifice size of 1 mm².

Summers-Smith [48]

Estimated constant leak rate for a reciprocating compressor with segmented packing and vented to atmosphere
- Lubricated
  0.003 - 0.03 x 10^{-3} \text{ kg.s}^{-1}
- Unlubricated
  0.03 - 0.3 x 10^{-3} \text{ kg.s}^{-1}

8.6 Small-bore connections

A limited amount of data were located concerning the size of non-rupture emission sizes which may be expected from small-bore pipe connections.

British Gas Engineering Standard PS/SHA1 [34]

Compression fitting leak size
= 0.25 \text{ mm}^{2}

Screwed connection leak size
= 0.25 \text{ mm}^{2}

ICI draft procedure [45]

For the small-bore pipework serving compressors the leak size was estimated as 0.1 mm².

8.7 Fugitive emissions

Fugitive emissions are routine emissions from plant components such as mechanical seals on pumps which emit very small amounts of process fluid as normal operating practice. Several studies have been made with a view to quantify the total fugitive emissions from petrochemical
installations. Although these invariably concentrated on the pollution rather than the fire hazard aspect, the studies provide some potentially useful leak data. The results from these studies have been summarised elsewhere [50]. The studies usually involved taking a random sample of potential emission sources and enclosing them in bags which collected the expected small emissions. An average emission rate, or factor, was then calculated for the entire sample. However, because the produced emission factors incorporated non-leaking sources they were very low. For example, Bierl [51] quoted a mean emission factor of 0.014 grams per hour per valve for a sample of 125 ball valves which clearly does not represent a credible hazard. Some studies however went on to give details of leak size distribution within the sample. Table 8.4 presents component leak size distributions which resulted from a study of thirteen petroleum refineries performed by the Radian Corporation for the US Environmental Protection Agency [52]. Table 8.4 shows that for sources handling gas and flashing liquids the leak size distributions were heavily skewed i.e. typically 5 percent of the total sources sampled were responsible for 95 percent of the total emission rate. The Radian data are still only of limited use since the upper leak range is only 1 lb/hour which is equivalent to \(0.126 \times 10^3\) kg.s\(^{-1}\). However, the data were used by Powell [53] to derive the following estimates of "worst-case" leak rates for plant operator exposure:

Leak rate of worst 5% pumps  
in light (i.e. flashing) liquid service \(0.341 \times 10^3\) kg.s\(^{-1}\)

Leak rate of worst 5% valves  
in light liquid service \(0.038 \times 10^3\) kg.s\(^{-1}\)

8.8 Application of surveyed data

The hole size data presented in this section were used in two ways for the purposes of this study. Firstly, typical hole sizes were derived from the available data which were used to illustrate the use of the mathematical models listed in Section 9 in calculating source emission sizes and thus hazard distances around items of process equipment. The hazard distance calculation results are given in Section 10. Secondly, crude values of hole size were chosen which corresponded with the selected leak frequencies listed in Section 7.3. The leak frequencies and sizes were thus used as input in the fire and explosion study described in Section 12. The rationale for specifying values for these purposes for each equipment item is given below.
Pipework

The example hazard distance calculations in Section 10 do not include the consequences of pipe fracture. The calculation given is that of an opened drain or sample point which represents the full-bore emission through a 15 mm nominal bore pipe.

The pipe leak size data used for the fire and explosion study were derived from the estimates given in the Rijnmond Report (see Figure 8.1). The basic hole size here was taken to be a full-bore release such that the available leak area was the cross-sectional area \( A \) of the pipe. A major leak was thus taken to be \( 0.1 \times A \) whilst a minor leak was \( 0.01 \times A \) (for further explanation of the terminology adopted for this part of the study see Section 7.3). This then resulted in an overall leak size distribution of \( 10:1:0.1 \) for rupture:major:minor leak sizes. In the paragraphs below this leak size distribution is assumed to apply to most leak sources. In some cases there were not enough data available to wholly justify this rather sweeping assumption. However, the preliminary estimates of leak size and frequency were made here to enable the establishment of a crude methodology. Whilst these values are readjusted in the manner described in Section 12, more sophisticated refinement can only be achieved with considerable effort in future data gathering, which is beyond the scope of this study.

Flanges/Gaskets

The following hole sizes were adopted as example values to illustrate the calculation of hazard distances around gasketted joints:

<table>
<thead>
<tr>
<th>Leak Type</th>
<th>Area</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>25 mm²</td>
</tr>
<tr>
<td>Minor leak</td>
<td>2.5 mm²</td>
</tr>
<tr>
<td>Small leak</td>
<td>0.25 mm²</td>
</tr>
</tbody>
</table>

The basic case for the above leak sizes was taken to be the major leak case i.e. a large portion of a CAF gasket segment being blown out from between two adjacent bolt holes. The relative leak sizes for the minor and small leaks were derived from the ICI values given for SWJ and RTJ leak sizes.
For the fire and explosion study, the number of leak size categories was reduced to two, as it was felt that very small gasket leaks (i.e. 0.25 mm²) did not present a significant flammable risk on the type of plant considered in Section 12. The major leak case was therefore taken to be the same as above, in other words the size of the hole resulting from a segment of CAF gasket being blown out. A minor leak was taken to be one-tenth this area. In the basic simplified analysis given in Section 12 the difference in leak sizes between CAF and reinforced gaskets was not considered beyond this brief categorisation.

Valves

The hazard distances around valves were not used as examples in the calculation results given in Section 10. From the data presented here it appeared that valve leaks may be expected to be of a similar magnitude to gasket leaks. However, the leak frequency data presented in Section 7.2.2 indicated that more serious leaks may cause significant levels of flammable risk. In particular, the data item originating from WASH 1400 [25] and quoted by the Rijnmond Report [28] was concerned with the frequency of complete valve rupture. It was therefore decided to adopt the leak size distribution used for rupture: major: minor pipework failures above to valve leaks also i.e. $A:0.1A:0.01A$, where $A$ is the cross-sectional area of the pipe. The increased severity of these leaks compared to the data presented here is also reflected by the relatively low leak frequencies adopted in Section 7.3 in comparison with most of the data presented in Section 7.2.2.

Pumps

For the example hazard distance calculations in Section 10 a range of centrifugal pump leak sizes were chosen, based on the leak sizes given in Section 8.4. The values were also chosen to enable the resulting hazard distances to be compared with the Zone 2 distances given by existing codes of practice in Figure 2.1. For simplicity the same hole sizes as were used for gaskets were also used for pumps. This chosen range was also intended to take in the range of pump sizes since it was suspected that the magnitude of pump seal leaks, like compressor seal leaks (see below), is proportional to pump shaft size.
For the fire and explosion study it was again considered that most of the pump leak sizes given in Section 8.4 (i.e. a few grams per second of fluid or vapour) would present a very small flammable hazard on the typical "standard" plant formulated in Section 12. It was decided therefore to adopt the hole size distribution used for pipe and valve leaks for pump leaks also, and to adjust the pump leak frequencies accordingly as described in Section 7.3.

Centrifugal compressors

The example hole sizes selected for use with the calculational models in Section 10 were derived using the formulae given in Section 8.5 which described the dependency of leak hole size on shaft diameter. Accordingly two hole sizes were chosen as being typical of seal leaks on small and large compressors: 25 and 250 mm².

Compressors were not included in the fire and explosion analysis and so a leak size distribution was not formulated.

Small-bore connections

Leakage from small-bore connections was not considered in the hazard range examples given in Section 10 beyond the aforementioned sample/drain point leakage.

Argument was made in Section 7.2.7 that the major hazard from small-bore connections is presented by fractures. Consequently two leak levels were used in the fire and explosion study: complete full-bore rupture and partial fracture, here taken to be one-tenth the area of a full-bore rupture.

In conclusion, a literature survey was carried out for leak source data concerning typical plant items. The available data seemed to relate to mostly small leaks that corresponded to the high leak frequencies for these items given in Section 7. Example hole sizes were selected for two purposes: firstly to enable the calculation of example hazard distances; and secondly as input to the fire and explosion analysis. These are described in Sections 10 and 12 respectively. It was realised that the selected hole sizes for the fire and explosion study were judgemental and that the acquisition of further data was necessary to wholly justify their adoption.
Table 8.1 Pipework leak severity distribution (Blything and Parry [30])

<table>
<thead>
<tr>
<th>Pipe diameter (mm)</th>
<th>Leaks</th>
<th>Ruptures</th>
<th>Ruptures/Leaks</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Study 1 (CHEM 1):</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$d &lt; 150$</td>
<td>29</td>
<td>2</td>
<td>0.07</td>
</tr>
<tr>
<td>$150 &lt; d &lt; 380$</td>
<td>12</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>Overall</td>
<td>41</td>
<td>2</td>
<td>0.05</td>
</tr>
<tr>
<td><strong>Study 2 (STEAM 1):</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$d &lt; 150$</td>
<td>115</td>
<td>9</td>
<td>0.08</td>
</tr>
<tr>
<td>$150 &lt; d &lt; 380$</td>
<td>26</td>
<td>7</td>
<td>0.27</td>
</tr>
<tr>
<td>$d &gt; 380$</td>
<td>6</td>
<td>1</td>
<td>0.17</td>
</tr>
<tr>
<td>Unspecified</td>
<td>18</td>
<td>2</td>
<td>0.11</td>
</tr>
<tr>
<td>Overall</td>
<td>165</td>
<td>19</td>
<td>0.12</td>
</tr>
</tbody>
</table>
Table 8.2 Flange bolting arrangement data - bolt and bolt hole dimensions [42]

<table>
<thead>
<tr>
<th>ANSI class bore (in)</th>
<th>Flange nominal bore (in)</th>
<th>Number of holes</th>
<th>Hole diameter (in)</th>
<th>Bolt diameter (in)</th>
</tr>
</thead>
<tbody>
<tr>
<td>150</td>
<td>0.5 - 1.5</td>
<td>4</td>
<td>0.62</td>
<td>0.5</td>
</tr>
<tr>
<td></td>
<td>2 - 3</td>
<td>4</td>
<td>0.75</td>
<td>0.625</td>
</tr>
<tr>
<td></td>
<td>3.5 - 4</td>
<td>8</td>
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<td>1.12</td>
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<td>1.125</td>
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<td>24</td>
<td>20</td>
<td>1.38</td>
<td>1.25</td>
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|                      | 300                      | 0.5              | 0.62              | 0.5               |
|                      | 0.75 - 1.25              | 4                | 0.75              | 0.625             |
|                      | 1.5                      | 4                | 0.88              | 0.75              |
|                      | 2                        | 8                | 0.75              | 0.625             |
|                      | 2.5 - 5                  | 8                | 0.88              | 0.75              |
|                      | 6                        | 12               | 0.88              | 0.75              |
|                      | 8                        | 12               | 1.0               | 0.875             |
|                      | 10                       | 16               | 1.12              | 1.0               |
|                      | 12                       | 16               | 1.25              | 1.125             |
|                      | 14                       | 20               | 1.25              | 1.125             |
|                      | 16                       | 20               | 1.38              | 1.25              |
|                      | 18 - 20                  | 24               | 1.38              | 1.25              |
|                      | 24                       | 24               | 1.62              | 1.5               |

91
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<th>Flange nominal bore (in)</th>
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<th>ANSI class 300 diametric distance (in)</th>
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Table 8.4 Radian Corporation fugitive emission size distributions [52]

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<tr>
<th>Valves</th>
<th>Gas/Vapour</th>
<th>Light liquid/Two phase</th>
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<tbody>
<tr>
<td>Leak range (lb/hr)</td>
<td>A</td>
<td>B</td>
</tr>
<tr>
<td>&gt;1.0</td>
<td>1.2</td>
<td>70.0</td>
</tr>
<tr>
<td>0.1 - 1.0</td>
<td>3.2</td>
<td>23.3</td>
</tr>
<tr>
<td>0.01 - 0.1</td>
<td>7.6</td>
<td>5.8</td>
</tr>
<tr>
<td>0.001 - 0.01</td>
<td>8.7</td>
<td>0.8</td>
</tr>
<tr>
<td>0.00001 - 0.001</td>
<td>6.6</td>
<td>0.1</td>
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<tr>
<th>Pump seals</th>
<th>Light liquid</th>
<th>Heavy liquid</th>
</tr>
</thead>
<tbody>
<tr>
<td>Leak range (lb/hr)</td>
<td>A</td>
<td>B</td>
</tr>
<tr>
<td>&gt;1.0</td>
<td>4.0</td>
<td>70.6</td>
</tr>
<tr>
<td>0.1 - 1.0</td>
<td>15.5</td>
<td>24.6</td>
</tr>
<tr>
<td>0.01 - 0.1</td>
<td>22.7</td>
<td>4.4</td>
</tr>
<tr>
<td>0.001 - 0.01</td>
<td>16.4</td>
<td>0.4</td>
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<table>
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<tr>
<th>Compressor seals</th>
<th>Hydrocarbon</th>
<th>Hydrogen</th>
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</thead>
<tbody>
<tr>
<td>Leak range (lb/hr)</td>
<td>A</td>
<td>B</td>
</tr>
<tr>
<td>&gt;1.0</td>
<td>16.2</td>
<td>74.3</td>
</tr>
<tr>
<td>0.1 - 1.0</td>
<td>33.8</td>
<td>24.3</td>
</tr>
<tr>
<td>0.01 - 0.1</td>
<td>16.9</td>
<td>1.4</td>
</tr>
<tr>
<td>0.001 - 0.01</td>
<td>4.9</td>
<td>0.0</td>
</tr>
<tr>
<td>0.00001 - 0.001</td>
<td>2.1</td>
<td>0.0</td>
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continued
Table 8.4 continued

<table>
<thead>
<tr>
<th>Leak range (lb/hr)</th>
<th>A</th>
<th>B</th>
<th>A = Percent of total sources screened with sampled leak rates within leak range.</th>
<th>B = Percent of total mass emissions attributable to sources within leak range.</th>
</tr>
</thead>
<tbody>
<tr>
<td>&gt;1.0</td>
<td>0.0</td>
<td>0.0</td>
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<td>0.1 - 1.0</td>
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<td>0.01 - 0.1</td>
<td>0.6</td>
<td>30.1</td>
<td></td>
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<tr>
<td>0.001 - 0.01</td>
<td>1.3</td>
<td>6.0</td>
<td></td>
<td></td>
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<tr>
<td>0.00001 - 0.001</td>
<td>0.9</td>
<td>0.7</td>
<td></td>
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</table>
Figure 8.1 Summary of pipe leak size distributions given by Hawksley [29]
9. Calculation models for emission and dispersion

In order to quantify the extent of the hazardous area around the sources listed in Table 5.1 it is necessary to model the emission and the dispersion of the release. This section describes a literature survey for calculational models which describe these processes. Where several models were found which described the same process the model preferred for application in Section 10 was usually the one commonly used in HAC work. The selected models listed below tend to be simple in application and frequently empirical in nature but will provide at least a first order estimate of the physical phenomena involved. Section 10 describes the application of these models to a variety of typical emission scenarios to show how hazard distances vary according to the emission and dispersion mode and according to the nature of the released flammable substance. The models chosen for use in Section 10 were not necessarily the most rigorous models available. However, the procedure for calculating hazard distances reflects the overall project strategy described in Section 3 as it was designed to provide a framework which allows for models to be substituted without affecting the overall strategy. Thus in cases where the user expresses a preference for a more detailed or mechanistic model to the one chosen here, it may be incorporated.

The main objective of modelling a flammable release is to establish an area around the potential release source within which the probability of ignition may be estimated. This may be expressed either as an absolute value or a value relative to other ignition sources. The estimation of the flammable risk posed by a collection of different emission sources incorporates the results of the studies of leak source inventories, leak frequencies and hole size estimates and is discussed further in Section 12.

The formation of a flammable atmosphere from an initial emission may require several physical processes to be considered for modelling. For example, if a pump seal starts to leak, the processes to be modelled are:
Liquid emission
Pool spread (for a severe leak)
Vaporisation of the liquid pool
Dispersion of the flammable vapour in the ambient atmosphere

Thus each release scenario may require its own model system to be adequately described. As expressed above, each model system may be expected to account for the nature and conditions of the process fluid and the effects of changes in these conditions. The emission and dispersion of the releases from the sources listed in Table 5.1 fall into one or more of the following categories which may be separately modelled:

**Emission**

- Gas/vapour flow
- Liquid flow
- Flashing liquid flow (i.e. two-phase flow)
- Vaporisation of liquids

**Dispersion**

- Gas/vapour jets
- Liquid jets
- Gas dispersion

Several model systems have been published which either completely or partly describe the behaviour of flammable releases. These may be found in the following:

- Lees [10]
- Lihou (ed.) [54]
- Cude [55]
- Mecklenburgh [56]
- BS 5345 Part 2 draft appendix [57]
A list of the models included in each of these references is presented in Table 9.1. These texts are frequently referred to throughout this section since they provide various models currently in use for hazard assessment. The relevance of other models are also discussed where appropriate. Some of these model systems (e.g. The Rijnmond Report) were devised to estimate the effects of very large releases, therefore some of the recommendations made as to which model to use may not be applicable to the smaller releases that HAC is designed to protect against. These cases will be highlighted where applicable.

9.1 Emission

9.1.1 Gas/vapour emission

The emission rate of gas through an orifice such as a hole is calculated according to whether the flow is subsonic or sonic. For subsonic flow the mass flow rate $G$ (kg.s$^{-1}$) is given by:

$$G = C_D A \rho \left[ \frac{2\gamma}{\gamma-1} \frac{RT_1}{M} \left( 1 - \left( \frac{P_a}{P_1} \right)^{\frac{\gamma-1}{\gamma}} \right) \right]^{0.5} \tag{9.1}$$

where:

$G$ = mass flow (kg.s$^{-1}$)  
$C_D$ = discharge coefficient  
$\rho$ = gas density (kg.m$^{-3}$)  
$A$ = orifice area (m$^2$)  
$\gamma$ = gas adiabatic coefficient  
$R$ = gas constant (8314 J.kmol$^{-1}$.K$^{-1}$)  
$T_1$ = containment temperature (K)  
$M$ = molar weight (kg.kmol$^{-1}$)  
$P_a$ = atmospheric pressure (Pa)  
$P_1$ = containment pressure (Pa)
The criterion for sonic or choked flow is that the containment pressure $P_1$ reaches a critical value $P_c$, where:

$$\frac{P_c}{P_1} = \frac{2}{\gamma+1} \quad (9.2)$$

Above $P_c$ the flow is no longer dependent on the ambient pressure ($P_1$ is approximately 2 bar abs. for choked flow). For sonic conditions:

$$G = C_D A P_1 \left[ \frac{\gamma M}{z RT_1} \left( \frac{2}{\gamma+1} \right)^{\gamma-1} \right]^{0.5} \quad (9.3)$$

where $z$ is the gas compressibility. For most gases at modest pressures the ideal gas law applies and the compressibility may be assumed to be 1. However, at elevated pressures the gas may assume a non-ideal state so that the compressibility is less than 1. The change in gas compressibility $z$ can be accounted for by using the Redlich-Kwong equation of state as expressed by Smith and Van Ness [60]:

$$z^3 - z^2 + (A - B^2 - B)z - AB = 0 \quad (9.4)$$

where

$$A = \frac{a P_1}{R^2 T_i^{2.5}}$$

$$B = \frac{b P_1}{R T_i}$$

$$a = \frac{0.42748 R^2 T_i^{2.5}}{P_c}$$

$$b = \frac{0.08664 R T_i}{P_c}$$

and where $T_i$ and $P_c$ are the critical temperature and pressure of the gas. The non-ideality of the gas at high pressures also affects the value of $\gamma$. However, since Equation (9.3) is not very sensitive to small changes in $\gamma$, this was ignored when investigating the use of the model for setting zone distances in Section 10. Equation (9.3) is used in all the references listed in Section 9.1 to calculate gas flows since sonic flows are usually encountered in leak calculations. The derivations of all the gas flow equations given here are found in Lees [10].
9.1.2 Liquid emission

The equation commonly used for the emission rates of liquids below the flash point is the orifice liquid flow equation derived from the Bernoulli equation:

\[ G = \frac{C_d A \sqrt{2 \rho_L (P_1 - P_2)}}{\sqrt{R}} \]  

(9.5)

where \( \rho_L \) is the liquid density (kg.m\(^{-3}\)) and \( P_1 \) and \( P_2 \) are the pressures upstream and downstream of the orifice (Pa).

9.1.2.1 Liquid jet atomisation

Upon emission the liquid jet may either form a continuous jet or atomise into a spray. Since this obviously affects the subsequent dispersion of the release it is necessary to test the emission for the formation of spray. Wheatley [61] gave a simple predictive method for a liquid jet that essentially forecasts atomisation when the jet reaches a critical velocity \( u_{\text{crit}} \). At this velocity the drag forces caused by the air overcome the surface tension of the jet which then breaks up. This happens when

\[ \text{We}_j \geq \text{Re}_j^{-0.45} \times 10^6 \]  

(9.6)

where \( \text{Re}_j \) is the jet Reynolds number and \( \text{We}_j \) is the jet Weber number given by the following:

\[ \text{Re}_j = \frac{u d \rho_L}{\mu_L} \]  

(9.7)

\[ \text{We}_j = \frac{u^2 d \rho_L}{\sigma_L} \]  

(9.8)

\( u \) is the initial liquid jet velocity, \( d \) is the initial jet diameter (m), \( \mu_L \) is the liquid viscosity (kg.m\(^{-1}\).s\(^{-1}\)) and \( \sigma_L \) is the coefficient of surface tension (N.m\(^{-1}\)). Wheatley further suggested that an experimental correlation by Brodkey [62] may be used to estimate the maximum atomised drop diameter \( d_p \). Brodkey related the maximum drop diameter to the drop Weber number \( \text{We}_D \) as follows:

\[ \text{We}_D = \frac{u^2 d_p \rho_a}{\sigma_L} = 20 \]  

(9.9)

where \( \rho_a \) is the air density (kg.m\(^{-3}\)).
9.1.2.2 Rain-out from atomised liquid jets

If the liquid jet atomises then it is necessary to determine how much of the spray descends to the ground or "rains out". If a significant amount of the jet rains out this affects subsequent dispersion calculations. The maximum drop size \( d_o \) may be used to estimate the maximum velocity at which drops descend from the jet. Wheatley called this the gravitational settling velocity of the drop \( u_o \) which is derived using a force balance of the drag of the drop and gravity, giving,

\[
\eta Re_o^2 C_D(Re_o) = \frac{4}{3} d_o^2 \left( \frac{\rho_L - \rho_a}{\rho_a} \right) g \mu_a^2
\]  

(9.10)

\( Re_o \) is the drop Reynolds number:

\[
Re_o = \frac{u_o d_o \rho_a}{\mu_a}
\]  

(9.11)

where \( \mu_a \) is the viscosity of air (kg m\(^{-1}\) s\(^{-1}\)). \( \eta \) is a correction factor applied to the rigid sphere discharge coefficient \( C_D \) for the motion of the fluid within the drop:

\[
\eta = \frac{1 + \mu_u/\mu_L}{1 + 2\mu_u/3\mu_L}
\]  

(9.11)

and the drag coefficient is expressed as

\[
C_D = \frac{24}{C_D} + C'_D
\]  

(9.13)

\( C'_D \) is a correction factor for inertia and turbulence effects not included in Stoke's law. Wheatley concluded that drops must have a diameter greater than about 0.3 mm before the terminal settling velocity exceeds 1 m s\(^{-1}\) and that the drop will stay in the jet if its terminal velocity is less than or equal to one-tenth of the jet velocity.

9.1.3 Determination of discharge coefficient

The mass flow equations given above require an estimate of the discharge coefficient \( C_D \) to be made. Typical values are as follows:

- Sharp-lipped orifice \( C_D \) 0.6
- Short nozzle stub (Length/Diameter < 3) \( C_D \) 0.75 - 0.8
The value of 0.6 for the orifice $C_D$ has been given by Lees and suggested by the ICI draft procedure first referred to in Section 2.2 [45] as a "general purpose" $C_D$ for use with compressor leaks. The nozzle $C_D$ was also given by Lees.

For pipe breaks where the pipe is longer than a stub it becomes necessary to incorporate the friction effect of the pipe into the $C_D$ value for a better estimate of emission rate. This is achieved in different ways depending on the phase of the emitted fluid. For liquids the friction losses are expressed in terms of a number of velocity heads. These account for friction losses at the entrance and exit of the pipe, in the fittings used and in the pipe itself. The velocity head losses of the pipe entrance, exit and associated fittings may be straightforwardly obtained from lists such as the one given in Perry [63]. The friction loss of the pipe is expressed as a function of pipe length, diameter and the Fanning friction factor $f$ thus:

$$K = \frac{4fL}{D} \quad (9.14)$$

where $K$ is the velocity head loss, $f$ is a function of viscosity and turbulence for which a typical value is 0.005 for fully turbulent flow in a 25 mm pipe. The method for determining the friction loss in pipes was given by Perry and was used for the modelling of hazardous releases by Ramskill and by the Rijnmond Report. The losses of the pipe and the fittings are summed to arrive at an overall discharge coefficient thus:

$$C_D = \frac{1}{\sqrt{1 + \Sigma K}} \quad (9.15)$$

The discharge coefficient for pipework handling gas may be obtained in a similar manner by summing the velocity head losses for the pipe and fittings. However, whilst the fittings losses are the same as those used for liquid flows the pipe losses are obtained differently since the friction causes pressure and temperature changes. Ramskill suggested the use of a procedure given by Perry which employs Lapple charts to provide a first-guess estimate of pipe friction. It is suggested that this method is rigorous enough for HAC as Carter [64] has stated that a more refined solution may only be obtained with the use of complex iteration routines to estimate outlet conditions which can only be solved using a computer.
9.1.4 Two-phase emissions

Wheatley [61] identified three types of superheated liquid flow through an outlet, which may be qualitatively described as follows:

1. The fluid remains a superheated liquid up to the outlet. This is the case if the leak path is very short so that the liquid has no time to flash. In these cases the mass flow is calculated using the liquid flow orifice Equation (9.5).

2. The fluid flow at the outlet is a two-phase mixture with the liquid and vapour phases at equilibrium at the choke pressure. Since the flow is choked the liquid is superheated. This occurs if the leak path is long and results in a lessened discharge rate compared to single-phase liquid flow.

3. Between the two cases above there is a period in which the flow is two-phase but not in equilibrium. The flow thus depends on the state of equilibrium reached.

All the references listed in Section 9.1 recommended the use of the model first published by Fauske [65] to calculate flashing releases of flammable material. The model was formulated primarily to describe releases of superheated water, and is given by

\[ G = C_g A \sqrt{2 \rho_M (P_1 - P_e)} \]  

(9.16)

where \( \rho_M \) is the average density of the flow at the outlet and \( P_e \) is the choke pressure. The value of \( P_e \) was assumed to be correlated by the length to diameter ratio \( (L/D) \) of the leak path as follows:

\( L/D = 0 \) \hspace{1cm} \text{In other words orifice flow.} \ P_e \text{ is ambient pressure and the flow is determined using Equation (9.5).}

\( L/D > 12 \) \hspace{1cm} \text{The flow is choked and the phases are at equilibrium. The flow is assumed to be homogenous i.e. the gas phase is dispersed evenly within the liquid phase as tiny bubbles so that the phases are travelling with the same velocity.} \ P_e \text{ is assumed to be the same as for a single phase gas system and is thus obtained}
from Equation (9.2). Since the phases are assumed to be at equilibrium the flash fraction may be obtained from the following expression:

$$(1-q)C_{PL}(-dT) = \lambda d\ q$$

which upon integration gives

$$q = 1 - \exp[-(T_i - T_e)C_{PL}/\lambda]$$

(9.17)

where $C_{PL}$ is the liquid specific heat (J.kg$^{-1}$.K$^{-1}$), $T_i$ is the liquid temperature, $T_e$ is the equilibrium temperature at the choke pressure $P_e$ and $\lambda$ is the latent heat of vaporisation (J.kg$^{-1}$). The flash fraction is used to calculate the mixture density $\rho_M$:

$$\rho_M = \left[ \left(\frac{q}{\rho_f} \right) + \left(\frac{1-q}{\rho_L} \right) \right]^{-1}$$

(9.18)

which enables $G$ to be calculated from Equation (9.16).

$0 < L/D < 12$ Between the two above extremes the flow is presumed to go through a metastable region at $L/D = 3$, where the flow is calculated directly using Equation (9.16) with $\rho_M = \rho_L$ and where $P_e$ is obtained from Equation (9.2). After the metastable region the flowrate is interpolated for $L/D$ values between 3 and 12.

The use of this model has been criticised for two main reasons. Firstly, the model was developed based on experiments using water. However, there is no justification presented in any of the references cited in Section 9.1 as to why it should be applicable to substances other than water. Second, and more importantly, the assumption that choke pressure and therefore flow rate is correlated by the pipe $L/D$ ratio has been shown to be unjustified. More recent experimental work by Fletcher and Johnson [66] using water and Freon-11 indicated that it is the pipe length alone which correlates the state of equilibrium reached by the flow. Figure 9.1 presents two flow curves for the fluids used by Fletcher and Johnson which may be used to estimate two-phase flowrates (the term excess pressure in the figure is a measure of the amount of superheat). The mass flux $G_F$ (shown as $G$ in Figure 9.1) may be obtained from the empirical relationship shown between pipe length $L$ (shown as $l$ in Figure 9.1) and the group

$$\frac{G_F^2}{C_P^2 P_1}$$

(9.18)

where $P_1$ is the stagnation, or containment pressure (Pa). Note that the curves for the different substances give similar results which suggests that they may be used for other flashing liquids.
In a more recent reference [67], Fauske agreed that pipe length is the appropriate criterion for estimating the stability of two-phase pipe flow, and gave the following equation for a first-order estimation of mass flow which he called the equilibrium rate model (ERM):

\[
G_p = \frac{dP}{dT} \sqrt{\frac{T}{NC_{pl}}} \tag{9.19}
\]

where

\[
N = \left( \frac{dP}{dT} \right)^{0.5} \frac{T}{2\Delta P \rho_c C_s^2 C_{pl}} + 10L \tag{9.20}
\]

and where \( T \) is the absolute temperature of the containment (K), \( dP/dT \) is the differential saturation pressure change at temperature \( T \) (Pa·K\(^{-1}\)) and \( \Delta P \) is the total available pressure drop (Pa). Fauske estimated that equilibrium conditions are reached after about 0.1 m and that above this value \( N = 1 \).

Both the Fletcher and Johnson curves and the Fauske ERM depend upon experimental observations to estimate mass flows and do not involve detailed considerations of parameters such as the choke pressure. Other simple models which have been derived from first principles rather than experimental data include those of Nyren and Winter [68] and Wheatley [61]. These were both formulated for modelling the equilibrium flow of flashing ammonia releases. The Nyren and Winter model uses the following system of equations:

\[
G = A \sqrt{\frac{P_0 - P_e}{V_c \left( 1 + \frac{L}{\frac{D^2}{4}} \right) - \frac{V_c}{C_{lp} \left( 1 - \frac{1}{2C_{lp}} \right)}}} \tag{9.21}
\]

where

\[
V_c = X_e V_{c,G,SAT} \frac{T_e}{T_{SAT}} + (1 - X_e)V_L
\]

\[
X_e = \frac{T_e}{\lambda_c} \left( S_{l,SAT}(P_0) - S_{l,SAT}(P_e) + C_{pl} \ln \frac{T_e T_{SAT}(P_e)}{T_e T_{SAT}(P_0)} \right)
\]

\[
T_e = T_0 - n_e (T_0 - T_{SAT}(P_e))
\]

\[
G = \text{mass flow (kg·s}^{-1})
\]

\[
A = \text{cross sectional area of pipe (m}^2\)
\]

\[
C_{lp} = \text{liquid specific heat (J·kg}^{-1}·K}^{-1})
\]

\[
C_D = \text{discharge coefficient for pipe inlet}
\]
\[\lambda_c = \text{heat of vaporisation at outlet conditions (J.kg}^{-1}\]\n\[D = \text{pipe diameter (m)}\]
\[L = \text{pipe length (m)}\]
\[n_c = \text{phase equilibrium parameter}\]
\[P_0 = \text{static pressure (Pa)}\]
\[P_c = \text{critical (choke) pressure (Pa)}\]
\[S = \text{entropy (J.K}^{-1}\r\]
\[T_c = \text{temperature at the choke (K)}\]
\[T_0 = \text{initial liquid temperature (K)}\]
\[V_c = \text{specific volume of liquid at choke (m}^3\text{.kg}^{-1}\]
\[X_c = \text{weight fraction of vapour at choke (i.e. quality)}\]
\[\alpha F = \text{simplified friction factor}\]

**Subscripts**

- **G** = vapour phase
- **L** = liquid phase
- **SAT** = saturated condition

\[G\] is found by iteratively varying the choke pressure until a maximum value is found. However, the authors gave no guidance as to how a value for the phase equilibrium parameter \(n_c\) may be interpolated between \(n_c = 1\) for a fully evaporated liquid and \(n_c = 0\) for a superheated liquid. The model is therefore very difficult to use in practice.

The Wheatley model calculates the choke pressure of a release along a tube by iteratively solving the expression

\[V_L(P_0 - P_c) + C_{pl}(T_0 - T_c) - C_{pl}T_c \ln \left(\frac{T_0}{T_c}\right) + gh \frac{C_{pl}T_c^2}{P_c^2A^2}\]

\[\times \left[1 + \left(\frac{A^0}{T_c^2} - 2\right)\ln \left(\frac{T_0}{T_c}\right)\right] = \frac{1}{2} \left[ V_L + C_{pl} \frac{T_c^2}{P_cA^2}\ln \left(\frac{T_0}{T_c}\right) \right]^2 \]  \hspace{1cm} (9.22)

with the temperature at the choke \(T_c\) as the unknown. The flow is choked if \(P_c\), the vapour pressure at \(T_c\), is greater than atmospheric pressure. The model assumes that the flow is

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homogenous, frictionless and adiabatic. The nomenclature of Equation (9.22) is the same as (9.21) with the addition of $A^0$ which is a coefficient in the Antoine vapour pressure correlation:

$$P_{SAT} = e^{(-A^0/T_c) + B}$$  \hspace{1cm} (9.23)

The quality of the outlet flow is given by:

$$X_c = \frac{CPLT_c^2}{AP_c} \left[ \ln \left( \frac{T_o}{T_c} \right) \right]$$  \hspace{1cm} (9.24)

where $V_{V_c}$ is the specific volume of the choked vapour. This is calculated using the Ideal Gas Law:

$$P_cV_{V_c} = \frac{RT_c}{M}$$  \hspace{1cm} (9.25)

where $M$ is the molecular weight (kg.kmol$^{-1}$). The mean specific volume of the choked mixture is obtained from:

$$V_c = V_L + X_c(V_V - V_L)$$  \hspace{1cm} (9.26)

and the velocity (m.s$^{-1}$) from

$$\frac{1}{2}u_c^2 = C_D \left[ \frac{(P_0 - P_c)}{\rho_L} + C_{PL}(T_0 - T_c) - C_{PL} \ln \left( \frac{T_o}{T_4} \right) + gh \right]$$  \hspace{1cm} (9.27)

where $\rho_L$ is the liquid density (kg.m$^{-3}$). The mass flow $G$ is then found from

$$G = \frac{u_cA}{V_c}$$  \hspace{1cm} (9.28)

where $A$ is the cross sectional area of the flow path.

For the purposes of this study the theoretical model given by Wheatley was compared against the empirical models of Fletcher and Johnson and Fauske by considering a typical example of a flashing propane release at 10 bar abs and 290 K through a pipe 0.5 m long with a diameter of 0.025 m i.e. such as may occur through a carelessly opened drain point.
Using Figure 9.1 (water curve):

When \( L = 0.5 \text{ m} \),

\[
\frac{G_p^2}{0.61^2 P_1} \approx 200 \text{ kg.m}^{-3}
\]

therefore

\[
G_p = \sqrt{(0.61)^2 \cdot 10^6 \cdot 200}
\]

since

\[
G = G_p \cdot A
\]

where \( A \) is the cross sectional area then

\[
G = 0.00049 \cdot \sqrt{(0.61)^2 \cdot 10^6 \cdot 200}
\]

\[
G = 4.2 \text{ kg.s}^{-1}
\]

Fauske ERM

Using Equation (9.19):

\[
G_p = \frac{dP}{dT} \sqrt{\frac{T}{NC_{pl}}}
\]  \hspace{1cm} (9.19)

for \( L = 0.5 \text{ m}, N = 1 \). Using physical property data correlations given in Appendix E the following data were derived for propane at 290 K:

\[
C_{pl} = 2600 \text{ J.kg}^{-1}\text{K}^{-1};
\]

\[
\frac{dP}{dT} = 20800 \text{ Pa.K}^{-1}
\]

therefore

\[
G = 0.00049 \cdot 20800 \cdot \sqrt{\frac{290}{2600}}
\]

\[
G = 3.4 \text{ kg.s}^{-1}
\]
The temperature at the choke was first obtained using Equation (9.22). The following physical
property data were applied:

\[ A^0 = 6724; \]
\[ V_L = 2.0 \times 10^{-3} \text{ m}^3 \text{kg}^{-1} \]

Iteration of Equation (9.22) gave a solution where \( T_c = 281 \text{ K} \). From vapour pressure data derived
from the vapour pressure correlation in Appendix E the choke pressure \( P_c \) was approximately
6 bar abs. The quality at the choke conditions was obtained using Equation (9.24) where \( V_{vc} \)
was obtained from Equation (9.25):

\[ V_{vc} = \frac{8314 \cdot 281}{44 \cdot 6 \cdot 10^5} = 0.088 \text{ m}^3 \text{kg}^{-1} \]

\[ X_c = \frac{2600 \cdot 281^2 \cdot \ln(290/281)}{6724 \cdot 6 \cdot 10^5(0.088 - 0.0020)} = 0.019 \]

The average specific volume at the choke was given by Equation (9.26):

\[ V_c = 0.0020 + 0.019(0.088 - 0.0020) \]

The velocity was then obtained using Equation (9.27) with a \( C_D \) value suggested by Wheatley
to be 0.8 for a short pipe:

\[ \frac{1}{2} V_c^2 = 0.8^2 \left( \frac{(10 - 6)10^5}{500} + 2600(290 - 281) - 2600 \cdot 281 \ln(290/281) \right) \]

\[ V_c = 39 \text{ m.s}^{-1} \]

hence from Equation (9.28):

\[ G = \frac{39 \cdot 0.00049}{0.0036} = 5.4 \text{ kg.s}^{-1} \]

The modelled flow rates were similar in all three cases shown. The use of the Fletcher and
Johnson correlation was preferred for use in Section 10 because:

1. It provides a more conservative estimate of flow rate than the Fauske ERM, whilst the
approximate nature of the flow estimation implicitly reflects the considerable
uncertainties involved in evaluating two-phase flows.
2. It is simpler to use than the Wheatley model and may also be used to roughly estimate the non-equilibrium flows where the flow path is relatively short.

In conclusion to this section on two-phase flow model selection it should be noted that the above models were only intended for use in situations where pipe friction is negligible i.e. for relatively short pipes. For long pipes this is not the case as has already been pointed out in Section 9.1.3. For single-phase flows it is necessary to modify the value of $C_D$ to account for the friction of the pipe. This is also true for two-phase flows although the problem is considerably more complex. Leung and Grolmes [69] presented a model which adapts Lapple charts for use with two-phase flow in horizontal ducts, whilst Carter [70] used an incremental approach to solve the differential two-phase flow equation given by Perry [63]:

$$G_F(x, y) = K u^\prime_{w} (2^n) x^{2(2^n)} y^{2(2^n)}$$  \hspace{1cm} (9.31)

and for a circular pool radius $r$:

$$G_F(r) = K' u^\prime_{w} (2^n r^{(4+2)(2^n)})$$  \hspace{1cm} (9.32)

9.1.5 Vaporisation of liquids

The emission rate from a spill or pool of liquid may depend primarily on whether the pool is a volatile liquid below its boiling point e.g. a solvent pool or a cryogenic liquid spill e.g. LNG.

9.1.5.1 Vaporisation of volatile liquids

The model commonly used for estimating evaporation rates from pools of volatile liquid is that of Sutton [71] modified by Pasquill [72]. The vapourisation rate from a rectangular pool $x$ m downwind by $y$ m crosswind is given by:

$$G_F(x, y) = K u^\prime_{w} (2^n) x^{2(2^n)} y^{2(2^n)}$$  \hspace{1cm} (9.31)

and for a circular pool radius $r$:

$$G_F(r) = K' u^\prime_{w} (2^n r^{(4+2)(2^n)})$$  \hspace{1cm} (9.32)

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where

\[ K \text{ or } K' = K_{DVA}^{2n(2+n)} k^{4(1-n)/(2+n)} z_1^{-n/(2+n)} \]  
(9.33)

and where a wind velocity profile is assumed:

\[ \left( \frac{u_w}{u_{w1}} \right) = \left( \frac{z}{z_1} \right)^{\frac{n}{2+n}} \]

and:

\[ \chi_0 = \frac{P_{SAT} \cdot M}{RT} \]

Nomenclature:

- \( G_F \) = vaporisation flux (kg.m\(^2\).s\(^{-1}\))
- \( u_w \) = wind speed at height \( z \) m above ground (m.s\(^{-1}\))
- \( z_1 \) = reference height (m)
- \( u_{w1} \) = wind velocity at reference height \( z_1 \) (m.s\(^{-1}\))
- \( n \) = Sutton atmospheric stability constant
- \( k \) = Von Karman’s constant = 0.4
- \( R \) = gas constant = 8314 J.kmol\(^{-1}\).K\(^{-1}\)
- \( P_{SAT} \) = vapour pressure at temperature \( T \) K (Pa)
- \( D_{VA} \) = diffusion coefficient of vapour in air (m\(^2\).s\(^{-1}\))

Lihou gave empirical values of \( K \) and \( K' \) as a function of \( n \), which was itself an expression of atmospheric stability as given in Table 9.2. Mecklenburgh rearranged Equation (9.32) to give a solution for the volumetric flow \( Q \) m\(^3\).s\(^{-1}\) from a circular pool with diameter \( d_p \) m:

\[ Q = K' u_w d_p^2 \frac{2 \pi P_{SAT}}{R} \left( \frac{2}{u_{w1}^2} \right)^{\frac{n}{2+n}} \]  
(9.34)

where here \( P_A \) is atmospheric pressure (Pa) and \( n_* = n/(2+n) \). Empirical values of \( n \) and \( K' \) are given in Table 9.3. For dispersion over short distances the values usually taken for both the Lihou and the Mecklenburgh empirical coefficients are those corresponding to typical i.e. neutral atmospheric conditions.

If volatile pool evaporation is to be modelled completely then the reduction in vaporisation rate produced by the cooling effect should be taken into account. However it was neglected here in order to give a conservative estimate.
9.1.5.2 Vaporisation of cryogenic liquids

The model generally used to describe the rapid boiling of a spreading pool of cryogenic liquid is that of Shaw and Briscoe [73]. The instantaneous vaporisation rate $G(R)$ from a spreading pool of radius $R$ is given by:

$$G(R) = \frac{Ck(T_s - T_L)}{\lambda(\pi\alpha)^{1/2}} \int_0^R \frac{2\pi r}{(t - t')^{1/2}} \, dr$$

(9.35)

where $(t - t')$ is the time the pool takes to cover a pool annulus of width $dr$. Other nomenclature is:

- $C = 1$ for impenetrable surfaces
- $C = 3$ for soils and other penetrable surfaces
- $k$ = surface thermal conductivity ($W.m^{-1}.K^{-1}$)
- $T_s$ = surface temperature (K)
- $T_L$ = liquid temperature (K)
- $\alpha$ = surface thermal diffusivity ($m^2.s^{-1}$)
- $\lambda$ = latent heat of vaporisation ($J.kg^{-1}$)

Shaw and Briscoe devised a computer code to numerically solve the integral. Separate analytical solutions of Equation (9.35) have been presented by Lihou, TNO and Mecklenburgh for relatively large instantaneous releases. However these are not useful for HAC which is concerned with protecting against the ignition of small continuous releases, or planned releases such as the draining of a pipeline. However, Mecklenburgh also presented a solution to give an approximation of the vaporisation of a continuous leak using the following expression derived from the Shaw and Briscoe model:

$$W = \frac{2Ck(T_s - T_L)}{\lambda} \left( \frac{t}{\pi\alpha} \right)^{1/2} \frac{\pi d_p^2}{4}$$

(9.36)

$W$ kg is the total amount boiled off in time $t$ seconds, where $t$ is either one minute or the time taken for the average volumetric boil-off rate $V/2t \, m^3.s^{-1}$ to decrease to the evaporation rate given by Equation (9.32), whichever is shorter. The pool diameter $d_p$ is a function of $t$ such that it takes the smallest value given by Equations (9.37) and (9.38):

$$d_p^2 = \left( \frac{512gGt^3}{9\pi \rho_L} \right)^{1/2}$$

(9.37)
\[ d_r^2 = \frac{4Gt}{\pi \rho \ell h_{\text{MIN}}} \]  

(9.38)

where \( G \) is the leak rate of the cryogenic liquid (kg s\(^{-1}\)) and \( h_{\text{MIN}} \) is the minimum pool height expected on a given surface. Values for \( h_{\text{MIN}} \) given by Mecklenburgh are tabulated in Table 9.4, whilst Table 9.5 lists values given by Mecklenburgh for the function \( k/(\alpha)^{1/2} \) for different surfaces. Other nomenclature for Equations (9.36) to (9.38) is the same as for the Shaw and Briscoe Equation (9.35). Note that the amount vaporised \( W \) cannot be greater than the total amount emitted \( Gt \). Use of the model for small emissions indicated that the boil-off rate very quickly reaches the emission rate. Since this represents the most pessimistic vaporisation rate estimate the model was preferred despite its limitations.

9.2 Dispersion

9.2.1 Gas jets

The dispersion of a jet is characterised according to whether the flow is subsonic or sonic. The dispersion may also depend on the shape of the leak orifice.

The simplest case is for a subsonic jet issuing from a circular orifice. The established model is that of Ricou and Spalding [74] which gives the axial jet concentration \( C_x \) as:

\[ \frac{C_x}{C_0} = 5 \frac{d_o}{x} \left( \frac{\rho_A}{\rho_o} \right)^{1/2} \]  

(9.39)

where

\[
\begin{align*}
C_x & = \text{axial volumetric concentration at } x \text{ m (m}^3/\text{m}^3) \\
C_0 & = \text{volumetric concentration at outlet (m}^3/\text{m}^3) \\
d_o & = \text{outlet diameter (m)} \\
x & = \text{distance along axis (m)} \\
\rho_A & = \text{ambient density (kg.m}^3) \\
\rho_o & = \text{gas density at outlet (kg.m}^3) 
\end{align*}
\]

The value of 5 is an empirical value of \( 1/\tan^2\theta \), where \( \theta \) is the jet half-angle i.e. the angle made between the jet axis and the edge of the jet at the outlet. This value was given by Birch et al. [75] as applying to jets of ethylene and also separately by Rajaratnam [76] as a general value.
Equation (9.39) applies to circular holes only. For slot-shaped holes such as may occur after a gasket section blow-out the subsonic jet dispersion was modelled by the following expression given by Rajaratnam [76] and Chen and Rodi [77]:

\[
\frac{C_x}{C_0} = 2\left(\frac{2b_o}{x}\right)^{1/2}\left(\frac{\rho_A}{\rho_o}\right)^{1/2}
\]

(9.40)

where \(b_o\) is half the slot width (m).

For sonic underexpanded jet dispersion a separate treatment is required since the flow is choked and the outlet pressure may be several times the ambient. As the flow leaves the outlet it expands and accelerates. A series of expansion waves develop around the expansion point which are reflected as compression "barrel shocks" around the sonic region. These are illustrated in Figure 9.2. Depending on the speed of the jet there may be a series of barrel shocks as the jet velocity approaches subsonic levels at which point the dispersion proceeds as for subsonic jets. Ewan and Moodie [78] modelled the concentration decay of sonic jets using original work by Kleinstein [79]. The concentration decay along the jet axis is given by:

\[
\frac{M_x}{M_0} = 1 - \exp\left[-\frac{1}{k\left(\frac{\rho_A}{\rho_{e2}}\right)^{0.5} x^* \sqrt{\frac{\rho_{e2}}{\rho_o}} - X_t}\right]
\]

(9.41)

where

\[
x^* = x - 2x_b
\]

\[
x_b = 0.77d_o + 0.068d_0^{1.35}N
\]

(9.41a)

\[
\rho_{e2} = \rho_o\left(\frac{P_A}{P_C}\right)
\]

\[
P_C = \text{choke pressure} = P_i\left(\frac{2}{\gamma + 1}\right)^{\frac{\gamma}{\gamma-1}}
\]

\[
N = \frac{P_C}{P_A}
\]

and

\[
d_{e2} = d_0\left(\frac{C_D P_C}{P_A}\right)^{1/2}
\]
Nomenclature:

\[ M/M_0 \quad = \quad \text{mass fraction on jet axis (kg/kg)} \]
\[ x \quad = \quad \text{axial distance along jet (m)} \]
\[ x_b \quad = \quad \text{barrel length (m)} \]
\[ x^* \quad = \quad \text{modified axial distance (m)} \]
\[ d_o \quad = \quad \text{outlet diameter (m)} \]
\[ d_{EQ} \quad = \quad \text{equivalent outlet diameter (m)} \]
\[ r_{EQ} \quad = \quad \text{equivalent outlet radius (m)} \]
\[ \rho_c \quad = \quad \text{density at choke pressure } P_c \text{Pa (kg.m}^3) \]
\[ \rho_{EQ} \quad = \quad \text{equivalent density (kg.m}^3) \]
\[ \rho_A \quad = \quad \text{ambient density (kg.m}^3) \]
\[ C_D \quad = \quad \text{outlet discharge coefficient} \]
\[ k' \quad = \quad \text{eddy viscosity coefficient taken to be 0.107 for under expanded jets} \]
\[ X_c \quad = \quad \text{non-dimensional core-length taken to be 0.7} \]

The units of Equation (9.41a) are mm.

The equivalent values of density and outlet diameter are the values at which subsonic dispersion commences after the series of barrel shocks.

The calculated concentration on the jet axis is only an average value and may not be the maximum at any given time. To allow for this, the peak-to-mean concentration ratio of 1.5 suggested by Long [80] was applied to the calculated distance to the LEL to arrive at the overall hazard distances shown in Section 10.

The sonic jet models given here are only applicable to jets from round holes which are unimpeded. Unfortunately no expression equivalent to Equation (9.40) has been found for slot-shaped sonic jets, and as a first approximation in Section 10 Equation (9.41) was used for this case. If a jet impinges on a solid object then some of the momentum is lost and the dispersion is impeded. How much of the momentum is lost depends on each individual situation and so generalisations cannot be made. In Section 10 it was necessary to conservatively assume that the jet lost all momentum in situations where impingement seemed likely.
9.2.2 Liquid jets

For an intact liquid jet (see Section 9.1.4.1) it is necessary to estimate the distance travelled before it strikes the ground and forms a pool. This may be calculated using the relation for the horizontal distance travelled by a projectile.

For horizontal jets:

\[ x = u_L \left( \frac{2y}{g} \right)^{1/2} \]  

(9.42)

and for inclined jets at grade:

\[ x = \frac{u_L^2}{g} \sin 2\theta \]  

(9.43)

where \( y \) is the height of the source above grade (m), \( g \) is the gravitational constant (m.s\(^{-2}\)), \( u_L \) is the initial liquid jet velocity (m.s\(^{-1}\)) inclined at \( \theta \) degrees to the horizontal. Equations (9.41) and (9.42) neglect both air friction and the break up of the jet along its trajectory which may both affect the overall distance travelled. No correction for jet break up has been found, however a correction for air friction may be derived using theory presented by Brown [81] who assumed that friction is proportional to particle velocity in the horizontal and vertical directions:

\[ \frac{d^2x}{dt^2} = -k_D \frac{dx}{dt} \]  

(9.44)

\[ \frac{d^2y}{dt^2} = -k_D \frac{dy}{dt} - g \]  

(9.45)

Solving the two equations by substitution and using the boundary conditions:

when \( t = 0, \quad x = 0, \quad \frac{dx}{dt} = u_L \cos \theta, \quad \frac{dy}{dt} = u_L \sin \theta, \quad y = 0 \)

gives

\[ x = \frac{u_L \cos \theta}{k_D} \left( 1 - e^{-bt} \right) \]  

(9.46)

\[ k_D y + gt = \frac{k_D u_L \sin \theta + g}{k_D(1 - e^{-bt})} \]  

(9.47)
Empirical evaluation is then required for the drag coefficient \( k_D' \), which Ooms et al. estimated as 0.3 for plumes [82]. This model gives a more refined estimate than the rule-of-thumb quoted by O’Shea [83] that in practice a liquid jet travels about half the theoretical value.

The dispersion of two-phase spray flows and atomised liquid jets is considerably more complex to model. As has been indicated, where spray is predicted rain-out is seldom a problem assuming there are no physical objects impeding the flow. For the calculations described in Section 10 it was tentatively assumed that two-phase flashing releases could be treated the same as underexpanded sonic gas jets as a first approximation, which resulted in reasonable hazard distances. A similarly tentative assumption was made for the estimation of spray dispersion in Section 10, which were treated as neutrally buoyant plumes with no rain-out and rapidly vaporising droplets due to the violent motion of the jet upon emission.

9.2.3 Atmospheric gas dispersion

Atmospheric gas dispersion methods are applicable when the turbulence forces of the atmosphere are greater than the momentum and buoyancy forces of the emitted gas plume e.g. a jet of gas of about the same density of air will be dispersed by the atmosphere when its velocity reduces to that of the ambient wind [55].

The established method of estimating atmospheric dispersion is to model the emitted plume using the Gaussian, or normal, distribution. The model most widely used is that of Pasquill and Gifford [84] based on work by Sutton [71]. The version of the model used in Section 10 to estimate hazard distances is that of TNO [59] where the concentration \( C \) (kg.m\(^{-3}\)) at a point \((x,y,z)\) downwind of a source with origin \((0,0,h)\) is given by:

\[
C(x,y,z) = \frac{G}{2\pi u_w \sigma_y \sigma_z} \left[ \exp\left(\frac{-y^2}{2\sigma_y^2}\right) \right] \times \left[ \exp\left(\frac{-(z-h)^2}{2\sigma_z^2}\right) + \exp\left(\frac{-(z+h)^2}{2\sigma_z^2}\right) \right]
\]  

where \( G \) is the source strength (kg.s\(^{-1}\)) and \( u_w \) is the windspeed (m.s\(^{-1}\)). The plume centre-line concentration may be calculated simply by setting \( x \) and \( y = 0 \) and \( z = h \). TNO assumed the
standard deviations $\sigma_y$ and $\sigma_z$ are primarily functions of distance and atmospheric stability i.e. the amount of atmospheric turbulence such that:

\begin{align*}
\sigma_y &= a x^b \\
\sigma_z &= c x^d
\end{align*}

(9.49)
(9.50)

where $a$, $b$, $c$ and $d$ are empirically determined constants. Table 9.6 lists TNO values of these constants for the range of Pasquill stability categories from A to F where A is the most unstable (promoting dispersion) and F is the most stable (restricting dispersion). The examples considered in Section 10 concerned dispersion over relatively short distances, and so category D (i.e. neutral) conditions were assumed.

TNO also introduced empirical correction factors for $\sigma_y$ and $\sigma_z$ to account for surface roughness and peak-to-mean plume concentration variation. The surface roughness factor $C_{z_0}$ is applied as follows:

\begin{equation}
\sigma_z = C_{z_0} x^d
\end{equation}

(9.51)

where

\begin{equation}
C_{z_0} = (10 \sigma_0)^{0.51}\times^{-0.22}
\end{equation}

(9.52)

and $z_0$ is a measure of the ground roughness termed the roughness length. The correction is only applicable for dispersion over large distances, however. For short range dispersion $C_{z_0}$ is set to one i.e. $z_0 = 0.1$ m. The peak-to-mean concentration correction factor $C_{TAV}$ is applied in a similar fashion:

\begin{equation}
\sigma_y = C_{TAV} x^b
\end{equation}

(9.53)

where

\begin{equation}
C_{TAV} = \left( \frac{t_{AV}}{600} \right)^{0.2}
\end{equation}

(9.54)

and where $t_{AV}$ is the sample time for the average plume concentration (s). TNO specified that $C_{TAV}$ should have a minimum value of 0.5, which corresponds to $t_{AV} = 20$ s or a peak-to-mean concentration ratio of 2. This value was used for the dispersion calculations described in Section 10 to calculate hazard distances to half-LEL. This error margin was chosen as a result of conclusions drawn by Britter and McQuaid [85] that Gaussian dispersion models provide concentration estimates accurate to about ±40%.
9.2.3.1 Applicability of Gaussian methods to short-range dispersion

Gaussian dispersion models are usually applied to estimate the dispersion of large releases over hundreds of metres e.g. chimney stack emissions. It was therefore necessary to validate the use of Gaussian models for short distances i.e. less than 50 m, since these hazard distances correspond to the magnitude of the emissions usually encountered in HAC. Such validation may be found from several sources. Sutton [86] concluded that his model was applicable to columns of hot air in a laboratory as well as large outdoor releases. Elsewhere, Katan [87] adapted the Sutton model to estimate the dispersion of aviation fuel vapour from filling points between distances of 0.20 m to 25 m. The applicability of the Sutton model to dispersion over short distances was also discussed by Long [80] who found no fundamental reason why it should not be used. Other work where broad agreement has been found between Gaussian predictions and experimental observations includes that of Dicken and Shaw [88] who measured concentrations within small chlorine plumes; Welker, who experimented with small propane releases [89]; and Lapin and Foster [90] who worked with small vaporising pools of liquid oxygen. It was thus concluded from the above references that Equation (9.48) is applicable to neutral buoyancy gas dispersion over short distances, and also dense gas dispersion under certain conditions which are described below.

9.2.3.2 Applicability of Gaussian methods to small dense gas releases

Gaussian dispersion methods may be used to model dense releases when the atmospheric turbulence forces are greater than the gravitational force exerted on the plume. The relative magnitude of these forces is expressed by the dimensionless Richardson number \( Ri \) which was defined by Hanna and Drivas [91] as:

\[
Ri = \frac{\text{Potential energy of cloud}}{\text{Turbulent energy of environment}}
\]  

(9.55)

For a continuous source the Richardson number at the outlet may be defined as:

\[
Ri_0 = \frac{g}{\bar{u}_0^2} \left( \frac{\rho_i - \rho_a}{\rho_a} \right) \left( \frac{G}{\rho_i \bar{u}_w d_{cw}} \right)
\]  

(9.56)

where
\[ \rho_i = \text{initial release density (kg.m}^{-3} \) \\
\rho_a = \text{ambient density (kg.m}^{-3} \) \\
g = \text{gravitational constant} = 9.81 \text{ m.s}^{-2} \\
G = \text{continuous source strength (kg.s}^{-1} \) \\
d_{cw} = \text{crosswind diameter of continuous release source (m) } \\
u_w = \text{wind velocity (m.s}^{-1} \) \\
\]

and \( u_* \) is the friction velocity of the wind close to the ground. \( u_* \) may be obtained from:

\[
\frac{1}{u_*} = \frac{1}{k u_w} \ln \left( \frac{z_w}{z_0} \right)
\]

(9.57)

where \( k \) is the von Karman constant i.e. approximately 0.4, \( z_w \) is a reference height at which the windspeed is \( u_w \) and \( z_0 \) is the roughness length from Equation (9.52) i.e. 0.1 m [10].

Havens and Spicer [92] stated that for \( Ri < 1 \) the dispersion is controlled by atmospheric turbulence whereas for \( Ri > 32 \) the dispersion is gravity controlled and Gaussian models may not be applicable. Between these figures the dispersion mode is uncertain although Hanna and Drivas concluded that a reasonable critical value \( R_{ic} \) to take for the transition point between atmospheric and gravity dispersion is \( R_{ic} = 10 \). This then gives a criterion as to whether Gaussian models may be applied to an initial vapour release.

If the outlet Richardson number is above the critical value Hanna and Drivas estimated the transition distance \( x_T \) travelled by the plume until the local Richardson number falls below \( R_{ic} \) using:

\[
x_T = \frac{E_0}{(V_0)^{1/3} u^2 R_{ic}}
\]

(9.58)

where \( R_{ic} \) is the critical Richardson number, \( V_0 \) the initial volumetric source strength (\( \text{m}^3.\text{s}^{-1} \)) and \( E_0 \) is the initial potential energy of the release given by:

\[
E_0 = g V_0 \left( \frac{\rho_i - \rho_a}{\rho_a} \right)
\]

(9.59)

After the plume has travelled \( x_T \) the plume dispersion may be estimated using Equation (9.48). The total distance gives a first-guess estimate of overall dispersion distance for dense releases.


9.2.3.3 Applicability of Gaussian methods in low windspeeds

Gaussian dispersion methods depend on reliable estimates of windspeed to enable downwind concentrations to be estimated. Whilst there is no theoretical limit to the maximum windspeed that may be modelled, Equation (9.48) is unreliable if the windspeed falls below a certain limit given by TNO as 1 m.s\(^{-1}\) [59]. Due to the relative low probability of such low windspeeds occurring outdoors the problem of modelling dispersion in near-calm conditions is seldom addressed. For example, British Gas [34] adopted 2 m.s\(^{-1}\) as the most pessimistic windspeed for outdoor HAC problems since it was estimated that there is an 85% probability that the average UK windspeed lies between 2 and 10 m.s\(^{-1}\). However, in most indoor situations the airsPEED is nearly always below 1 m.s\(^{-1}\) and it is here that there is a problem with estimating dispersion. The modelling of releases in indoor plant is considered in Section 13.4.

To summarise, this section has described a literature survey for mathematical models which may be used to estimate the size and consequences of typical flammable emissions. In considering the types of models essential to adequately describe each release it was necessary to consider the release behaviour. Consequently, models were selected which could be used sequentially in model "chains". These are used in Section 10 to show the effects of release size, phase and process conditions on the resulting hazard distances for selected release scenarios. The models described in this section were therefore primarily chosen for illustrative purposes. They were not intended to form a definitive list and where more sophisticated, realistic models exist it should be possible to substitute them into the model chain without affecting the overall modelling strategy.
Table 9.1 Published models for calculating the hazard distances resulting from flammable releases

<table>
<thead>
<tr>
<th>Process</th>
<th>Referenced text (see below)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>1</td>
</tr>
<tr>
<td>Liquid flow</td>
<td>x</td>
</tr>
<tr>
<td>Liquid $C_D$</td>
<td>x</td>
</tr>
<tr>
<td>Gas/vapour flow - sonic</td>
<td>x</td>
</tr>
<tr>
<td>- subsonic</td>
<td>x</td>
</tr>
<tr>
<td>Gas $C_D$</td>
<td>x</td>
</tr>
<tr>
<td>Two-phase flow</td>
<td>x</td>
</tr>
<tr>
<td>- flash fraction</td>
<td>x</td>
</tr>
<tr>
<td>- spray fraction</td>
<td>x</td>
</tr>
<tr>
<td>Volatile liquid evaporation</td>
<td>x</td>
</tr>
<tr>
<td>Cryogenic pool boiling</td>
<td>x</td>
</tr>
<tr>
<td>- instantaneous release</td>
<td></td>
</tr>
<tr>
<td>- continuous release</td>
<td></td>
</tr>
<tr>
<td>Liquid jet throw</td>
<td>x</td>
</tr>
<tr>
<td>Gas jet dispersion</td>
<td>x</td>
</tr>
<tr>
<td>- sonic</td>
<td></td>
</tr>
<tr>
<td>- subsonic</td>
<td></td>
</tr>
<tr>
<td>Atmospheric dispersion</td>
<td>x</td>
</tr>
<tr>
<td>- point source at grade</td>
<td></td>
</tr>
<tr>
<td>- elevated point source</td>
<td></td>
</tr>
<tr>
<td>Peak-to-mean correction</td>
<td>x</td>
</tr>
</tbody>
</table>

Column nomenclature:

Table 9.2 Values of $n$, $K$ and $K'$ for use with the Sutton-Pasquill evaporation model (from Lhou [54])

<table>
<thead>
<tr>
<th>Stability</th>
<th>$n$</th>
<th>$K \times 10^3$</th>
<th>$K' \times 10^3$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Unstable</td>
<td>0.20</td>
<td>1.278</td>
<td>3.846</td>
</tr>
<tr>
<td>Neutral</td>
<td>0.25</td>
<td>1.570</td>
<td>4.685</td>
</tr>
<tr>
<td>Stable</td>
<td>0.30</td>
<td>1.786</td>
<td>5.285</td>
</tr>
</tbody>
</table>

Table 9.3 Values of $n$, and $K'$ for use with the Sutton-Pasquill evaporation model (from Mecklenburgh [56])

<table>
<thead>
<tr>
<th>Stability</th>
<th>Pasquill stability category</th>
<th>$n$</th>
<th>$n_*$</th>
<th>$K' \times 10^3$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Unstable</td>
<td>A</td>
<td>0.17</td>
<td>0.078</td>
<td>1.0</td>
</tr>
<tr>
<td></td>
<td>B</td>
<td>0.20</td>
<td>0.091</td>
<td>1.2</td>
</tr>
<tr>
<td></td>
<td>C</td>
<td>0.25</td>
<td>0.111</td>
<td>1.5</td>
</tr>
<tr>
<td>Neutral</td>
<td>D</td>
<td>0.30</td>
<td>0.130</td>
<td>1.7</td>
</tr>
<tr>
<td></td>
<td>E</td>
<td>0.35</td>
<td>0.149</td>
<td>1.8</td>
</tr>
<tr>
<td>Stable</td>
<td>F</td>
<td>0.44</td>
<td>0.180</td>
<td>1.8</td>
</tr>
</tbody>
</table>
Table 9.4 Estimated values of minimum pool height $h_{\text{MIN}}$ for use with the Shaw and Briscoe model (from Mecklenburgh [56])

<table>
<thead>
<tr>
<th>Surface</th>
<th>Minimum pool height (mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rough, sandy soil</td>
<td>25</td>
</tr>
<tr>
<td>Farm land</td>
<td>20</td>
</tr>
<tr>
<td>Smooth sand, gravel</td>
<td>10</td>
</tr>
<tr>
<td>Concrete, stone</td>
<td>5</td>
</tr>
</tbody>
</table>

Table 9.5 Values of the conductivity/diffusivity function $k/(\alpha)^{1/2}$ for use with the Shaw and Briscoe model (from Mecklenburgh [56])

<table>
<thead>
<tr>
<th>Surface</th>
<th>$k/(\alpha)^{1/2}$ (W.s$^{1/2}$.m$^{-2}$.K$^{-1}$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Concrete</td>
<td>1430</td>
</tr>
<tr>
<td>Soil (average)</td>
<td>1420</td>
</tr>
<tr>
<td>Soil (dry sandy)</td>
<td>580</td>
</tr>
<tr>
<td>Soil (80% moisture and sand)</td>
<td>1020</td>
</tr>
</tbody>
</table>
Table 9.6  Atmospheric stability constants given by TNO [59] for use with the Pasquill-Gifford model

<table>
<thead>
<tr>
<th>Stability and Pasquill category</th>
<th>a</th>
<th>b</th>
<th>c</th>
<th>d</th>
</tr>
</thead>
<tbody>
<tr>
<td>Very unstable</td>
<td>A 0.527</td>
<td>0.865</td>
<td>0.28</td>
<td>0.90</td>
</tr>
<tr>
<td>Unstable</td>
<td>B 0.371</td>
<td>0.866</td>
<td>0.23</td>
<td>0.85</td>
</tr>
<tr>
<td>Slightly unstable</td>
<td>C 0.209</td>
<td>0.897</td>
<td>0.22</td>
<td>0.80</td>
</tr>
<tr>
<td>Neutral</td>
<td>D 0.128</td>
<td>0.905</td>
<td>0.20</td>
<td>0.76</td>
</tr>
<tr>
<td>Stable</td>
<td>E 0.098</td>
<td>0.902</td>
<td>0.15</td>
<td>0.73</td>
</tr>
<tr>
<td>Very stable</td>
<td>F 0.065</td>
<td>0.902</td>
<td>0.12</td>
<td>0.67</td>
</tr>
</tbody>
</table>
Figure 9.1 Two-phase flow curves for (top) Freon-11 and (bottom) water. Taken from Fletcher and Johnson [66]
Figure 9.2 Flow regions around a nozzle emitting a sonic under expanded jet ($M = \text{Mach number}$). Taken from Ewan and Moodie [78]
10. Estimation of hazard ranges

This section discusses how selected models from the literature search described in Section 9 may be used to estimate the extent of typical hazardous zones around some of the leak sources given in Table 5.1. Example hazard distances were calculated, with the results presented at the end of this section, to show how the estimation of hazard distances may be quantified and to investigate the assumptions that have to be made in the course of the estimation process.

To facilitate the use of the calculation models in sequential fashion the relevant equations from each model where appropriate were transcribed into a series of short interactive FORTRAN 77 programs so that the answers from one routine could be used in the next step in the model chain. The program listings for the following models may be found in Appendix D:

1. Sonic gas emission rate  \hspace{1cm} \text{Equation (9.3)}
2. Liquid emission rate  \hspace{1cm} \text{Equation (9.5)}
3. Test for liquid atomisation  \hspace{1cm} \text{Equation (9.6)}
4. Evaporation rate from volatile pool  \hspace{1cm} \text{Equation (9.34)}
5. Underexpanded jet dispersion  \hspace{1cm} \text{Equation (9.41)}
6. Liquid jet throw  \hspace{1cm} \text{Equation (9.44)}
7. Atmospheric dispersion  \hspace{1cm} \text{Equation (9.48)}
8. Test for gravity dispersion  \hspace{1cm} \text{Equation (9.56)}

Other models used in this section which were given in Section 9 were either charted correlations or one-line equations, for example the models describing two-phase emission rate and the estimation of flash fraction. These more simple models were used manually. The leak scenarios considered in this section were the following:

1. Flange leaks
   - gas
   - flashing liquid
   - non-flashing liquid
2. Compressors - gas
3. Pumps - flashing liquid
   - non-flashing liquid
4. Sample/drain points - flashing liquid

As well as comparing the hazard distances generated by different equipment items the other objective of this analysis was to investigate how the hazard distances varied with different flammable materials of the same type e.g. gases and liquids. To illustrate this, six representative flammable substances were selected for use with the calculation models. These were:

1. Hydrogen (H₂)
2. Methane (CH₄)
3. Ethylene (C₂H₄)
4. Propane (C₃H₈)
5. Acetone (C₃H₆O)
6. Methyl Ethyl Ketone (MEK) (C₄H₇O)

The behaviour of these substances upon release was considered typical of flammable materials handled in the process industries. In order to model the release of a particular substance it was necessary to obtain physical property data pertaining to the substance. Data for the six substances named above and the correlations used in the data derivation are given in Appendix E. A short FORTRAN routine is listed in Appendix D which was written to apply these correlations to generate temperature dependent physical property data. The correlations are applicable to pure substances only therefore when dealing with mixtures care should be taken that appropriate data are applied.

As discussed in Section 8, leak hole sizes were selected as being representative of the values given for each equipment item. A preliminary assessment of leak severity was made to show the difference in hazard distances caused by severe leaks. The hole sizes indicated do not represent recommended values for global use but are an indication of the order of magnitude that may be expected for each category of leak based on the information given in Section 8. For example, the flange leak sizes used in Tables 10.1 - 10.3 were selected to represent the following cases:
Small leak on a SWJ/RTJ gasket = 0.25 mm²
Large leak on a SWJ/RTJ gasket =
Small leak on a CAF gasket = 2.5 mm²
Large leak on a CAF gasket = 25 mm²

Similarly, the leak sizes used for pumps and compressors in Tables 10.4 and 10.5 were chosen to reflect both leak severity and equipment (i.e. shaft) size. The assumption of hole size distributions for process equipment is approached again in the formulation of a process plant fire and explosion model in Section 12.

10.1 Calculation of hazard distances

Tables 10.1 - 10.6 list results of the hazard distance calculations carried out. In all cases the temperature of the substance before emission was taken to be ambient which has been set to 298 K here. The other assumptions made are given below where appropriate.

Table 10.1 Hazard distances for flange leaks: gas

It was assumed that the gas dispersed as a circular underexpanded jet with no obstructions and that the overall hazard distance was the distance to 0.67 × LEL to allow for possible peak-to-mean concentration variations (see Section 9.2.1). In the absence of topographical information about each individual leak site it was assumed for all the leak scenarios considered in Tables 10.1-10.6 that the hazardous zone formed a spherical volume concentric to the leak source.

Table 10.2 Hazard distances for flange leaks: non-flashing liquid

For the examples considered in Table 10.2B, the liquid jets were tested for atomisation using Equation (9.6) and were estimated to atomise with a maximum drop size of about 0.5 mm at 10 bar abs and 0.2 mm at 20 bar abs using Equation (9.9). It was assumed that the drops quickly vaporised and that the subsequent hazard distance for each leak shown in Table 10.2B. could be estimated by combining the Gaussian dispersion distance calculated from Equation (9.48) and the gravity/Gaussian dispersion transition distance calculated from Equation (9.58). The overall figures are listed with the transition distance shown in brackets. This method provided a conservative estimate of the hazard distance without the use of far more sophisticated
dense gas dispersion methods (a brief review of some of these models e.g. DEGADIS, by Havens and Spicer [92]; or CRUNCH, by Jagger [93], is given by Puttock [94]), because it was assumed that there was negligible dispersion before the transition point. It also provided a simple method of assessing the relative importance of gravity-induced dispersion for a given release and gave an albeit approximate measure of the overall dispersion distance. However, note that for large release rates e.g. for a leak hole size of 25 mm², the transition distance becomes appreciable; hence for these cases analysis of the release using more sophisticated models may become necessary if the zone distance is to be realistic.

A windspeed of 2 m.s⁻¹ was assumed for all the Gaussian dispersion calculations listed in this section, together with neutral atmospheric stability (i.e. Pasquill category D). A peak-to-mean concentration ratio of 2 was also assumed, thus giving an overall hazard distance as the distance to 0.5 × LEL.

The figures given in Table 10.2C show the more conservative case where the liquid jet was assumed to remain intact and was projected in the worst case at an angle of 45° upwards and away from the leak source. The hazard distance was estimated by combining the throw of the jet incorporating air resistance, the radius of the ensuing pool at the point of impact, and the dispersion distance for the evaporated vapour. The pool size was taken to be the value at which the rate of pool spread ceased to be significant, which was considered here to be below 1 mm.s⁻¹. Note that although the jet was assumed to atomise in Table 10.2B, this calculation was performed to show how the results compare with the atomisation case.

Table 10.3 Hazard distances for flange leaks: flashing liquids

For the case of a flashing liquid propane jet from a gasket leak it was assumed that there was no flashing upstream of the orifice and that the jet flashed completely on emission. It was further assumed that the subsequent dispersion could be approximately modelled as an under expanded jet and that the overall hazard range was the distance to 0.67 × LEL.

Table 10.4 Hazard distances for compressor leaks

The release was assumed to be a sonic jet emitted from the failed seal which was impeded by the compressor body and consequently lost all momentum. The subsequent dispersion was therefore assumed to be Gaussian according to the conditions assumed above. It was not
overlooked that both hydrogen and methane are buoyant gases which rise upon release; however the specification of a spherical hazardous zone around each leak source was intended to reflect the fact the plume may disperse in any direction.

Table 10.5 Hazard distances for pump leaks

For the purposes of these calculations it was assumed that leaks of non-flashing liquids from pump seals collect in a 1 m$^2$ bund and that the dispersion of the evaporating vapour was consequently Gaussian. For leaks of flashing liquid it was assumed that the liquid was emitted as a jet which impinged on the pump body and flashed completely. The hazard distance was therefore obtained by combining the gravity/Gaussian transition distance (shown in brackets) with the Gaussian dispersion distance.

Table 10.6 Hazard distances for drain/sample point leaks: flashing liquid

The results shown are of a full-bore release from a drain/sample point pointing downwards at the ground. The spray/flash fraction was estimated using Equation (9.17) to be roughly 0.8. The subsequent dispersion was calculated using the same procedure as for the flashing liquid release in Table 10.5.

10.2 Interpretation of results

The results shown in Tables 10.1 - 10.6 indicate that the difference in physical properties for each substance does not have a great effect on the calculated hazard distances, in situations where the substances are of the same phase and emitted under the same conditions. For example, the hazard distances around ethylene and methane compressors are similar for the same leak hole sizes. The exception to this general rule is hydrogen, which is anomalous due to its low density which results in a high mass rate for an emission through a given leak hole size. However, open-air hydrogen releases do not normally create an extraordinary hazard compared to an equivalent release of, say, LPG, since its high positive buoyancy results in the release being rapidly transported upwards and thus away from the release source and surrounding ignition sources. This aids the dispersion of the release so that large flammable clouds of hydrogen seldom have chance to accumulate.
Hazard distances vary significantly according to mode of release for a given substance. An example is an ethylene emission through a 25 mm$^2$ leak hole in a gasket (Table 10.1) and a compressor seal (Table 10.4). In both cases the gas was assumed to emit as a sonic jet. However, it was envisaged that the compressor leak would probably be impeded by the equipment casing. Since it was not possible to accurately model the rate at which the jet loses its momentum in such circumstances it was conservatively assumed that all momentum was lost and that the proceeding dispersion was Gaussian. The hazard distances shown in Tables 10.1C and Tables 10.4C show that where Gaussian dispersion is assumed this results in hazard distances which are double those produced by jet dispersion for the release sizes that HAC is concerned with.

Similarly, the results indicate that flashing liquid releases do not disperse as quickly as gas emissions for releases through the same leak hole size i.e. releases occurring on the same type of equipment item. The hazard distances in Table 10.5B show that whereas an impeded flashing propane release through a 25 mm$^2$ hole at 20 bar abs. disperses to a safe level over 41 m, an impeded ethylene release under the same conditions disperses to a safe level after a distance of 12 m. It was concluded that this effect is principally due to the high mass rate of flashing liquid releases compared to equivalent gas releases, which then both disperse as gases. This results in a greater ignition hazard for flashing releases, which is discussed further in Section 12. However, this effect may apply only to impeded jet releases, since the hazard distances for unimpeded jets shown in Table 10.3C and Table 10.1C indicate that the violent mixing action inherent in jet dispersion results in similar dispersion distances for jets both of flashing liquid and of gas, subject to the assumptions made.

Returning to flange leaks, two sets of non-flashing liquid release calculations were performed according the assumptions laid out in Section 10.1. It was concluded from these calculations that the results in Table 10.2C are rather unrealistic. For example, they indicate that a liquid jet of about 0.5 mm diameter remains intact when projected 64 m. However, it was also felt that the spray dispersion results listed in Table 10.2B were probably pessimistic, since it was assumed that there was no rain-out from the spray. For releases in relatively confined areas this may not be the case since if the plume impinges on an object some of the spray will rain out on surrounding obstacles, resulting in a decrease in the flammable inventory of the plume and hence a reduction in the hazard distance. Notwithstanding the possibility of rain-out it was concluded that the hazard distance resulting from the atomised spray provided the more credible estimate of the true situation.
Lastly, it is possible to compare some of the results given in this Section with the Zone 2 distances for selected equipment items given in Figures 2.1 and 2.2. For example, the results in Table 10.5 for propane should be expected to compare with the Zone 2 distances for LPG pumps shown in Figure 2.1. However, with the exception of the ICI/RoSPA zone, the codes are evidently based on smaller size releases than are considered in Table 10.5, or are seemingly based on releases of non-flashing liquids. This points up one of the main inadequacies of current codes which specify standard hazard distances. The authors of the different codes may have had releases of specific size and phase in mind when setting zone distances; however it is seldom clear what the details of these releases were. It is therefore very difficult for an industrial operator who wishes to comply with a given code to know what release rate and phase the code distance was based upon, and thus what degree of conservatism is involved. This is also true of the zone distances given in Figure 2.2 for LPG pipe fittings: whereas the ICI/RoSPA distance is based on a stated hole size of $40 \text{ mm}^2$ (see Section 8.2) it is not clear how the other codes achieve smaller zone distances for apparently the same leak scenario. It was thus concluded that whilst the models discussed in Section 9 provided a reasonable estimate of the behaviour of small releases under different conditions, to obtain better estimates of zone distances requires quantification of the size of release that HAC is expected to provide ignition protection against. More data gathering is therefore required to quantify leak size and severity fraction.

To summarise, example calculations were performed to illustrate how the mathematical models described in Section 9 may be used to estimate hazard distances around typical leak sources. A range of leak sizes was used in each case to show how hazard distances may vary with leak severity. Also, a selection of flammable substances of differing physical properties was used with each leak scenario to investigate how this affected the calculated hazard distance. The calculations showed some inadequacies in the models selected, in particular in the dispersion of atomised sprays. However, it was possible to conclude that hazard distances were mostly dependent on the phase of the emitted substance rather than individual physical parameters such as density, vapour pressure etc. Hazard distances were also shown to be highly dependent on the mode of dispersion. A comparison of the calculated hazard distances with Zone 2 distances suggested by various HAC standards and codes showed that for some release scenarios (most notably flashing liquid releases) the codes apparently underestimate the dispersion distance to a safe level. It is therefore necessary that codes which give standard zone sizes should quantify the conditions from which the distances were derived, and should give guidance to the bounds of applicability for each case.
Table 10.1 Hazard distances for flange leaks: gas

A. Flow (kg.s\(^{-1}\) × 10\(^3\))

<table>
<thead>
<tr>
<th>Pressure (Pa × 10(^5))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.092</td>
<td>0.92</td>
</tr>
<tr>
<td>Methane</td>
<td>0.26</td>
<td>2.6</td>
</tr>
<tr>
<td>Ethylene</td>
<td>0.34</td>
<td>3.4</td>
</tr>
</tbody>
</table>

B. Distance to LEL (m)

<table>
<thead>
<tr>
<th>Pressure (Pa × 10(^5))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.62</td>
<td>2.0</td>
</tr>
<tr>
<td>Methane</td>
<td>0.19</td>
<td>0.60</td>
</tr>
<tr>
<td>Ethylene</td>
<td>0.26</td>
<td>0.84</td>
</tr>
</tbody>
</table>

C. Distance to 0.67 × LEL (m)

<table>
<thead>
<tr>
<th>Pressure (Pa × 10(^5))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.94</td>
<td>3.0</td>
</tr>
<tr>
<td>Methane</td>
<td>0.28</td>
<td>0.89</td>
</tr>
<tr>
<td>Ethylene</td>
<td>0.30</td>
<td>1.3</td>
</tr>
</tbody>
</table>
Table 10.2 Hazard distances for flange leaks: non-flashing liquid

A. Flow (kg.s\(^{-1}\) \(\times 10^3\))

<table>
<thead>
<tr>
<th>Pressure (Pa (\times 10^5))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>6.0</td>
<td>60</td>
</tr>
<tr>
<td>MEK</td>
<td>6.0</td>
<td>60</td>
</tr>
</tbody>
</table>

B. Distance to LEL assuming atomisation of liquid jet (m)*

<table>
<thead>
<tr>
<th>Pressure (Pa (\times 10^5))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>2.5</td>
<td>8.6</td>
</tr>
<tr>
<td>(0.6)</td>
<td>2.8</td>
<td>(13)</td>
</tr>
<tr>
<td>MEK</td>
<td>2.6</td>
<td>9.3</td>
</tr>
<tr>
<td>(0.8)</td>
<td>3.6</td>
<td>(17)</td>
</tr>
</tbody>
</table>

Distance to 0.5 \(\times\) LEL assuming atomisation of liquid jet (m)*

<table>
<thead>
<tr>
<th>Pressure (Pa (\times 10^5))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>3.2</td>
<td>11</td>
</tr>
<tr>
<td>MEK</td>
<td>3.4</td>
<td>11</td>
</tr>
</tbody>
</table>

*see footnote on page 135

continued
Table 10.2 continued

C. Overall hazard distance assuming that liquid jet does remains a continuous stream (m)

C.1 Throw of liquid jet (m)

<table>
<thead>
<tr>
<th>Pressure (Pa x 10^5)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>37</td>
<td>65</td>
</tr>
<tr>
<td>MEK</td>
<td>36</td>
<td>64</td>
</tr>
</tbody>
</table>

C.2 Radius of pool at cessation of significant growth rate (m)

<table>
<thead>
<tr>
<th>Pressure (Pa x 10^5)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm^2)</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>0.22</td>
<td>1.7</td>
</tr>
<tr>
<td>MEK</td>
<td>0.23</td>
<td>2.0</td>
</tr>
</tbody>
</table>

C.3 Pool evaporation rate at cessation of significant growth rate (kg.s^-1 x 10^3)

<table>
<thead>
<tr>
<th>Pressure (Pa x 10^5)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm^2)</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>0.4</td>
<td>17</td>
</tr>
<tr>
<td>MEK</td>
<td>0.2</td>
<td>11</td>
</tr>
</tbody>
</table>

continued
Table 10.2 continued

C.4 Overall distance to LEL (m) (jet throw + pool radius + plume length to LEL)

<table>
<thead>
<tr>
<th></th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Pressure (Pa (\times 10^5))</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>0.25</td>
</tr>
<tr>
<td>Acetone</td>
<td>38</td>
<td>66</td>
</tr>
<tr>
<td>MEK</td>
<td>37</td>
<td>65</td>
</tr>
<tr>
<td><strong>Get throw</strong> + pool radius + plume length to LEL</td>
<td>25</td>
<td>25</td>
</tr>
<tr>
<td><strong>Get throw</strong> + pool radius + plume length to 0.5 (\times) LEL</td>
<td>10</td>
<td>10</td>
</tr>
</tbody>
</table>

C.5 Overall hazard distance (m) (jet throw + pool radius + plume length to 0.5 \(\times\) LEL)

<table>
<thead>
<tr>
<th></th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Pressure (Pa (\times 10^5))</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>0.25</td>
</tr>
<tr>
<td>Acetone</td>
<td>38</td>
<td>67</td>
</tr>
<tr>
<td>MEK</td>
<td>37</td>
<td>66</td>
</tr>
</tbody>
</table>

*The figures given in brackets in Tables 10.2B, Table 10.5 and Table 10.6 represent the transition distance in each example for the change from gravity-controlled dispersion to Gaussian dispersion. It was conservatively assumed that no dilution takes place until the transition distance is reached to give a crude measure of the effect of gravity-controlled dispersion on dense plumes. The distances to LEL and half-LEL given in these tables include the bracketed figures. Where no bracketed figures are given in the distance to LEL for examples where the vapour is heavier than air (e.g. Table 10.2C.4), the transition distance was calculated as zero i.e. the plume was Gaussian upon emission.*
Table 10.3 Hazard distances for flange leaks: flashing liquids

A. Flow (kg/s$^1 \times 10^3$)

<table>
<thead>
<tr>
<th>Pressure (Pa $\times 10^5$)</th>
<th>Hole size (mm$^2$)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.25</td>
<td>2.5</td>
<td>25</td>
</tr>
<tr>
<td></td>
<td>0.25</td>
<td>2.5</td>
<td>25</td>
</tr>
<tr>
<td>Propane</td>
<td>4.7</td>
<td>47</td>
<td>470</td>
</tr>
<tr>
<td></td>
<td>0.30</td>
<td>0.96</td>
<td>3.0</td>
</tr>
<tr>
<td></td>
<td>0.30</td>
<td>0.44</td>
<td>1.4</td>
</tr>
</tbody>
</table>

B. Distance to LEL (m)

<table>
<thead>
<tr>
<th>Pressure (Pa $\times 10^5$)</th>
<th>Hole size (mm$^2$)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.25</td>
<td>2.5</td>
<td>25</td>
</tr>
<tr>
<td></td>
<td>0.25</td>
<td>2.5</td>
<td>25</td>
</tr>
<tr>
<td>Propane</td>
<td>0.30</td>
<td>0.96</td>
<td>3.0</td>
</tr>
<tr>
<td></td>
<td>0.30</td>
<td>0.44</td>
<td>1.4</td>
</tr>
</tbody>
</table>

C. Distance to 0.67 $\times$ LEL (m)

<table>
<thead>
<tr>
<th>Pressure (Pa $\times 10^5$)</th>
<th>Hole size (mm$^2$)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.25</td>
<td>2.5</td>
<td>25</td>
</tr>
<tr>
<td></td>
<td>0.25</td>
<td>2.5</td>
<td>25</td>
</tr>
<tr>
<td>Propane</td>
<td>0.45</td>
<td>1.5</td>
<td>4.5</td>
</tr>
<tr>
<td></td>
<td>0.45</td>
<td>1.5</td>
<td>4.5</td>
</tr>
</tbody>
</table>

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Table 10.4 Hazard distances for compressor leaks

A. Flow (kg.s\(^{-1}\) × 10\(^3\))

<table>
<thead>
<tr>
<th>Hole size (mm(^2))</th>
<th>Pressure (Pa × 10(^{5}))</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>10</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>9.2</td>
</tr>
<tr>
<td>Methane</td>
<td>26</td>
</tr>
<tr>
<td>Ethylene</td>
<td>34</td>
</tr>
</tbody>
</table>

B. Distance to LEL (m)

<table>
<thead>
<tr>
<th>Hole size (mm(^2))</th>
<th>Pressure (Pa × 10(^{5}))</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>10</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>8.9</td>
</tr>
<tr>
<td>Methane</td>
<td>4.8</td>
</tr>
<tr>
<td>Ethylene</td>
<td>5.6</td>
</tr>
</tbody>
</table>

C. Distance to 0.5 × LEL (m)

<table>
<thead>
<tr>
<th>Hole size (mm(^2))</th>
<th>Pressure (Pa × 10(^{5}))</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>10</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>13</td>
</tr>
<tr>
<td>Methane</td>
<td>6.8</td>
</tr>
<tr>
<td>Ethylene</td>
<td>8.0</td>
</tr>
</tbody>
</table>
Table 10.5 Hazard distances for pump leaks

A. Flow (kg.s\(^{-1}\) × 10\(^3\))

<table>
<thead>
<tr>
<th>Pressure (Pa × 10(^{5}))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>6.0</td>
<td>60</td>
</tr>
<tr>
<td>MEK</td>
<td>6.0</td>
<td>60</td>
</tr>
<tr>
<td>Propane</td>
<td>4.7</td>
<td>47</td>
</tr>
</tbody>
</table>

B. Distance to LEL (m)*

<table>
<thead>
<tr>
<th>Pressure (Pa × 10(^{5}))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>2.1</td>
<td>all</td>
</tr>
<tr>
<td>MEK</td>
<td>1.3</td>
<td>cases</td>
</tr>
<tr>
<td>Propane</td>
<td>2.2</td>
<td>7.5</td>
</tr>
<tr>
<td></td>
<td>(0.3)</td>
<td>(1.5)</td>
</tr>
</tbody>
</table>

C. Distance to 0.5 × LEL (m)*

<table>
<thead>
<tr>
<th>Pressure (Pa × 10(^{5}))</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hole size (mm(^2))</td>
<td>0.25</td>
<td>2.5</td>
</tr>
<tr>
<td>Acetone</td>
<td>2.9</td>
<td>all</td>
</tr>
<tr>
<td>MEK</td>
<td>1.8</td>
<td>cases</td>
</tr>
<tr>
<td>Propane</td>
<td>3.0</td>
<td>10</td>
</tr>
</tbody>
</table>

* see footnote on page 135
Table 10.6 Hazard distances for drain/sample point leaks: flashing liquid

The orifice is a 15 mm diameter sample point 500 mm long

A. Flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th>Pressure (Pa x 10⁻⁵)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Propane</td>
<td>1.5</td>
<td>2.1</td>
</tr>
</tbody>
</table>

B. Distance to LEL (m)*

<table>
<thead>
<tr>
<th>Pressure (Pa x 10⁻⁵)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Propane</td>
<td>50</td>
<td>60</td>
</tr>
<tr>
<td></td>
<td>(15)</td>
<td>(19)</td>
</tr>
</tbody>
</table>

C. Distance to 0.5 x LEL (m)*

<table>
<thead>
<tr>
<th>Pressure (Pa x 10⁻⁵)</th>
<th>10</th>
<th>20</th>
</tr>
</thead>
<tbody>
<tr>
<td>Propane</td>
<td>64</td>
<td>77</td>
</tr>
</tbody>
</table>

*see footnote on page 135
11. Probability of leak ignition

So far, for an individual leak source, data have been presented to enable estimates to be made of:

- Leak frequency (Section 7)
- Leak hole size (Section 8)
- Emission rate (Section 9.1)
- Hazard distance (Section 9.2)

A further requirement to classifying a hazardous area is a method of establishing the probability of leak ignition within the hazard distance. If such information can be established then a link may be made to the overall criterion used for setting the tolerability of flammable risk, thus leading to the classification of the hazardous zone.

11.1 Ignition sources

It was initially necessary to define the ignition sources that may give rise to fire and explosion incidents. Some information has already been given in Section 4 concerning the contribution made to the overall number of fire and explosion incidents in the UK by different ignition sources for 1986-1987. The data for "closed" i.e. pressurised plant are summarised in Table 11.1. More ignition source data concerning offshore incidents have been published by Forsth of Det Norske Veritas (DNV)[95][96]. The published ignition data related to 133 incidents in the Norwegian North Sea and 133 incidents in the Gulf of Mexico and are summarised in Table 11.2. Both Tables 11.1 and 11.2 were adjusted to omit ignition incidents involving hotwork since it was noted in Appendix B that where leaks were ignited by hotwork it was the hotwork itself which initiated the leak. The tables were also adjusted to omit incidents where the ignition source remained unknown.
Care should be exercised in comparing offshore and onshore ignition data as the dispersion of leaks on an offshore plant is not the same as for the equivalent onshore plant due to its compactness. Despite this, there are some basic similarities between the ignition data presented in the two tables, e.g. the majority of igninations were caused by flames and hot surfaces, including engines and their exhausts, whilst electrical-type ignitions were in the minority. This suggests that despite obvious differences in plant layout, fires and explosions are initiated in a similar manner for both types of plant. More comparisons are drawn between onshore and offshore ignitions in the literature survey described next.

11.2 Ignition probability

In the literature survey for estimates of leak ignition probability, it was found necessary to distinguish between the probability of ignition and the probability of explosion given ignition since statistics and estimates have been published for both. The explosion probability literature survey is described in Section 11.3.

Kletz [97]

\[
\text{Ignition probability of small leaks} = 10^4
\]

This was a judgemental estimate based on small vapour jet leaks on a polyethylene plant. Kletz also provided the following leak ignition estimate for hydrogen/hydrocarbon plant at 250 bar:

\[
\text{Ignition probability} = 0.033
\]

The above figure is for a range of mostly small leak sizes. Kletz argued that ignition probability increases with release size so that for large vapour clouds, i.e. greater than 10 tons, the ignition probability is between 0.1 and 0.5. The formulation of a crude ignition model in Section 11.2.1 is based on this proposition.
Browning [98]

Relative ignition estimates were given as follows:

<table>
<thead>
<tr>
<th>Release type</th>
<th>Relative ignition probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Massive LPG release</td>
<td>$10^1$</td>
</tr>
<tr>
<td>Flammable liquid with flashpoint below 110°F (43°C)</td>
<td>$10^2$</td>
</tr>
<tr>
<td>or with temperature above flashpoint</td>
<td></td>
</tr>
<tr>
<td>Flammable liquid with flashpoint 110-200°F (43-93°C)</td>
<td>$10^3$</td>
</tr>
</tbody>
</table>

The figures given were based on conditions of no obvious source of ignition and with explosion-proof, i.e. protected electrics. It was assumed here that the figures relate to releases of equal size although this was not made clear in the reference. For releases into unprotected areas Browning recommended that the appropriate value be multiplied by 10. Elsewhere [99] Browning gave absolute values of ignition probability based on the protected case of $10^{-1}$ - $10^{-2}$ per spill.

The Canvey Reports [23][100]

The first Canvey Report assumed the following ignition probabilities:

<table>
<thead>
<tr>
<th>Release type</th>
<th>Ignition probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Limited release (tens of tons)</td>
<td>0.1</td>
</tr>
<tr>
<td>Large releases</td>
<td>1</td>
</tr>
</tbody>
</table>

The second Canvey Report used a different method of estimation by giving the following values for on-site ignitions:

<table>
<thead>
<tr>
<th>Sources of ignition</th>
<th>Ignition probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Nominally zero</td>
<td>0.1</td>
</tr>
<tr>
<td>Very few</td>
<td>0.2</td>
</tr>
<tr>
<td>Few</td>
<td>0.5</td>
</tr>
<tr>
<td>Many</td>
<td>0.9</td>
</tr>
</tbody>
</table>
Examples of areas where these ignition source categories exist were given as follows:

Nominally zero: "None readily identifiable" e.g. limited release of liquid hydrocarbon into a bund after the overfilling of a tank.

Very few: Large release of liquefied gas under pressure after a catastrophic failure of a tank in a tank farm.

Few: Release of flammable material near to non-continuous operations e.g. an LPG release from a tank next to a road/rail facility.

Many: Release of flammable material near to a plant or resulting from a nearby fire or explosion.

This method provides relative ignition probabilities for a fixed release size emitted into an area with varying ignition source density. It does not necessarily provide a method for differentiating between release sizes.

The Rijnmond Report [28]

The Rijnmond Report assigned ignition probabilities to sources in the path of the plume. For a given plume size and wind direction the overall ignition probability is therefore the sum of the individual probabilities and so varies according to each major release considered. However, the report gave fixed estimates of initial release ignition probability which was the probability of an ignition source close to the release source causing a fire before the plume is fully developed (this may include auto-ignition). This was set to 0.1 for all major release sources.

Dahl et al.[101]

The foregoing references all gave values of probability essentially based on expert judgement. Actual data regarding ignition frequency from which to derive probability estimates are very sparse. One such set of data is that of Dahl et al. which referred to blow-outs on offshore drilling and production platforms. In an analysis of 172 blow-outs occurring between 1956 and 1980 Dahl listed some 135 that resulted in ignition, distinguishing between incidents involving explosion and fire, and those just involving fire. The total number is given in Table 11.3 together with the derived ignition frequency. 90% confidence limits for the derived frequency using the $\chi^2$ distribution are also shown. The numbers of ignitions are specific to the fluid indicated so
that for example blow-outs listed as involving both oil and gas were omitted. It was inferred that the ignition frequencies shown relate to massive releases; and that since they are expressed as ignition frequency given release, they are equivalent to the ignition probabilities given above.

11.2.1 Selection of ignition probability values

As discussed previously, the sparsity of available ignition data for either small or large releases made it difficult to derive a simple estimate to use. The confidence limits shown for the Dahl data indicate a very wide spread of values within which the true value may lie. Nevertheless such a selection was necessary to enable the quantification of ignition probability and to give meaning to the hazard distances calculated in Section 10. It was therefore decided to devise an ignition model, according to the previously discussed suggestion by Kletz: that ignition probability varies according to emission rate. Ignition probability also varies with ignition source density as pointed out by the Canvey report. However, the intention here was to take an average ignition source density over a wide range of plants that reflected typical zoning practice with respect to the location of ignition sources. In other words, the derived probability values were intended to represent the result of applying the mixture of codes of practice shown in Figures 2.1 - 2.3. This enabled the foregoing spread of expert judgement estimates to be put into the context of leak size alone.

The probability estimates for three levels of leak size were therefore defined for this study: minor, major; and massive. A minor leak was defined as being below 1 kg.s\(^{-1}\); a massive leak was defined as being above 50 kg.s\(^{-1}\). An intermediate, or major leak lay between the two limits. Ignition probabilities were therefore assigned to these leak levels. Furthermore, since the Browning estimates and the Dahl data offered some evidence of ignition probability dependency on phase, a distinction was thus made between the ignition probabilities of gas and liquid releases.

The massive leaks were first considered as follows. The Dahl frequency estimate of 0.3 for massive gas leaks is consistent with Kletz’s upper probability value of 0.5 for land-based vapour clouds. A working value of 0.3 was therefore adopted. Also, the Dahl estimate of 0.08 for the ignition frequency of a massive oil release is close to the upper limit suggested by Browning of 0.1 for flammable gas/liquid spills. Hence 0.08 was adopted as a working value of massive liquid release ignition probability (the overall preference was for smaller values where there was some justification).
For minor liquid spills as defined above, the lower limit of the Browning ignition probability was adopted, i.e. 0.01. This value was also taken for minor gas releases as it lies between the Kletz estimates of 0.033 for mostly small high pressure leaks and $10^{-4}$ for very small ethylene leaks.

A crude ignition model was therefore constructed by plotting the estimates made on log-log paper as shown in Figure 11.1. The archetypal leak sizes which were taken to exemplify the limits minor leaks $< 1 \text{ kg.s}^{-1}$ and major leaks $> 50 \text{ kg.s}^{-1}$ were 0.5 and 100 kg.s$^{-1}$ respectively.

The treatment of ignition probability so far described is very speculative because it was designed to circumvent the paucity of available data. It was realised that, taken by itself, the ignition model described above contains too much uncertainty to be applied with any degree of confidence. With this consideration in mind, a series of cross-checks was devised to test the model using annual fire and explosion statistics. This is described in Section 12.

11.3 Explosion probability

Values of explosion probability were examined concurrently to those of ignition probability. Before giving the results it is necessary to distinguish between the probability of explosion given a leak and the probability of explosion given ignition. The former is the product of the probability of explosion given ignition and the probability of ignition given a leak. It is the latter expression which is most commonly given in the literature.

Kletz [97]

Kletz gave the following values of explosion probability given ignition:

<table>
<thead>
<tr>
<th>Size of cloud (tons)</th>
<th>Explosion probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>&gt; 0.1</td>
</tr>
<tr>
<td>&lt; 1</td>
<td>0.01 - 0.001</td>
</tr>
</tbody>
</table>

Kletz also stated that there is no minimum amount of flammable material required to form an explosive mixture.
The Canvey Reports [23][100]

The first Canvey Report provided the following value of explosion probability for refineries which was taken here to relate to massive releases:

Probability of explosion given major fire = 0.5

For large vapour clouds the following values were given:

<table>
<thead>
<tr>
<th>Release type</th>
<th>Explosion probability given ignition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Large vapour clouds (~ 1000 tons)</td>
<td>1</td>
</tr>
<tr>
<td>Smaller clouds of gases other than methane (&lt; 100 tons)</td>
<td>0.1</td>
</tr>
<tr>
<td>Smaller clouds of methane (&lt; 100 tons)</td>
<td>0.01</td>
</tr>
</tbody>
</table>

Dahl et al. [101]

As well as ignition frequency data, the blow-out event information mentioned in Section 11.2 yielded explosion data which is given in Table 11.3. Again, this was taken to relate to massive vapour releases.

Sofyanos [102]

More data concerning the number of offshore fires and explosions in the Gulf of Mexico were published by DNV which related to 326 incidents occurring between 1956 and 1979. These are summarised in Table 11.4. The incidents are ranked in five damage categories where Category I is the most severe and involves the loss of the platform whilst Category V incidents are deemed of no consequence. The categories were based on property damage rather than loss of life; therefore these data were particularly useful because it was possible to infer a degree of severity from them.
11.3.1 Selection of explosion probability values

A basic explosion model similar to the ignition model presented in Section 11.2.1 was derived from the values given above. The main difference between the two models is that since explosions are due to releases of vapour rather than liquid there is thus no need to specify the probability of explosion given ignition of a release of liquid.

For massive leaks above 50 kg s\(^{-1}\), the explosion probability at the archetypal leak size of 100 kg s\(^{-1}\) was taken to be 0.5 as a first estimate. This corresponds to the Dahl and the Sofyanos Damage Category I estimated frequencies. It is also in broad agreement with the judgemental values given by Kletz and by the Canvey Reports.

For minor leaks below 1 kg s\(^{-1}\) the explosion probability of a leak of 0.5 kg s\(^{-1}\) was taken to be the Sofyanos Category V explosion probability of 0.04.

These points were plotted on log-log paper in Figure 11.2 so that intermediate (major) leak explosion probabilities could be interpolated.

To summarise, a literature survey was carried out to obtain data from which to derive ignition probabilities which could be applied to the source strength calculations. From the limited amount of information available concerning major incidents, crude ignition and explosion models were constructed. It was realised that there was considerable uncertainty in the data which should be improved on if the models are to be of use in HAC. A method of producing refined leak and ignition estimates from the event information described in the foregoing sections is described in Section 12.
Table 11.1 Ignition source summary adjusted to omit hotwork and unknown sources (from Appendix B)

<table>
<thead>
<tr>
<th>Ignition source</th>
<th>Number of incidents</th>
<th>Percent (unadjusted)</th>
<th>Percent (adjusted)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flame : general</td>
<td>8</td>
<td>9.3</td>
<td>16.7</td>
</tr>
<tr>
<td>: LPG fired eqpt.</td>
<td>2</td>
<td>2.3</td>
<td>4.1</td>
</tr>
<tr>
<td>Hot surfaces</td>
<td>10</td>
<td>11.6</td>
<td>20.8</td>
</tr>
<tr>
<td>Friction</td>
<td>4</td>
<td>4.7</td>
<td>8.4</td>
</tr>
<tr>
<td>Electrical</td>
<td>8</td>
<td>9.3</td>
<td>16.7</td>
</tr>
<tr>
<td>Hot particles</td>
<td>3</td>
<td>3.5</td>
<td>6.3</td>
</tr>
<tr>
<td>Static electricity</td>
<td>6</td>
<td>7.0</td>
<td>12.5</td>
</tr>
<tr>
<td>Smoking</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Autoignition</td>
<td>7</td>
<td>8.1</td>
<td>14.5</td>
</tr>
<tr>
<td>Unknown</td>
<td>38</td>
<td>44.2</td>
<td>-</td>
</tr>
<tr>
<td>Total</td>
<td>86</td>
<td>100.0</td>
<td>100.0</td>
</tr>
</tbody>
</table>
Table 11.2 Ignition sources for offshore fire and explosion incidents [95][96]

<table>
<thead>
<tr>
<th>Ignition source</th>
<th>Incident number</th>
<th>Percent (unadj.)</th>
<th>Percent (adj.)</th>
<th>Incident number</th>
<th>Percent (unadj.)</th>
<th>Percent (adj.)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Welding/cutting/</td>
<td>80</td>
<td>59</td>
<td>-</td>
<td>24</td>
<td>18</td>
<td>-</td>
</tr>
<tr>
<td>grinding</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Engines/exhausts</td>
<td>5</td>
<td>4</td>
<td>13.5</td>
<td>45</td>
<td>34</td>
<td>51.8</td>
</tr>
<tr>
<td>Sparks</td>
<td>1</td>
<td>1</td>
<td>2.7</td>
<td>5</td>
<td>4</td>
<td>5.7</td>
</tr>
<tr>
<td>Electrics</td>
<td>12</td>
<td>9</td>
<td>32.4</td>
<td>21</td>
<td>16</td>
<td>24.2</td>
</tr>
<tr>
<td>Hot surfaces</td>
<td>8</td>
<td>6</td>
<td>21.6</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Self-ignition</td>
<td>2</td>
<td>1.5</td>
<td>5.4</td>
<td>1</td>
<td>1</td>
<td>1.1</td>
</tr>
<tr>
<td>Cigarette, lighter, match</td>
<td>2</td>
<td>1.5</td>
<td>5.4</td>
<td>1</td>
<td>1</td>
<td>1.1</td>
</tr>
<tr>
<td>Other</td>
<td>7</td>
<td>5</td>
<td>18.8</td>
<td>14</td>
<td>11</td>
<td>16.1</td>
</tr>
<tr>
<td>Unknown</td>
<td>16</td>
<td>12</td>
<td>-</td>
<td>22</td>
<td>17</td>
<td>-</td>
</tr>
<tr>
<td>Total</td>
<td>133</td>
<td>100</td>
<td>100.0</td>
<td>133</td>
<td>100</td>
<td>100.0</td>
</tr>
</tbody>
</table>

Table 11.3 Ignition and explosion frequency data derived from Dahl [101]

<table>
<thead>
<tr>
<th>Fluid</th>
<th>Total blow-outs</th>
<th>Number of ignitions</th>
<th>90% lower limit</th>
<th>Ignition frequency</th>
<th>90% upper limit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>123</td>
<td>35</td>
<td>0.2</td>
<td>0.3</td>
<td>0.4</td>
</tr>
<tr>
<td>Oil</td>
<td>12</td>
<td>1</td>
<td>0.004</td>
<td>0.08</td>
<td>0.4</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Fluid</th>
<th>Total blow-outs</th>
<th>Number of ignitions</th>
<th>90% lower limit</th>
<th>Explosion frequency</th>
<th>90% upper limit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>35</td>
<td>12</td>
<td>0.2</td>
<td>0.3</td>
<td>0.6</td>
</tr>
</tbody>
</table>
Table 11.4 Explosion frequency data (derived from Sofyanos [102])

<table>
<thead>
<tr>
<th>Damage category</th>
<th>Number of fires and explosions</th>
<th>Number of explosions</th>
<th>90% lower limit</th>
<th>Explosion frequency</th>
<th>90% upper limit</th>
</tr>
</thead>
<tbody>
<tr>
<td>I</td>
<td>9</td>
<td>5</td>
<td>0.22</td>
<td>0.56</td>
<td>1</td>
</tr>
<tr>
<td>II</td>
<td>13</td>
<td>3</td>
<td>0.063</td>
<td>0.23</td>
<td>0.60</td>
</tr>
<tr>
<td>III</td>
<td>33</td>
<td>6</td>
<td>0.079</td>
<td>0.18</td>
<td>0.36</td>
</tr>
<tr>
<td>IV</td>
<td>128</td>
<td>22</td>
<td>0.12</td>
<td>0.17</td>
<td>0.25</td>
</tr>
<tr>
<td>V</td>
<td>143</td>
<td>6</td>
<td>0.018</td>
<td>0.042</td>
<td>0.083</td>
</tr>
<tr>
<td>Total</td>
<td>326</td>
<td>42</td>
<td>0.098</td>
<td>0.13</td>
<td>0.17</td>
</tr>
</tbody>
</table>
Figure 11.1 Initial estimate of ignition probability with respect to leak flow rate.
Figure 11.2 Initial estimate of explosion probability given ignition with respect to leak flow rate
12. Fire and explosion model

The data so far presented constitute the elements of a model to predict the frequency of ignition for a collection of leak sources such as may be found on a process plant, or section of process plant. This section describes the construction and purpose of such a model which links the component data sources discussed previously, i.e:

Leak source inventory (Section 5)
Source leak frequency (Section 7)
Source leak hole size (Section 8)
Source strength (Section 9)
Ignition probability (Section 11)

The final result of the proposed model is the estimation of the number of fires and explosions per plant. Consequently, by using the major plant inventory data given in Section 6, overall numbers of ignition incidents may be generated. These may then be compared to historical data to adjust the best-estimates from the above list which form the basis of the model. Broad agreement with the historical statistics would then suggest that the selected best-estimates were representative of actual conditions.

12.1 Construction of the model

To arrive at an estimate for the overall number of fire and explosion incidents with which to make a comparison, a typical "standard plant" was constructed using the leak source inventory information given in Section 5. The intention was to establish an average plant containing an average number of each leak source to which could be applied the leak and ignition information data to arrive at an average ignition frequency. This would then be multiplied by the number of plants at risk to arrive at the final estimate. This somewhat crude concept employs a number of
assumptions which are sometimes sweeping in nature (further refinement of the model is discussed in Section 14). However, it was necessary to devise this model for the purposes of HAC for the following reasons.

As outlined in the project strategy given in Section 3 the ideal quantification method for classifying the hazardous area around a source of release would be to take the leak source and, by using confidence tested leak frequency and hole size distributions, derive a source strength distribution. From the source strength the estimation of ignition probability should be possible in principle. However, it became apparent during the course of the study that whilst estimates for each of these parameters were available there were considerable uncertainties in their derivation due to the scarcity of data. The resulting data therefore incurred very wide confidence limits. For individual source ignition probability estimates the combined confidence limits for these data would render the results meaningless. It was therefore decided to use the collected data in a different way: to devise a model that would not only generate overall ignition estimates but also leak and ignition frequencies by size, source and fluid phase. The model would therefore need to incorporate and combine leak data for all the sources so far considered. By combining the data in this way it was thought that the model could be validated not only in terms of the overall ignition event frequencies but also crosschecked in terms of the intermediate figures where historical data existed. For example, not only should the overall frequency of fires be consistent with available statistics but also the distribution of fire frequencies with respect to fluid phase. It was also envisaged that by crosschecking final and intermediate results, errors made in selecting raw data for input would be propagated by the effect produced on the subsequent calculational stages. If the event frequencies could be made to agree with historical data it was thought that this point provided a convincing basis from which to examine the nature of the ignition incidents resulting in fires and explosions. For example, it would be necessary to establish the proportion of ignition incidents which may be prevented by the control of ignition sources within the hazard area. It should be stressed that the fire and explosion model described in this section is primarily useful in the estimation of realistic leak data. The further work necessary before the model may be directly applied to HAC is discussed in Section 14.

The formulation of a "standard" plant provided a starting point around which to base available data. The synthesis of this plant is described below.
12.1.1 Leak source inventory

The principal leak sources on typical process plant were identified in Section 5 as:

- Flanges
- Valves
- Pipes
- Pumps/Compressors
- Small bore connections

(Emissions from relief vents were discounted since these were taken to be controlled emissions i.e. emissions which are usually vented through a stack.)

With the exception of small bore connections, inventory examples were provided from which a typical overall inventory was constructed. In this case an inventory for a medium-sized plant was drawn up based mainly on the detailed information given in Material take-off 1 for the hydrocarbon plant. To produce a more manageable profile the number of pipe diameters was condensed to four: 25, 50, 100 and 300 mm nominal bore. Numbers of small connections (taken to be 10 mm NB) were assumed using data given by Anyakora et al. concerning plant instruments [103]. The final inventory is shown in Table 12.1.

To enable the calculation of source strengths it was also necessary to define a fluid phase split for the listed leak sources. It was assumed that petrochemical plants typically handle mainly flashing liquids in the most quantities, which is partly supported by the estimated phase split of Material take-off 1 given in Table 5.19. For the standard plant the fraction handling flashing liquids was conservatively estimated to be higher, as shown in Table 12.1.

12.1.2 Leak frequency distribution

For each source considered a leak frequency distribution was assumed using the data sources listed in Section 7. Each distribution is in the form of a $x \times y$ array, where $x$ is the source connection diameter and $y$ is the severity of the leak (explained in Section 12.1.3). The original best-estimates of the source frequencies used in the model are given in brackets in Table 12.2.
Section 12.4 explains how these initial estimates were iterated to achieve ignition event frequencies comparable with historical data, with the result that the final leak frequencies are shown unbracketed in Table 12.2.

12.1.3 Leak hole size distribution

As described in Section 8.8, the hole size distribution for each source type was partitioned into a number of severity levels for the purposes of the fire and explosion model. For pipes, valves, pumps and small connections the worst-case leak was taken to be complete rupture or breakage with consequent leak area given by $A$, the cross-sectional area of the source. A major leak was taken to be $0.1 \times A$ whilst a minor leak was taken to be $0.01 \times A$ with the exception of small bore connections where minor leaks were neglected. This is in general agreement with the leak severity information given in Sections 7 and 8, and resulted in a $4 \times 3$ array for each source (except for small connection leaks which were described by a $4 \times 2$ array). For flanges a major leak was taken to represent a blow-out of a section between two bolts whilst a minor leak was $0.1 \times$ the blow-out area, resulting in an overall $4 \times 2$ array. In this simplified case gasket types were not differentiated except as part of the overall gasket leak frequency i.e. most of the major releases may be expected to be CAF gaskets whilst the minor releases would include all the reinforced gasket failures.

12.2 Description of fire and explosion model

Having produced a series of arrays describing source inventory, source leak frequency and leak size the devised fire and explosion model utilises them in the manner shown in Figure 12.1. This shows how the calculation sequence is executed to arrive at the overall ignition event frequencies. A FORTRAN computer program was written to enable numerical analysis of the model. This listing of this program (named IGNITION) is given in Appendix F and the calculation sequence is described below.

1. The overall inventories of pumps, pipes etc. are split according to size and fluid phase.
2. The relevant leak frequency array is applied to each source inventory array element to arrive at an \( x \times y \) array for source leak frequency per plant.year.

3. Using the simple hole size distribution array a source strength array is derived by applying the process conditions to simplified emission models given in Section 9. For each model the following conditions were assumed:

<table>
<thead>
<tr>
<th>Liquid phase</th>
<th>Gas phase</th>
</tr>
</thead>
<tbody>
<tr>
<td>average liquid density</td>
<td>molecular weight</td>
</tr>
<tr>
<td>average plant pressure</td>
<td>adiabatic gas coefficient</td>
</tr>
<tr>
<td>average plant temperature</td>
<td>(ratio of specific heats)</td>
</tr>
<tr>
<td>leak orifice ( C_D )</td>
<td></td>
</tr>
<tr>
<td>guillotine pipe break ( C_D )</td>
<td></td>
</tr>
</tbody>
</table>

\[
\begin{align*}
\text{average liquid density} & = 600 \text{ kg.m}^{-3} \\
\text{average plant pressure} & = 15 \text{ bar abs.} \\
\text{average plant temperature} & = 288 \text{ K} \\
\text{leak orifice } C_D & = 0.6 \\
\text{guillotine pipe break } C_D & = 0.8 \\
\text{molecular weight} & = 44 \text{ kg.kmol}^{-1} \\
\text{adiabatic gas coefficient} & = 1.4 \\
\end{align*}
\]

Two-phase flow was assumed to be approximately \( 0.8 \times \) the equivalent liquid flow. Guillotine pipe break emission rates were multiplied by 1.5 to account for the fraction of fracture cases involving flow from both ends of the broken pipe. It was felt that whereas it is usual for risk assessment purposes to assume flow from both pipe ends in all such cases, this model should attempt to simulate actual conditions where two-way flow may not invariably occur after every pipe fracture.

4. The simple ignition model described in Section 11 is applied to the emission flow arrays for each source size, type and fluid phase to produce corresponding fire and explosion probability and frequency arrays. Modifications made to the original ignition model included ignition probability cut-off points for emission rates less than \( 0.5 \text{ kg.s}^{-1} \) and greater than \( 100 \text{ kg.s}^{-1} \) which correspond to the fixed point ignition probabilities inputted at this stage. The assumption was made that there is a finite ignition level for very large leaks beyond which the ignition probability does not increase (this is implied by the literature information given in Section 11). Equally, for very small leaks it was assumed that there is always a small probability of ignition. This reasoning was also applied to the estimation of explosion probability given ignition where a similar cut-off point is
used for very large leaks. However, there is no such cut-off used for very small leaks since at low emission rates it is virtually impossible to form an explosive mixture in an open-air situation such as envisaged here.

At this point the program prints out the arrays so far calculated.

5. The program then sorts each frequency array element depending on whether the corresponding emission rate element is less than 1 kg.s\(^{-1}\) (minimum), above 50 kg.s\(^{-1}\) (maximum), or between 1 and 50 kg.s\(^{-1}\) (medium). These are then summed for each source type and fluid phase to give intermediate frequency totals e.g. the frequency of fires resulting in flashing liquid leaks below 1 kg.s\(^{-1}\). The intermediate totals are then summed to give overall totals of ignition incidents per plant per year. Lastly, all totals are printed.

12.3 Fire and explosion frequency literature survey

A survey of available fire and explosion statistics was conducted for information with which to compare the estimates provided by the above model.

12.3.1 Fires

Information on refinery fires is given in Reported Fire Losses in the Petroleum Industry published annually by the API. The reports for the years 1983, 1984 and 1985 [104],[105],[106] gave the number of refinery fires occurring in the years 1982-85 as 173, 201, 144 and 104 respectively. These figures were broken down by cost as shown in Table 12.3 which was assumed to be directly proportional to fire size. It was assumed that major fires corresponded to damage losses of over $100,000, hence for the years in question about one third of the fires were classified as major. The fire frequency was estimated using refinery inventory information derived in a similar fashion to the refinery inventory data given in Section 6 [107],[108],[109]. The frequencies for the years 1982, 1984 and 1985 are summarised in Table 12.4, where the overall fire frequency per refinery/year is about 0.25. This estimate is of the same order of magnitude as the major fire frequency estimate of 0.1 per refinery/year given by the first Canvey Report [23]. The difference may be related to the cut-off point for major fires as defined by the Canvey report being defined
by other criteria as well as cost. Finally, in Section 6 it was assumed that each refinery consists
on average of five processing units. For each refinery unit the major fire frequency was therefore
taken to be 0.02 - 0.05 fires per plant.year, which was the estimate that the fire and explosion
model was tested against.

Further data on fires in the UK process industry have been given by Redpath [110] which
is referred to more in more detail in Section 12.4.

12.3.2 Explosions

Two surveys of vapour cloud explosions have been performed by Davenport [111][112].
In the second survey Davenport recorded 34 explosions occurring between 1962 and 1982
inclusive which were broken down by location thus:

<table>
<thead>
<tr>
<th>Location</th>
<th>Number</th>
</tr>
</thead>
<tbody>
<tr>
<td>United States</td>
<td>19</td>
</tr>
<tr>
<td>Western Europe</td>
<td>9</td>
</tr>
<tr>
<td>Other</td>
<td>6</td>
</tr>
</tbody>
</table>

Davenport also gave details of the leak source or event leading to the explosion in each case.
These are listed and split into source categories in Tables 12.5 and 12.6.

An estimate was made in Section 6 of the number of plants where there was a major risk
of explosion. This excluded such sites as LPG and natural gas storage installations and arrived
at an overall estimate of 2996 plants for the US and Western Europe. From these data the
explosion frequency per plant was estimated as follows:

\[
\text{explosion frequency} = \frac{\text{number of explosions}}{(\text{time} \times \text{plants at risk})}
\]

\[
= \frac{19 + 9}{21 \times 2996}
\]

\[
= 4.5 \times 10^{-4} \text{ explosions/plant.year}
\]

This figure was then compared against estimates generated by the fire and explosion model.
12.4 Numerical analysis of fire and explosion model

The initial estimates of plant fire and explosion frequency using the model did not show basic agreement with the statistical data. It therefore proved necessary to iterate the fire and explosion model several times to achieve both intermediate and final results that were comparable with the statistical estimates given above. For the iteration procedure the original estimates of source leak frequency and the setpoints on Figure 11.2 for the explosion probability given ignition were altered. The iteration of these values was felt to be justified because the initial estimates were largely based on intuitive reasoning only, given the judgemental estimates listed in Sections 7 and 11. The source inventories, hole size distributions and setpoints on the ignition probability graph (Figure 11.1) were kept constant.

The original estimates of source leak frequency are shown in brackets alongside the final estimates in Table 12.2 where applicable. The degree of adjustment necessary for each figure may be expressed as the ratio of the original to the final estimate and are given below for each source:

<table>
<thead>
<tr>
<th>Source</th>
<th>Adjustment Ratio</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>2</td>
</tr>
<tr>
<td>Flanges</td>
<td>3</td>
</tr>
<tr>
<td>Valves</td>
<td>10</td>
</tr>
<tr>
<td>Pumps</td>
<td>1</td>
</tr>
<tr>
<td>Small-bore connections</td>
<td>5</td>
</tr>
</tbody>
</table>

Note that in most cases the final estimate of leak frequency was less than the original estimate, reflecting the conservative nature of the original estimates. Since the degree of adjustment was no more than a factor of 10 it was concluded that the original estimates were reasonably close to the final figures.

Regarding the explosion frequency estimation, it was found that the probability set points were set too high to give figures comparable to the statistical estimates. Consequently the lower set point as shown in Figure 11.2 was reset from 0.04 to 0.025, whilst the upper set point was reduced from 0.5 to 0.25. These new estimates were still in broad agreement with the explosion frequency statistical data presented in Tables 11.3 and 11.4. Iteration of the ignition probability
set points shown in Figure 11.1 was not necessary, as the discussion in Section 12.4.1 concludes that reasonable agreement with historical fire frequency data was found with the original estimates.

The complete output from the model program IGNITION with the above changes incorporated is given in Appendix G. Attention is drawn to the ignition event frequency summary breakdown by leak source given on page G22 and the summary breakdown by fluid phase given on page G26. These are referred to frequently in the following discussion.

12.4.1 Analysis of overall fire frequency estimate

The frequency of fires estimated from the model is 0.03 fires per plant\textperiodcentered year. The overall estimate was corrected for the fact that only a proportion of leak sources are included in the model. Redpath [110] gave data concerning large loss fires on process plant that categorised the fires according to the original plant item causing the fire. These are listed in Table 12.7 and show that of the 101 fires that were directly attributable to process equipment 58 of them, i.e. 57\%, may be taken as due to the sources considered here. Hence to convert the modelled estimate to an overall estimate it was necessary to multiply 0.03 by 101/58 to arrive at a final figure of 0.05 fires per plant\textperiodcentered year. This is at the upper limit of the statistical estimate of 0.025 - 0.05 fires/plant\textperiodcentered year given in Section 12.3.1. However there are two allowances for adjustment that were considered when comparing the figures.

1. The vessel/pipework ratio is probably higher for the plants considered by Redpath since fires due to tanks and vessels form a large fraction of the total. Also, since vessels and tanks tend to be "compound" leak sources i.e. they include flanges, nozzles etc. it is uncertain whether fires originating from vessel failures were instead due to the leakage of an ancillary component. This would tend to decrease the 101/58 correction factor given above.

2. Allowance should be made for the effect of fire protection measures which would keep the historical number of major fires low. The figures for fire loss given in Table 12.3 indicate that the majority of fires resulted in losses of less than $10^5. It is unknown how many of these fires would have been larger if protection measures were not taken.
It was surmised that if the above two conflicting factors were taken into account then factor 2 has a greater effect than factor 1. and so the model probably underestimates the overall fire frequency. More data would have to be available to arrive at a firm conclusion.

12.4.2 Analysis of intermediate fire frequency estimates

Not enough data were found to permit a detailed analysis of the fire frequency breakdown by leak source shown on page G22. It was possible however to condense the results into a form roughly comparable to the historical data presented in Table 12.7. For example, Table 12.7 indicates that about 62% of fires attributable to leaks were from pipe fracture. If "fracture" is assumed also to cover less severe pipe damage then the combined modelled contribution of pipes and small bore connections to the overall modelled fire frequency is 53%, which is similar to the statistical estimate. However, from Table 12.7 it is apparent that the contribution of flanges, pumps and valves to the overall fire frequency is low compared to the model estimate.

The other principal cross-check made was by fluid phase. Data given by the Fire Protection Association and reproduced by Lees [10] are given in Table 12.8. These concern a survey of large loss fires in Great Britain occurring between 1971 and 1973 which categorises each fire according to the material first ignited. The table shows that of 46 fires due to releases of gas, vapour or liquid, 10 were due to gas. Of these, 7 were due to hydrogen which were discounted from the analysis as not being typical within the "standard plant" approach. For other substances the fraction of fires due to gas releases is 3/39 i.e. 8%. This compares well with the modelled estimate of 10%, although it is uncertain how many of the vapour fires contribute to the gas total. The combined gas/vapour fraction from Table 12.8 is 19/39 or 48%, which is similar to the modelled gas/two-phase estimate of 59%.

12.4.3 Analysis of overall explosion frequency estimate

The modelled estimate of $6 \times 10^4$ explosions per plant.year was compared to explosion frequency data derived from Tables 12.5 and 12.6. It was again necessary to correct the model estimate for the leak sources that were not incorporated into the model. From Table 12.6 the number of explosions attributable to the modelled leak sources is 23 out of a total of 34 i.e. 68%. The corrected model estimate was therefore calculated as $9 \times 10^4$ explosions per plant.year,
compared to the statistical estimate of $4.5 \times 10^{-4}$ explosions per plant.year given in Section 12.3.2. Although the model estimate is high, it was thought that allowance should be made for the fact that 30% of the explosions are due to leaks below 1 kg.s$^{-1}$. At very low leak rates some of the predicted explosions may not be physically possible in ordinary outdoor well-ventilated conditions. It was also considered possible that small explosions such as those occurring in process unit casings, e.g. pumps, are under-reported. These two factors combined led to the conclusion that the modelled estimate is in broad agreement with actual conditions.

### 12.4.4 Analysis of intermediate explosion frequency estimates

It was possible to extract explosion data from Table 12.6 to compare directly with the intermediate results given by the fire and explosion model. This resulted in the following comparison:

<table>
<thead>
<tr>
<th>Leak source</th>
<th>Number of explosions</th>
<th>Percent</th>
<th>Model estimate Per cent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipe</td>
<td>10</td>
<td>43</td>
<td>63*</td>
</tr>
<tr>
<td>Flanges and fittings</td>
<td>8</td>
<td>35</td>
<td>19</td>
</tr>
<tr>
<td>Valves</td>
<td>4</td>
<td>17</td>
<td>12</td>
</tr>
<tr>
<td>Pumps</td>
<td>1</td>
<td>4</td>
<td>7</td>
</tr>
</tbody>
</table>

*this figure contains both pipe and small connection explosion frequencies.

The above comparison indicated there was some overlap between the numbers of explosions due to pipe and small connections and those due to flanges and fittings. The contributions of valves and pumps to the overall explosion frequency are similar for both the statistical and modelled estimates.

The second cross-check made was with respect to fluid phase. Of the 34 incidents given in Table 12.5 Davenport listed about 11 that involved substances that were gas or vapour before release. Of these, 4 involved hydrogen and 4 ethylene. For all other materials there were therefore 3 cases out of 27 i.e. 11%. This is rather high compared with the model estimate of 6.6%.
12.5 Conclusion

This section has presented a fire and explosion model which estimates the frequency per plant year of ignition incidents using data presented in earlier sections. The purpose of the model is to enable better estimates of leak frequencies and emission sizes to be made relative to each other by inputting original estimates of leak frequencies etc., comparing the resulting ignition frequencies against historical event data and iterating until reasonable agreement is achieved. Although the model contains only a percentage of the leak sources that have been listed as causing explosions it may be expanded as necessary. The model has been partly validated by two methods: firstly in terms of the overall and intermediate estimates of event frequency; and secondly by the degree of adjustment to the original estimates necessary to achieve results comparable to the historical statistics. Better validation will be achieved as more incident data become available. The application of the fire and explosion model in this section was confined to simplified data derived from leak source inventories given in Section 5. To investigate the use of the fire and explosion model further, Appendix I describes how the model was used in a case study, to analyse the leak source inventory data of an actual plant.

The plant used for the case study was the medium-sized hydrocarbon plant whose leak source inventory was described in Section 5 as Material take-off 1. There were several reasons for performing the case study presented in Appendix I. Firstly, it would give a measure of the ease with which an engineer presented with both the model and a plant leak source inventory in the form of a material take-off could estimate the likely leak and ignition frequency. Applying the model to an actual plant would also be expected to highlight unusual results compared to intuitive engineering judgement. Lastly, it would indicate which leak sources contribute the most to the estimated ignition frequency, and thus suggest how this frequency could be improved upon.

The case study of the hydrocarbon plant emphasised the relative crudeness of some elements of the model. For example, the inventory data and failure data are required as input to the model in the form of ordered lists according to component size. As the number of component sizes increases this method becomes increasingly cumbersome, pointing towards the need for improvement, such as for example the use of a spreadsheet-style input.

The results of the case study showed that there were no anomalies in the predicted ignition frequencies compared with the results obtained using the simplified "standard plant" input.
described in the first part of this section. The analysis given in Appendix I showed that the model may be used to highlight the effect of individual item inventory changes, and also changes in phase split within the inventory data.

In conclusion, the formulation of a collective approach to the estimation of ignition frequency by taking a sample of leak sources suggests a zoning procedure that is based on leak source and ignition source density in a given area, rather than the commonly perceived notion of taking each leak source separately that is indicated by present standards and codes advocating the source-of-hazard method of HAC. However, the formulation of this fire and explosion model is only applicable to outdoor plant. Section 13 outlines the special problems that are met in classifying hazardous indoor areas whilst Section 14 suggests further work that may continue from the point reached here. This includes the further refinement of the fire and explosion model, and discussion of possible ways in which the model may be used for HAC.
Table 12.1 Inventory of standard plant for use with the fire and explosion model

A. Overall inventory

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipe lengths (m)</td>
<td>3750.0</td>
<td>4200.0</td>
<td>5400.0</td>
<td>1650.0</td>
</tr>
<tr>
<td>Valves</td>
<td>720</td>
<td>500</td>
<td>240</td>
<td>40</td>
</tr>
<tr>
<td>Flanges</td>
<td>900</td>
<td>1070</td>
<td>810</td>
<td>220</td>
</tr>
<tr>
<td>Gas fraction</td>
<td>0.00</td>
<td>0.15</td>
<td>0.15</td>
<td>0.25</td>
</tr>
<tr>
<td>Liquid fraction</td>
<td>0.50</td>
<td>0.40</td>
<td>0.40</td>
<td>0.35</td>
</tr>
<tr>
<td>Two-phase fraction</td>
<td>0.50</td>
<td>0.45</td>
<td>0.45</td>
<td>0.40</td>
</tr>
<tr>
<td>Pumps</td>
<td>0</td>
<td>15</td>
<td>10</td>
<td>0</td>
</tr>
</tbody>
</table>

Small bore connections ($d = 0.01$ m) = 700

continued
Table 12.1 continued

<table>
<thead>
<tr>
<th>B. Inventory by phase</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Pipe diameter (m)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Pipe lengths (m)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>0.0</td>
<td>630.0</td>
<td>810.0</td>
<td>412.5</td>
</tr>
<tr>
<td>Liquid</td>
<td>1875.0</td>
<td>1680.0</td>
<td>2160.0</td>
<td>577.5</td>
</tr>
<tr>
<td>Two-phase</td>
<td>1875.0</td>
<td>1890.0</td>
<td>2430.0</td>
<td>660.0</td>
</tr>
<tr>
<td><strong>Valves</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>0</td>
<td>75</td>
<td>36</td>
<td>10</td>
</tr>
<tr>
<td>Liquid</td>
<td>360</td>
<td>200</td>
<td>96</td>
<td>14</td>
</tr>
<tr>
<td>Two-phase</td>
<td>360</td>
<td>225</td>
<td>108</td>
<td>16</td>
</tr>
<tr>
<td><strong>Flanges</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>0</td>
<td>161</td>
<td>122</td>
<td>55</td>
</tr>
<tr>
<td>Liquid</td>
<td>450</td>
<td>428</td>
<td>324</td>
<td>77</td>
</tr>
<tr>
<td>Two-phase</td>
<td>450</td>
<td>481</td>
<td>365</td>
<td>88</td>
</tr>
<tr>
<td><strong>Pumps</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Liquids</td>
<td>0</td>
<td>11</td>
<td>6</td>
<td>0</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0</td>
<td>4</td>
<td>4</td>
<td>0</td>
</tr>
<tr>
<td><strong>Small bore connections</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><em>(d = 0.01m)</em></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas</td>
<td>105</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Liquid</td>
<td>280</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Two-phase</td>
<td>315</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Table 12.2 Leak frequency profile of standard plant

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Pipe leak frequency (leaks/metre/year \times 10^4)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rupture leak</td>
<td>0.0050</td>
<td>0.0050</td>
<td>0.0015</td>
<td>0.0005</td>
</tr>
<tr>
<td></td>
<td>(0.0100)</td>
<td>(0.0100)</td>
<td>(0.0030)</td>
<td>(0.0010)</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.0500</td>
<td>0.0500</td>
<td>0.0150</td>
<td>0.0050</td>
</tr>
<tr>
<td></td>
<td>(0.1000)</td>
<td>(0.1000)</td>
<td>(0.0300)</td>
<td>(0.0300)</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.5000</td>
<td>0.5000</td>
<td>0.1500</td>
<td>0.0500</td>
</tr>
<tr>
<td></td>
<td>(1.0000)</td>
<td>(1.0000)</td>
<td>(0.3000)</td>
<td>(0.1000)</td>
</tr>
<tr>
<td><strong>Valve leak frequency (leaks/valve/year \times 10^4)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rupture leak</td>
<td>0.0100</td>
<td>0.0100</td>
<td>0.0100</td>
<td>0.0050</td>
</tr>
<tr>
<td></td>
<td>(0.1000)</td>
<td>(0.1000)</td>
<td>(0.1000)</td>
<td>(0.1000)</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.1000</td>
<td>0.1000</td>
<td>0.1000</td>
<td>0.0500</td>
</tr>
<tr>
<td></td>
<td>(1.0000)</td>
<td>(1.0000)</td>
<td>(1.0000)</td>
<td>(1.0000)</td>
</tr>
<tr>
<td>Minor leak</td>
<td>1.0000</td>
<td>1.0000</td>
<td>1.0000</td>
<td>0.5000</td>
</tr>
<tr>
<td></td>
<td>(10.0000)</td>
<td>(10.0000)</td>
<td>(10.0000)</td>
<td>(10.0000)</td>
</tr>
<tr>
<td><strong>Pump leak frequency (leaks/pump/year \times 10^4)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rupture leak</td>
<td>0.3000</td>
<td>0.3000</td>
<td>0.3000</td>
<td>0.3000</td>
</tr>
<tr>
<td>Major leak</td>
<td>3.0000</td>
<td>3.0000</td>
<td>3.0000</td>
<td>3.0000</td>
</tr>
<tr>
<td>Minor leak</td>
<td>30.0000</td>
<td>30.0000</td>
<td>30.0000</td>
<td>30.0000</td>
</tr>
<tr>
<td><strong>Flange leak frequency (leaks/flange/year \times 10^4)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Major leak</td>
<td>0.3000</td>
<td>0.3000</td>
<td>0.3000</td>
<td>0.3000</td>
</tr>
<tr>
<td></td>
<td>(1.0000)</td>
<td>(1.0000)</td>
<td>(1.0000)</td>
<td>(1.0000)</td>
</tr>
<tr>
<td>Minor leak</td>
<td>3.0000</td>
<td>3.0000</td>
<td>3.0000</td>
<td>3.0000</td>
</tr>
<tr>
<td></td>
<td>(10.0000)</td>
<td>(10.0000)</td>
<td>(10.0000)</td>
<td>(10.0000)</td>
</tr>
<tr>
<td><strong>Small bore connection leak frequency (leaks/connection/year \times 10^4)</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(d = 0.01\ m)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Major leak</td>
<td>1.0000</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>(5.0000)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Minor leak</td>
<td>10.0000</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>(50.0000)</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

The figures given above in brackets are the initial estimates derived from the literature survey data in Section 7. The final estimates are shown unbracketed.
Table 12.3 Number of refinery fires reported to the API broken down by cost
[104][105][106]

<table>
<thead>
<tr>
<th>Year</th>
<th>1982</th>
<th>1983</th>
<th>1984</th>
<th>1985</th>
</tr>
</thead>
<tbody>
<tr>
<td>Size of loss (US dollars x $10^3$)</td>
<td>No.</td>
<td>%</td>
<td>No.</td>
<td>%</td>
</tr>
<tr>
<td>2.5 - 1000</td>
<td>111</td>
<td>64</td>
<td>148</td>
<td>74</td>
</tr>
<tr>
<td>100 - 1000</td>
<td>51</td>
<td>30</td>
<td>45</td>
<td>22</td>
</tr>
<tr>
<td>&gt; 1000</td>
<td>11</td>
<td>6</td>
<td>8</td>
<td>4</td>
</tr>
<tr>
<td>Total</td>
<td>173</td>
<td></td>
<td>201</td>
<td></td>
</tr>
</tbody>
</table>

Table 12.4 Frequency of major fires from statistics in Table 12.3

<table>
<thead>
<tr>
<th>Year</th>
<th>1982</th>
<th>1984</th>
<th>1985</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major fires Refineries</td>
<td>62</td>
<td>50</td>
<td>44</td>
</tr>
<tr>
<td>240</td>
<td>202</td>
<td>191</td>
<td></td>
</tr>
<tr>
<td>Fire frequency (fires/refinery/year)</td>
<td>0.26</td>
<td>0.25</td>
<td>0.23</td>
</tr>
<tr>
<td>Fire frequency (fires/refinery unit/year)*</td>
<td>0.052</td>
<td>0.050</td>
<td>0.046</td>
</tr>
</tbody>
</table>

* assumes an average 5 units per refinery
Table 12.5 Leak sources and events leading to vapour cloud explosions (listed by Davenport [112]) for the years 1962 - 1982

<table>
<thead>
<tr>
<th>Incident code number</th>
<th>Country of origin</th>
<th>Flammable substance</th>
<th>Leak source</th>
</tr>
</thead>
<tbody>
<tr>
<td>76</td>
<td>US</td>
<td>Ethylene Oxide</td>
<td>Tank - reaction</td>
</tr>
<tr>
<td>138</td>
<td>Middle East</td>
<td>Propane</td>
<td>Tank (refrigerated storage) - overpressure</td>
</tr>
<tr>
<td>1</td>
<td>US</td>
<td>Ethylene/Methane</td>
<td>Possible sight glass failure</td>
</tr>
<tr>
<td>18</td>
<td>US</td>
<td>Ethylene</td>
<td>Pipe fitting</td>
</tr>
<tr>
<td>11</td>
<td>Belgium</td>
<td>Ethylene</td>
<td>Pipe connection</td>
</tr>
<tr>
<td>7</td>
<td>US</td>
<td>Ethylene/Methane</td>
<td>Flare header</td>
</tr>
<tr>
<td>4</td>
<td>US</td>
<td>Ethylene Chloride</td>
<td>Pipe rupture</td>
</tr>
<tr>
<td>62</td>
<td>FRG</td>
<td>Methane</td>
<td>Pipe rupture</td>
</tr>
<tr>
<td>17</td>
<td>US</td>
<td>Butadiene</td>
<td>Valve - opened on air failure</td>
</tr>
<tr>
<td>6</td>
<td>US</td>
<td>Isobutylene</td>
<td>Valve - bonnet failure</td>
</tr>
<tr>
<td>13</td>
<td>Holland</td>
<td>Light H/C &lt; C&lt;sub&gt;10&lt;/sub&gt;</td>
<td>Tank - frothover</td>
</tr>
<tr>
<td>40</td>
<td>UK</td>
<td>Naphtha/H&lt;sub&gt;2&lt;/sub&gt;</td>
<td>Pipe rupture</td>
</tr>
<tr>
<td>77</td>
<td>US</td>
<td>H/C &gt; C&lt;sub&gt;10&lt;/sub&gt; + H&lt;sub&gt;2&lt;/sub&gt;</td>
<td>Reactor - reaction</td>
</tr>
<tr>
<td>32</td>
<td>US</td>
<td>Ethylene</td>
<td>Pipe connection</td>
</tr>
<tr>
<td>22</td>
<td>US</td>
<td>Butadiene</td>
<td>Pump</td>
</tr>
<tr>
<td>49</td>
<td>Japan</td>
<td>Ethylene</td>
<td>Flange</td>
</tr>
<tr>
<td>52</td>
<td>Japan</td>
<td>Vinyl Chloride</td>
<td>Valve - yoke failure</td>
</tr>
<tr>
<td>68</td>
<td>US</td>
<td>Propane</td>
<td>Hose rupture</td>
</tr>
<tr>
<td>9</td>
<td>UK</td>
<td>Cyclohexane</td>
<td>Pipe rupture</td>
</tr>
<tr>
<td>57</td>
<td>UK</td>
<td>Ethylene</td>
<td>Flange (on thermowell)</td>
</tr>
<tr>
<td>8</td>
<td>US</td>
<td>H/C &gt; C&lt;sub&gt;5&lt;/sub&gt;</td>
<td>Expansion joint</td>
</tr>
<tr>
<td>118</td>
<td>Holland</td>
<td>Gasoline</td>
<td>Possible pipe failure</td>
</tr>
<tr>
<td>60</td>
<td>Holland</td>
<td>Propylene</td>
<td>Flare header rupture</td>
</tr>
<tr>
<td>70</td>
<td>FRG</td>
<td>Naphtha + H&lt;sub&gt;2&lt;/sub&gt;</td>
<td>Pipe rupture</td>
</tr>
<tr>
<td>74</td>
<td>US</td>
<td>H&lt;sub&gt;2&lt;/sub&gt;</td>
<td>Vessel - failure</td>
</tr>
<tr>
<td>51</td>
<td>US</td>
<td>Ethylene</td>
<td>Mixing nozzle</td>
</tr>
<tr>
<td>136</td>
<td>US</td>
<td>Propane</td>
<td>Pipe rupture</td>
</tr>
<tr>
<td>88</td>
<td>Qatar</td>
<td>NGL</td>
<td>Tank (refrigerated storage) - failure</td>
</tr>
<tr>
<td>93</td>
<td>Middle East</td>
<td>Methane/LPG</td>
<td>Pipe failure</td>
</tr>
<tr>
<td>94</td>
<td>US</td>
<td>Propane</td>
<td>Pipe failure</td>
</tr>
<tr>
<td>95</td>
<td>Romania</td>
<td>Propane/Propylene</td>
<td>Pipe failure</td>
</tr>
<tr>
<td>81</td>
<td>US</td>
<td>Propane</td>
<td>Pipe elbow</td>
</tr>
<tr>
<td>99</td>
<td>US</td>
<td>Hexane</td>
<td>Plug valve</td>
</tr>
<tr>
<td>141</td>
<td>US</td>
<td>Gasoline</td>
<td>Tank - overfilling</td>
</tr>
</tbody>
</table>
Table 12.6  Categorised leak sources in vapour cloud explosion incidents (listed by Davenport [112]) for 1962 - 1982

<table>
<thead>
<tr>
<th>Leak source</th>
<th>Number of events</th>
<th>Proportion (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tanks - reaction</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Tanks - overfilling, frothover</td>
<td>2</td>
<td>5.9</td>
</tr>
<tr>
<td>Tanks - failure of refrigerated storage</td>
<td>2</td>
<td>5.9</td>
</tr>
<tr>
<td>Vessel - pressure vessel rupture</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Reactor - reaction</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Pipes</td>
<td>10</td>
<td>29.4</td>
</tr>
<tr>
<td>Flanges</td>
<td>2</td>
<td>5.9</td>
</tr>
<tr>
<td>Other fittings</td>
<td>6</td>
<td>29.4</td>
</tr>
<tr>
<td>Hoses</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Valves</td>
<td>3</td>
<td>8.8</td>
</tr>
<tr>
<td>Sight glasses</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Pumps</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Valve opened</td>
<td>1</td>
<td>2.9</td>
</tr>
<tr>
<td>Flare</td>
<td>2</td>
<td>5.9</td>
</tr>
<tr>
<td>Total</td>
<td>34</td>
<td>100.0</td>
</tr>
</tbody>
</table>
Table 12.7 Large loss fires in process industry between 1963 and 1973 (given by Redpath [110])

<table>
<thead>
<tr>
<th>Item of origin</th>
<th>Number</th>
<th>Number caused by flammable emissions</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Leakage from:</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fractured pipe</td>
<td>36</td>
<td>36</td>
<td>35</td>
</tr>
<tr>
<td>Leaking gland, flange or coupling</td>
<td>10</td>
<td>10</td>
<td>10</td>
</tr>
<tr>
<td>Unspecified leakage</td>
<td>12</td>
<td>12</td>
<td>12</td>
</tr>
<tr>
<td>Vat, still, pot or mixer</td>
<td>22</td>
<td>22</td>
<td>22</td>
</tr>
<tr>
<td>Tank</td>
<td>21</td>
<td>21</td>
<td>21</td>
</tr>
<tr>
<td>Electrical wiring or equipment</td>
<td>12</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Other known</td>
<td>49</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Unknown</td>
<td>44</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>206</strong></td>
<td><strong>101</strong></td>
<td><strong>100</strong></td>
</tr>
</tbody>
</table>

169
Table 12.8 Large loss fires in process industry between 1971 and 1973 (given by Lees [10])

<table>
<thead>
<tr>
<th>Classification by phase:</th>
<th>Number of fires</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>10</td>
</tr>
<tr>
<td>Vapour</td>
<td>16</td>
</tr>
<tr>
<td>Liquid</td>
<td>20</td>
</tr>
<tr>
<td>Solid</td>
<td>23</td>
</tr>
<tr>
<td>Unknown</td>
<td>10</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>79</strong></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Classification by material:</th>
<th>Number of fires</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrocarbons</td>
<td></td>
</tr>
<tr>
<td>- Gas</td>
<td>3</td>
</tr>
<tr>
<td>- Liquid/Vapour</td>
<td>18</td>
</tr>
<tr>
<td>- Solid</td>
<td>2</td>
</tr>
<tr>
<td>Other organics etc.:</td>
<td></td>
</tr>
<tr>
<td>- Liquid/Vapour</td>
<td>16</td>
</tr>
<tr>
<td>- Solid</td>
<td>7</td>
</tr>
<tr>
<td>- Cellulosic solids (timber, paper, cardboard, fireboard)</td>
<td>6</td>
</tr>
<tr>
<td>hydrogen</td>
<td>7</td>
</tr>
<tr>
<td>steel</td>
<td>2</td>
</tr>
<tr>
<td>sulphur</td>
<td>1</td>
</tr>
<tr>
<td>unknown</td>
<td>17</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>79</strong></td>
</tr>
</tbody>
</table>
Figure 12.1 Schematic flowsheet of plant fire and explosion model program IGNITION
13. Indoor plants

The classification of indoor plants presents special problems. Firstly, whilst outdoor plants tend to be "closed" in that the flammable inventory is normally contained in pipes and vessels, indoor plants are more likely to have open liquid surfaces where a volatile liquid can readily evaporate into the ambient atmosphere. Secondly, because an indoor emission is not exposed to the open air a flammable atmosphere readily builds up inside the enclosure which is not easily modelled. The special considerations necessary for indoor plant are outlined in this section.

13.1 Ventilation of indoor plants

Unlike outdoor releases the amount of air necessary to dilute an indoor release cannot be assumed to be available. The airflow needed to dilute the release to a safe level must be determined and provided if the default solution of blanket zoning the whole enclosure is to be avoided.

The simple principles of ventilation of enclosed spaces in which a flammable atmosphere is likely to occur are set out by Leach and Bloomfield [113] and Harris [114]. The relevant British Standard for building ventilation is BS 5925 [115]. However, this is primarily concerned with meeting the ventilation requirements of people working in buildings rather than flammable release dilution. Ventilation may be either natural or forced draught. Natural draught ventilation occurs via two mechanisms: either by pressure differences between the inside and the outside of the enclosure or by temperature gradients within the enclosure. The former is effected by the wind, whilst the latter usually arises through the activities in the enclosure producing a warm air "chimney effect". Harris suggested that typical natural ventilation rates for an enclosure with closed doors and windows are 0.5 - 3 room volumes per hour (ventilation rates may be given as either air changes per hour or room volumes per hour. The current trend is to use room volumes because the former term implies a complete change of air inside the enclosure which is not necessarily the case).
Forced draught or mechanical ventilation may be either background or local to a specific emission source. It is more reliable than natural ventilation and can provide far higher rates. The amount of background ventilation that may be supplied to an enclosure is limitless, although if the enclosure is occupied then the ventilation rate becomes constrained by the maximum air velocity it is comfortable to work in. A typical value suggested by BS 5925 is 0.3 m.s\(^{-1}\). However, in an emergency such as an unplanned emission within the enclosure there is no reason why higher rates could not be used to dilute the release. Similarly, local ventilation e.g. an exhaust hood is specifically designed to dilute a potential release or remove it to vent. In enclosed premises natural ventilation is hardly ever relied upon to remove a flammable hazard. For hazardous area classification purposes mechanical ventilation is usually specified if zone sizes are to be limited.

13.1.1 Ventilation requirements

For a given source term the calculation of the amount of air needed to dilute the release to a specified concentration is straightforward. This is the case for high momentum releases such as gas jets where the ambient air flow does not affect the shape of the flammable envelope. However for low momentum releases the ventilation pattern and buoyancy of the release assume greater importance. Typically the ventilation pattern of an enclosure is from the bottom to the top. Under such conditions Harris defined the behaviour of negatively and positively buoyant gases as follows:

For a positively buoyant gas such as methane, Harris found experimentally that a layer of gas/air mixture of uniform concentration (i.e. perfectly mixed) is readily established between the leak source height and the roof. This is shown in Figure 13.1 where the mixture layer is termed the gas accumulation space (GAS). The layer concentration then increases gradually to a steady-state concentration given by Equation 13.1. This refutes the assumption made by Leach and Bloomfield that a highly concentrated layer of buoyant gas forms first at ceiling level and then gradually moves down to the leak source level.
For a negatively buoyant release the buoyancy and atmospheric forces are opposing. Harris noted that in this case the GAS occupies the whole enclosure with a gradually increasing release concentration. For both cases Harris modelled the concentration change in the GAS as:

\[ \frac{dC}{dt} = \frac{C_i - C}{\tau} \]  

which is the established equation for a single stage backmixed reactor where

\[ C_i = \frac{Q_a}{\dot{Q}} \]

\[ \dot{Q} = \dot{Q}_A + Q_a \]

\[ \tau = \frac{V}{\dot{Q}} \]

and where

- \( Q_a \) = source strength (m³.s⁻¹)
- \( Q_A \) = ventilating air flow rate (m³.s⁻¹)
- \( C_i \) = "effective" inlet concentration (vol. fraction)
- \( C \) = concentration within GAS
- \( V \) = enclosure volume (m³)
- \( t \) = time (s)
- \( \tau \) = time/throughput constant for enclosure

Thus for an initial release, the concentration build-up within the enclosure is given by:

\[ \frac{C}{C_i} = 1 - \exp\left(-\frac{t}{\tau}\right) \]  

On the cessation of the release the concentration decay is modelled by:

\[ \frac{C}{C_0} = \exp\left(-\frac{t}{\tau}\right) \]  

where \( C_0 \) is the initial concentration within the enclosure. Hence for a given source strength it is possible to predict the steady state concentration within the GAS. Alternatively it is possible to calculate the amount of ventilation necessary to prevent the GAS concentration from reaching a specified hazardous concentration, in other words dilution ventilation. If dilution ventilation cannot be achieved then invariably the procedure should be to classify the whole enclosure as hazardous.
13.1.2 Ventilation requirements of standards and codes of practice

The approach to indoor emissions taken by the standards and codes discussed in Section 2 is varied. Typically the assumption is made that if a source of release is located indoors then the entire enclosure is classified as hazardous unless there is "adequate" ventilation. However there is usually no guidance to what constitutes "adequate" ventilation so this distinction seems of little benefit. This approach is followed by both the Institute of Petroleum (IP) and American Petroleum Institute (API) codes of practice [116][117]. Exceptions to this general approach are the British Gas Engineering Standard PS/SHA1 [34] and the draft of the proposed new IP code [118].

British Gas PS/SHA1

The British Gas standard envisages two ventilation categories A and B. Category A ventilation is essentially dilution ventilation as described above with the specified "safe" concentration being $0.25 \times \text{LEL}$. Under such conditions hazard distances are calculated broadly according to the methods given in Sections 9 and 10 but with a hazard distance double the value to $0.5 \times \text{LEL}$ as an added safety factor. Category B ventilation does not completely dilute the release resulting in a blanket zone for such enclosures. A third category is labelled unventilated where no special arrangements are made to ventilate the enclosure.

Proposed Institute of Petroleum code

The proposed IP code envisages three designed ventilation conditions: adequate, dilution and overpressure ventilation. A fourth category, inadequate ventilation, is equivalent to the British Gas unventilated category. Adequate ventilation of an enclosure is defined as being sufficient to prevent the persistence and build-up of a flammable atmosphere, and is based on a rule-of-thumb ventilation rate of 12 air changes per hour. However, since the estimation method described by Equations 13.1 - 13.3 correlates ventilation with source strength this figure should be treated with caution.

Dilution ventilation is defined by the proposed IP code as immediately diluting a potential release to a safe level such that the enclosure may be declared non-hazardous except for the
immediate vicinity of the release. This is hence a stricter definition than the one used above which assumes the development of a plume or jet. The IP code intends however that such ventilation should only be applied to small enclosures such as analyser houses.

Overpressure ventilation is commonly applied to such enclosures as instrument cubicles or control rooms which contain ignition sources within an overall area containing release sources. The enclosure is kept at a positive pressure relative to ambient so preventing the ingress of flammable vapour into the enclosure. It is not directly relevant to this discussion since such enclosures do not themselves contain release sources.

The literature sources considered reinforce the statement made at the beginning of this section that dilution ventilation is necessary if blanket zoning of an enclosure is to be avoided. Having determined the required ventilation rate the hazard distances around each source may be calculated using methods given in Sections 9 and 10. The special problems attached to this are considered in Section 13.4. However, for a release source such as an LPG cylinder situated indoors the amount of dilution ventilation necessary may be prohibitive for reasons of cost and working environment. In such situations the source may be controlled to limit the release. This is considered further below.

13.2 Control of indoor sources of release

There is a considerably wider range of potential emission sources for indoor plant than for installations in the open air simply due to the wider range of specialised processes that are carried out indoors. The following examples are of indoor installations encountered in this study where HAC was deemed necessary. They are given here to illustrate the difficulty in assessing the release sources in a general fashion similar to outdoor releases and to show the necessity of employing some sort of control over the maximum credible release size.

**Indoor pressurised process plants.** Plants which would otherwise be constructed outdoors have on occasion been constructed indoors for reasons of cost and convenience. These include pilot plants. The emission sources are the same as for outdoor plants although for pilot plants there is an added emission risk due to the tendency to break open lines to make plant modifications.
Indoor processes carried out at ambient conditions. It is quite usual for mixing operations such as paint manufacture which involve handling flammable materials in a range of quantities to be carried out indoors. For processes such as these, potential emission sources arise prominently from material transfer in barrels and containers, from spillages during filling operations and also from the liquid surface in the mixing vessel if it is uncovered.

Turbine houses. Natural gas grid transmission is carried out by jet turbines located in compressor houses known as "cabs". These are unoccupied during operation and so are provided with a large amount of ventilation for cooling purposes as well as to remove any flammable releases. Emission sources include the high pressure gas pipe line and fittings within the cab and also the fuel gas and lubricating oil lines supplying the turbine.

Sewage processing. Methane produced from activated sludge treatment is pumped to indoor electrical generators via ducts and also to a site boiler for sludge heating. Potential emission sources are as for outdoor equipment although the very low pressures involved means that the flammable hazard results from gas accumulation over a period of time rather than from sudden large releases.

Case histories of accidents in indoor premises made available to the study included that of an aerosol filling operation involving LPG as the propellant where the dropping of an aerosol bottle resulted in a large leak from the filling nozzle. Accidents were also common where the vapour from flammable liquids used for cleaning was ignited. Injury and fatality statistics for these types of events are given in Appendix B. The emission sources for these events may be categorised in terms of activities rather than equipment type. For example, a filled barrel by itself is not a source of release, whereas the filling or transport of barrels by forklift truck is. This makes it very difficult to formulate a general approach for zoning indoors since it is apparent that, depending on its circumstances, each individual incident may have different consequences. It would seem more practical to identify the activities which may lead to a release and to minimise or exclude them. A procedure should be instituted for each separate situation whereby a hazard and operability (HAZOP)-style study is used to identify which emissions may be considered normal and planned for, and which are unplanned. Emissions which are planned for may be contained and dealt with by the background ventilation whilst unplanned-for emissions may require other measures which will be pointed out by the leak study. These may include the following:
1. Altering the design of the process to remove the leak activity completely e.g. replacing material transfer in barrels by piped supply.
2. Altering the enclosure design by removing the leak activity outside, or by segregating the activity within the enclosure and providing local ventilation.
3. Changes in working practices e.g. evacuation procedures.
4. Emergency shut down interlocks and alarms linked to gas detectors.
5. Emergency ventilation interlocked with the gas detection system.
6. Excess flow valves on supply pipework leading into the enclosure to limit the potential emission.

Assessing each indoor situation individually does not imply that specialised guidance notes for each industry in the manner of those listed in Appendix A are the way forward. The guidance notes already in use do provide a useful starting point for a specialised study but because they also set hazard distances they apparently assume that the enclosure shape and ambient conditions are of little importance. This is unhelpful as it removes the onus on the user to consider the behaviour of possible emissions and may result in an overall impression of "playing safe" which is quite wrong.

13.3 An outline approach to classifying hazardous indoor areas

So far the overall concepts of ventilation and control of emission sources in enclosed areas have been considered. These may be incorporated to form an outline for a starting point for analysing indoor releases which is discussed in this section.

If the primary objective of removing or excluding the release source cannot be achieved it becomes necessary to model the way a flammable atmosphere may build up within the enclosure. The suitability of the models already given in Section 9 for modelling indoor releases is considered in Section 13.4. For this study the concern was to construct an overall approach within which the models could be used. It was envisaged that such an approach should take into account the parameters both of the release and of the enclosure. These parameters and their effect on the behaviour of indoor releases are listed in Table 13.1. By considering the interaction of the release and enclosure parameters a series of six basic cases was constructed from which further more detailed analyses may proceed. The six cases are shown in Table 13.2.
13.3.1 Practical considerations to approach

The six cases formulated in this study outline the basis of a theoretical approach to assessing indoor releases. However, it was appreciated that the theory must be related to the practical aspects of each individual situation. These include:

**Size of enclosure.** Below a certain enclosure size the segregation of the enclosure into hazardous and non-hazardous zones is inappropriate. This is related to the emission and ventilation rate within the enclosure as discussed above.

**Fluid phase.** In Sections 10 and 11 it was noted that outdoor hazard distances and ignition risk vary more as a function of fluid phase than any other fluid property. This effect is emphasised for indoor locations where a small release of flashing liquid may rapidly fill the entire enclosure. In such instances it may be more prudent to automatically classify the entire enclosure as hazardous.

**Nature of plant/activities.** If there is a profusion of open liquid surfaces within the enclosure then blanket zoning may be the only option regardless of case category if the measures described in Section 13.2 for controlling emission sources are not adopted.

**Occupancy of enclosure.** Indoor plants are usually more populated than outdoor plant due to the nature of the activities being carried out. This is relevant because the presence of operators tends to give rise to emissions and the introduction of additional potential ignition sources, mainly as a result of operator errors. If the enclosure is occupied then it follows that there is the additional chance of injury and fatalities in the aftermath of an ignition, although balanced against this is the increased chance of leak discovery before a large flammable cloud develops.

It was thought that the application of the above factors to the six representative cases listed in Table 11.2 would form the basis of a structured approach to the classification of indoor hazardous areas beyond which development has not been pursued in this study. Finally, it was necessary to assess the suitability of dispersion models in estimating the plume size resulting from releases in well-ventilated areas.
13.4 Modelling indoor releases

The segregation of enclosures into hazardous and non-hazardous areas is only feasible if the dispersion of potential indoor releases can be satisfactorily modelled. For a given situation this may either be achieved using practical methods or by empirical calculations. The practical methods may include measuring the flammable concentration around equipment items and spills using an explosimeter, or alternatively assessing the ventilation of an enclosure using smoke tracer tests and air velocity measurements. This study was not concerned with the practical testing of flammable atmospheres although such a study would greatly benefit the formulation of a method for indoor HAC. Instead the suitability of applying the dispersion models given in Section 9 to indoor releases was considered.

Indoor releases may be modelled in an initially similar fashion to outdoor releases: for a leaking piece of equipment a hole size distribution may be estimated and the corresponding source strength determined. To estimate the dispersion according to the principles adopted for outdoor releases two criteria must be fulfilled: dilution ventilation and (for low momentum releases) predictable air conditions. Dilution ventilation is necessary since dispersion calculations assume entrainment of fresh air into the jet or plume. If this can be guaranteed then jet dispersion calculations may be applied as for outdoor releases. For Gaussian dispersion the second criterion of atmospheric predictability must also be met. This is difficult for conditions of low air speed as has already been touched on in Section 9.4.3.3. In their methodology manual [59] TNO stipulated a minimum airspeed of 1 m.s\(^{-1}\) for reliable dispersion estimates: however, as indicated above the maximum air speed for comfortable working conditions in occupied enclosures is approximately 0.3 m.s\(^{-1}\) whilst actual air speeds may be somewhat less. If Gaussian methods are to be used then the airspeed must be brought up to the specified minimum level which may be achieved with emergency ventilation. If this cannot be achieved then Gaussian methods are of little use and practical measurements of likely dispersion patterns should be made.

In conclusion, the need to assess each situation separately must be emphasised as reliable methods for estimating indoor dispersion were not found. It was suspected that in many processes involving open liquid surfaces the flammable zone over the liquid surface would be very small and that a hazard may only arise in unplanned-for situations e.g. a container being knocked over.
However, little evidence was discovered concerning this and it is suggested that the formation of flammable atmospheres from open liquid surfaces could form the basis of a more practically based study.

To summarise, this section considered the difficulties of applying the techniques so far described as being relevant to outdoor installations to indoor installations also. The ventilation requirements of indoor plants were discussed, both from a theoretical view and from the view taken by current HAC codes of practice. It was concluded that the only way by which non-hazardous areas could be created in enclosures containing emission sources would be if the ventilation were sufficient to dilute a potential release to a safe level. In view of this fairly stringent criterion it was felt that a more realistic way of reducing the indoor flammable hazard would be to institute a hazop-style procedure to identify and remove the potential emission sources themselves. Where this is not possible then a basic theoretical approach was presented which related the parameters of the release to those of the enclosure. Finally, the methods for modelling releases presented in Section 9 were reviewed for their suitability indoors, and it was concluded that their use was limited to situations where dilution ventilation could be guaranteed.
Table 13.1 Parameters to be considered for indoor releases

<table>
<thead>
<tr>
<th>Enclosure parameters</th>
<th>Effect on indoor release</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ventilation</td>
<td></td>
</tr>
<tr>
<td>- good</td>
<td>Good ventilation is sufficient to prevent the build-up of a flammable atmosphere (FA) outside a defined envelope, or flammable atmosphere zone (FAZ). This is equivalent to dilution ventilation. Poor ventilation cannot do this.</td>
</tr>
<tr>
<td>- poor</td>
<td></td>
</tr>
<tr>
<td>Restoration</td>
<td>Restoration is the time taken to restore a non-hazardous atmosphere within the FAZ after the cessation of the release and so depends on detection and shut-off times as well as ventilation rate.</td>
</tr>
<tr>
<td>- short period</td>
<td></td>
</tr>
<tr>
<td>- long period</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Release parameters</th>
<th>Effect on indoor release</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fluid density</td>
<td>For the normal pattern of upward ventilation a lighter-than-air release results in a GAS above the leak source. For a dense release the GAS fills the entire enclosure.</td>
</tr>
<tr>
<td>- lighter than air</td>
<td></td>
</tr>
<tr>
<td>- heavier than air</td>
<td></td>
</tr>
<tr>
<td>Strength</td>
<td>Source strength affects FAZ size for good ventilation. For poor ventilation $FAZ = FA =$ entire enclosure</td>
</tr>
<tr>
<td>Leak frequency</td>
<td>Affects whether or not the leak is credible and also therefore the classification of the enclosure as hazardous or non-hazardous.</td>
</tr>
<tr>
<td>Leak momentum</td>
<td>Decides which ventilation model should be used if the ventilation is good</td>
</tr>
<tr>
<td>Leak duration</td>
<td>Also decides which dispersion model should be used</td>
</tr>
</tbody>
</table>
Table 13.2 Discrete cases for considering indoor releases

<table>
<thead>
<tr>
<th>Case</th>
<th>Release/Ventilation characteristics</th>
<th>Dispersion model</th>
<th>Hazard time</th>
<th>FAZ</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>High momentum, all durations, good ventilation</td>
<td>Jet</td>
<td>Release duration</td>
<td>Jet envelope</td>
</tr>
<tr>
<td>2</td>
<td>High momentum, all durations, poor ventilation</td>
<td>n/a</td>
<td>From start of release to elimination of FA in GAS$^1$</td>
<td>GAS</td>
</tr>
<tr>
<td>3</td>
<td>Low momentum, instantaneous release, good ventilation</td>
<td>Gaussian puff$^2$</td>
<td>From start of release to extinction of puff</td>
<td>Puff envelope</td>
</tr>
<tr>
<td>4</td>
<td>Low momentum, instantaneous release, poor ventilation</td>
<td>n/a</td>
<td>From start of release to elimination of FA in GAS$^1$</td>
<td>GAS</td>
</tr>
<tr>
<td>5</td>
<td>Low momentum, continuous release, good ventilation</td>
<td>Gaussian plume</td>
<td>Release duration</td>
<td>Plume envelope</td>
</tr>
<tr>
<td>6</td>
<td>Low momentum, continuous release, poor ventilation</td>
<td>n/a</td>
<td>From start of release to elimination of FA in GAS$^1$</td>
<td>GAS</td>
</tr>
</tbody>
</table>

Footnotes:
1. This simplifies the actual case because the FAZ will initially be inside the jet/plume/puff envelope. It is some time before the release builds up to fill the entire enclosure.
2. Gaussian puff dispersion is determined in a similar fashion to plume dispersion (Equation (9.48)). TNO have given an equivalent expression for puff dispersion [59].

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Figure 13.1 Dispersion patterns of indoor releases with upward ventilation.
14. Future work

The foregoing work on the basis of HAC for outdoor and indoor plants represents a way forward towards a definitive quantitative method. At the outset of the project it was thought that such a method could be formulated using data and methodologies currently available in the open literature. However, the study has shown that in each of the areas considered e.g. component leak frequency, leak size, ventilation modelling, further work is necessary before a more formalised methodology can be produced of the kind that may form the basis of any future standard. A framework for a formalised methodology is proposed in this section with consideration given to the advancement of the ideas described in this study for them to be incorporated in the methodology. This is described below.

14.1 Refinement of the fire and explosion model

The fire and explosion model described in Section 12 estimates the ignition probability based on historical fire frequency data. The data in turn reflect the performance of the current mixture of HAC codes and guidance documents. It is the aim of modelling ignition event frequencies in this way to be able to predict the effect that changes in HAC practice have on the overall fire frequency. To do this it is necessary to establish the fraction of ignitions which may be prevented by the use of HAC measures and those which may not. This criterion forms the basis of a sub-model to the overall fire and explosion model which should take account of the following in estimating the relevant ignition probability:

1. Effect of fluid phase on ignition probability
2. The probability of ignition for an unprotected location should increase with leak flow.

(N.B. 1. and 2. are accounted for in the basic ignition probability model given in Figure 11.1)
3. There should be a finite probability of ignition which allows for the failure to achieve perfect protection within the hazardous area and which should also increase with leak flow rate.

4. There should be a finite probability of leak self-ignition (explained below).

5. Within the hazardous area the ignition probability is the sum of 3. and 4. whilst outside the hazardous area the ignition probability is the sum of 2. and 4.

6. The overall ignition probability and also the fraction of self-ignitions should then agree with statistical evidence.

Self-ignition is the term used here to denote ignition at the point of release. In other words it means distance-independent ignitions such as may result from releases above the auto-ignition temperature; static electricity; or sparks and hot distressed surfaces occurring as a result of the release. Such ignitions would occur regardless of ignition protection measures taken as a result of HAC and do not therefore include ignitions by electrical equipment or hot surfaces situated around the leak source.

Points 1. to 6. are visualised in Figure 14.1 which is based on the gas/vapour ignition probability line given in Figure 11.1. It is suspected that at low flow rates the ignition probability is dominated by self-ignition incidents, although as the leak size grows larger the number ignitions due to imperfect zoning precautions become significant. For very large releases zoning precautions have a negligible effect on the overall ignition probability and line C. may approach line D.

This speculative model highlights two points where further work is necessary to clarify the nature of ignition incidents. Firstly it is necessary to determine the frequency of self-ignition incidents within the overall fire frequency. This may be achieved by analysing event data with regard to ignition sources. Such data have been summarily presented in Appendix B where the full case history for each incident may yield the required data. Since the data system described in Appendix B was instituted in 1986-87 (i.e. during the course of this study) not enough data were available to make a detailed analysis; however as more data become available this may become more feasible.

The second point highlighted by Figure 14.1 is that to reduce the overall ignition probability it is necessary to reduce the overall ignition source density within the zone (which would lead
to a decrease in the slope of line B) as well as increasing the distance between the leak and ignition sources. This is discussed further below. To conclude this section, it must also be emphasised that a crucial aspect of refining the fire and explosion model is the gathering of more data concerning leak frequency and leak size distribution to provide better validation of the final event frequency estimates.

14.2 Application of the fire and explosion model to HAC

It is envisaged that the fire and explosion model may be used in conjunction with a computer-based plant layout design package to investigate the effect of changing zone size. By taking a section of plant it should be possible to quantify the effect on the fire frequency produced by adopting different zoning policies i.e. the codes and guidance notes given in Appendix A. These could be incorporated by superimposing the zone distances on the plant layout with respect to the leak source locations. The ignition source density within each zone could then be adjusted to reflect actual conditions, in other words the fact that perfect protection cannot be achieved within the zone due to, for example, faulty equipment or the chance disregard shown to the HAC standard used. The ignition source density outside the zone would reflect typical unprotected conditions. The layout would then be tied in to the estimates of release size and frequency to produce a fire frequency which would depend on the release size and therefore the number of ignition sources in the ensuing flammable envelope. As well as fire frequency, the number of injuries and the amount of damage could be modelled depending upon the occupancy of the plant layout section concerned. A similar task is performed for entire sites by the IFAL (Instantaneous Fractional Annual Loss) model [119] although the fire and explosion model would be used on a very much smaller localised basis.

Such an approach would be expected to emphasise that the ignition of a leak from a given source is not only a function of the envelope size around the source but also of the combined envelopes around the leak sources in the immediate vicinity. Also emphasised would be the importance of the ignition source density as mentioned above. If the variation in zone sizes can be quantified with respect to a given level of ignition risk then this may lead to a method of setting areas of high and low ignition risk which would approximate to the current practice of setting Zone 1 and Zone 2 distances: however the modified zone distances would have a sound quantificational basis whereas the present fixed zone distances as expressed in Figures 2.1 - 2.3 do not. A possible approach to zone definition is discussed below.
14.3 Zone definition

So far the study has not directly addressed the problem of ranking areas according to the level of flammable risk present. The study has progressed through the estimation of source strength, the frequency of that source strength, the size of the flammable envelope around the source within which an ignition risk exists and, based on statistical evidence for a section of plant, the probability that the release will ignite. The question is on what basis shall the sections of plant be ranked as having a high or low ignition risk or, more fundamentally, a negligible ignition risk?

Currently the practice is to place a zone around a particular leak source and categorise the zone as either 0, 1 or 2 according to the likelihood of a flammable atmosphere occurring within the zone. A common progression from this point which is used by the ICI/RoSPA guide and the British Gas Engineering Standard PS/SHA1 (see Appendix A) is to describe each zone in terms of the number of hours per year that a flammable atmosphere may be expected to exist within it. The basis for this method is described by Parvin [120] as being based on a range of factors including the FAR for an operator in a hazardous area, the release frequency and duration and the sparking frequency of the electrical ignition sources in the vicinity of the leak source. The method is summarised in Appendix H whilst the results may be condensed as follows:

<table>
<thead>
<tr>
<th>Zone</th>
<th>Time of occupancy for flammable atmosphere per year</th>
</tr>
</thead>
<tbody>
<tr>
<td>Zone 0</td>
<td>&gt; 1000 hours</td>
</tr>
<tr>
<td>Zone 1</td>
<td>10 - 1000 hours</td>
</tr>
<tr>
<td>Zone 2</td>
<td>1 - 10 hours</td>
</tr>
<tr>
<td>Non-hazardous</td>
<td>&lt; 1 hour</td>
</tr>
</tbody>
</table>

The zone definitions were primarily intended to aid the installation of electrical equipment. For other ignition sources there is just one zone whose size is dependent on the size of release and which corresponds to Zone 2. Further work should certainly concentrate on refining the definition of the above zones with respect to all kinds of ignition risk and not just the risk of ignition from electrical equipment as is the case above. This has partly been accounted for in the formulation of the fire and explosion model. Whilst it seems clear that zone definition should be based on the overall ignition risk for the plant section concerned it is not certain which criteria should be used to define the cut-off level between hazardous and non-hazardous areas. Whilst the FAR
has been used in the past and is advocated in Section 4 as a measure of the effectiveness of HAC. It is difficult to use as a predictive tool because it can only be derived from existing data. It is probable that future work will investigate how the techniques of quantitative risk assessment (QRA) may be applied to arrive at levels of individual risk, which have already been touched upon in Section 4.1. It will be necessary to obtain improved estimates of event frequency and the event consequences such as risk from thermal radiation and explosion overpressures to arrive at an estimated risk of death to an operator working in a hazardous area. This may then be compared to levels of tolerable risk such as have already been identified by the Health and Safety Executive [121]. However, the modelling of such risk, exemplified by the Health and Safety Executive's computerised risk assessment method RISKAT [122], is chiefly concerned at present with levels of off-site risk i.e. the risk to the general public. For HAC purposes it is necessary to establish the individual risk to site personnel, and possibly process plant. The basic approach adopted by this work precluded such detailed study, especially since the crude estimate of the FAR due to ignition incidents given in Section 4 showed that there was a significant risk to plant operators without the need for such rigorous analysis. It was concluded that the use of QRA methods for defining hazardous areas will assume greater importance only after they have been refined to deal with ignition incidents of a less serious nature than complete vessel rupture, and when uncertainties in estimating the expected frequency of such incidents have been reduced.

Another idea that could be explored further with the aid of the proposed plant layout/fire and explosion model package is the reduction of calculated zone sizes in situations where the strict application of a zoning standard creates extreme operational difficulty e.g. when a zone overlaps a road. By varying the ignition source density within the calculated zone to account for the increased ignition probability incurred by reducing the zone size it should be possible to estimate the incremental increase in risk. Should this be small compared to the overall plant risk then an argument may be made for relaxing the zoning standard. However, such an approach may prove very difficult to regulate in practice and so the reduction of quantified zone distances should be approached with extreme caution.

14.4 Indoor plants

It has already been stated in Section 13 that for the zoning of an indoor plant to be put on the same quantitative basis as outdoor plants then the same degree of predictability must be established for the dispersion of the release in the ambient surroundings. For volatile liquids the
rate of evaporation is also rather unpredictable in conditions of low windspeed. Whilst the behaviour of indoor releases may be qualitatively described, the quantification of such behaviour requires a more practically based study of small releases into indoor areas to establish either new correlations or the modifications necessary to existing ones. The control of emission sources was also considered in Section 13 where it was proposed that a HAZOP-style procedure be instituted to limit the occurrence of open liquid surfaces. Further work is required to expand this approach and to formalise the basic rules for such a procedure.

14.5 Offshore plants

The overall approach to quantitative HAC so far described is intended to be applicable to any area of process plant where leak and ignition sources are present together. However, it is recognised that some types of plant pose more of a problem in the application than others. Offshore oil and gas production platforms in particular present a special problem because although a particular platform is very much out in the open a large proportion of the plant is totally or partly enclosed in modules. Also, although the platform is out in the open its compactness means that it may be relatively poorly ventilated. However, it has been shown in Section 11 that there are some similarities in ignition frequencies between onshore and offshore plant. Despite these similarities, and in view of the fact that the results so far discussed are provisional to further data-gathering and also in view of the suspected plant layout complications, the problems of dealing with offshore plant are thus noted but not pursued.

14.6 Minimum Inventory

A problem identified at the outset of the study was the specifying of the minimum inventory of flammable material present in a section of process plant below which the ignition hazard is negligible and the usual techniques of HAC are inapplicable. Equipment containing inventories below the minimum is most likely to present a problem indoors and so the concept of minimum inventory is taken to be an aspect of indoor plant.

The problem is not a recent one. There have been several attempts published in the literature to define the inventory level below which zoning precautions may be relaxed. The most rudimentary treatment is given by BS 5345: Part 2 [5] which stated that the source of hazard
approach to HAC may not be applicable where the total quantity of flammable substance available for release is less than 50 litres. However it was not specified whether the substance is gaseous, or is flashing or non-flashing liquid. As has been shown in Sections 10 and 11, fluid phase has a considerable effect on zone size and ignition probability and so this treatment may be considered inadequate.

The Dutch and Italian standards listed in Appendix A considered minimum inventories in more detail. Here the approach taken to inventories of flammable gases was to calculate the pressure-volume product at a given reference temperature and then compare the result with tabulated figures. Below a certain limit the relevant HAC procedure was deemed unnecessary. A similar procedure was followed for inventories of flammable liquid where the volume and the flashpoint were compared with tabulated figures. This procedure was only applicable to outdoor situations with sufficient ventilation to dilute the release to a safe level.

The most recent approach to considering levels of minimum inventory is that of the European standard making body CENELEC whose most recent draft procedure for HAC [123] eschews the direct quantification of inventory levels. Instead the procedure states that it is instead necessary for the operator to decide whether the equipment or plant under scrutiny is capable of releasing sufficient material to form a "dangerous volume" of flammable atmosphere and thus assess the applicability of HAC methods. This approach seems more satisfactory than either of the methods described above as it is intended to account for the manner in which the inventory may be released. For example, it is clear from the number of accidents listed in Appendix B concerning small containers that small amounts of flammable fluids readily give rise to ignition incidents when they are spilled, and it can be appreciated that under certain conditions there is no fixed inventory limit applicable to all situations below which a flammable atmosphere will not exist.

A practical procedure for relaxing zone distances where inventories are small may first entail the careful definition of the conditions necessary for this to be possible. The most important of these would probably be the fluid phase upon release. For example, for limited zoning the following conditions may apply:
The substance is not pressurised
The substance is not a flashing liquid
There should be some sort of containment to limit the release (e.g. a bund)
Dilution ventilation should be guaranteed

Further work on defining minimum inventories should emphasise the necessity of considering the ambient surroundings within which the release is likely to occur rather than just the inventory size.

To summarise, this section has outlined areas of further work which need to be addressed before a quantitative methodology for HAC can be established. These include the further analysis of ignition incidents, in particular the fraction of incidents which HAC precautions cannot prevent. The problem of which risk criteria may be used to rank hazardous areas had to be deferred, although it was argued that the use of the FAR as a crude measure of flammable risk was probably sufficient considering the current amount of uncertainty in the data so far presented. An outline method for using the fire and explosion model to classify hazardous areas in conjunction with a computer-based plant layout package was proposed. It was noted that both indoor and offshore plants require special consideration if they are not to be subjected to blanket zoning. Finally, it was concluded that there can be no minimum flammable inventory that is applicable to all situations below which there is a negligible risk of ignition.
Line A = Ignition probability within the hazardous area due to the ignition protection not being wholly successful.

Line B = Leak self-ignition probability

Line C = Sum of lines A and B and hence the overall estimate of ignition probability.

Line D = Ignition probability with no ignition protection measures taken.

Figure 14.1 Diagrammatic representation of ignition characteristic model
15. Conclusions and recommendations

A study was carried out which investigated the feasibility of putting hazardous area classification (HAC) on a quantitative basis. Up to the present time HAC has been carried out by largely qualitative methods; by using expert judgement to establish a zone around a leak source in which ignition sources are not allowed. This has led to a profusion of methodologies which are not always based on the same criteria. By returning to first principles for the problem it was hoped to produce more rigorous and uncontentious guidelines which would form the basis of future codes of practice.

In the end it was found not possible to achieve this ambitious goal. What has been achieved is the identification of one type of overall approach which accounts for all the factors which may be expected to contribute to a quantitative method for HAC. Where necessary, choices were made concerning leak source characteristics and release modelling that were subjective and although sometimes crude in nature, as explained below, a great deal of effort was put into devising a series of cross-checks for each step in the modelling process. However, the modular nature of the approach designed in this study means that where subjective decisions were made it is possible to substitute more refined alternatives without affecting the overall strategy.

15.1 Background

The development of HAC was reviewed from its origin in the oil drilling industry to the proliferation of guidance and codes of practice available to cater for most types of process industries where flammable substances are handled. The approach taken by the majority of national and international codes of practice is the setting of fixed distances around typical items of process equipment where flammable releases are anticipated. It was shown that there is sometimes a wide discrepancy in the distances given by different codes for the same item. The conclusion was drawn that this identified a need for a standardised treatment to model the behaviour of flammable releases.
15.2 Risk criteria for hazardous area classification

The risk to workers and operators from the ignition of flammable releases was reviewed using accident data from a database recently established during the course of this study. Consequently the available data were somewhat sparse although it was concluded that the risk to workers from both "closed" i.e. pressurised plant and plant consisting of open liquid surfaces was significant. Accordingly it was proposed that the fatal accident rate (FAR) would be a suitable criterion with which to measure the risk presented by different ignition hazards. It was noted that the FAR was particularly applicable since it applied to personnel at most risk from the hazard concerned. Whilst determining the FAR using accident data is one method of determining individual risk to plant operators from flammable ignitions, another method that may be used is the modelling of risk using consequence models i.e. quantitative risk assessment. This is current practice for assessing the risk to members of the general public from a planned industrial installation prior to its construction. However, it was felt that existing QRA methods were not refined enough to model the risk from smaller scale ignition incidents that may be prevented by using HAC methods, although this situation may change as new QRA methods are developed. Notwithstanding the need for future work, the use of a risk criterion based on property damage was discarded for the present as it was envisaged that it would eventually conflict with human risk criteria.

15.3 Assimilation of data necessary for a quantitative approach

The approach taken to assessing the ignition risk on process plant was firstly to consider outdoor plant. This selection was made because it was judged that outdoor process plants were mostly similar in type, construction and the hazards presented. It was felt that they were therefore easier to typify than indoor plants which are considerably more diverse in characteristics.

Basic information relevant to an envisaged quantitative approach was researched and assembled. This included:

- Inventories of plants at risk from fire and explosion
- Inventories of leak sources on each plant
- Estimates of leak frequency
- Leak hole size distributions from leak frequency data
Mathematical models for emission, vaporisation and dispersion of leaks
Representative physical property data required to operate the above models
Basic ignition and explosion frequency and probability information

The data were of variable quality. How this was dealt with is described below. The main thrust of the work was to develop the above data sources into a basic coherent approach for quantitative HAC, whilst bearing in mind the eventual need for a rigorous design method based on the results of the work.

The assimilation of the data was performed along two strands of work. Firstly, the estimates of hole size distributions were matched with the mathematical models for emission and dispersion to produce hazard distances. These were then used to investigate the effects of equipment item, size and fluid properties on the calculated distances. It was concluded that the ranges produced were broadly in line with the more conservative of the existing "fixed distance" zoning methods. The significance of flashing liquid leaks and spray leaks were highlighted as these resulted in large dispersion distances because of the high emission flow. However, the fact that the hole size distribution data was scarce meant that there was very little information on which to draw firm conclusions concerning the frequency of each release size. This led on the second strand of the work described below.

Basic risk assessment techniques were applied to the available event frequency data to determine the absolute risk of ignition presented by the calculated releases. The ideal solution would have entailed the use of the data listed above with associated confidence limits to arrive at the most likely ignition risk associated with a particular hazard distance. However, despite considerable effort directed towards the collection of ignition frequency and leak size and frequency data it was concluded that the data obtained were not of sufficient quality to permit such a rigorous approach to be taken.

15.4 Fire and explosion model

The alternative approach adopted was firstly to derive a typical "standard plant" using leak source inventory data and typical process conditions. This was then used to study the interaction of the basic information types listed above and their contribution to the overall plant fire and explosion frequency. A model was therefore devised for this task. The inputs were the
standard plant; the leak frequency and size data; and ignition and explosion probabilities based on statistical information and judgement estimates. The outputs were the frequencies of fire and explosion broken down by equipment type and fluid phase. The computed figures were then compared against statistical data so that the input data could be iterated to give better agreement. It was concluded that partial validation of the model was achieved because of the relatively small degree of adjustment needed to the original best-guess input data to arrive at intermediate figures which were comparable to historical data. The fire and explosion model was therefore judged to be an aid to obtaining improved estimates of the input frequency and probability data. The formulation of this model provided the culmination of the present work.

Further work using the basic fire and explosion model was envisaged to consist of the refinement of the overall ignition frequency to estimate which ignitions may be prevented by HAC precautions and which may not. The ignition source for the latter events would occur concurrently to the leak itself e.g. a hot bearing resulting in a leaking seal. Such events were termed "self-ignitions" and it is thought that small leaks are particularly prone to ignition in this way. Exploration of this phenomenon would require the gathering of incident data. Indeed a vital part of future work would be the gathering of further statistical data of all sorts used in this work to reinforce the data given here.

It was further envisaged that the refined fire and explosion model may be coupled with a computer-based plant layout package. This could then be used to investigate the effect of different zoning policies on the overall ignition frequency. The inputs would be the layout of the standard plant, the equipment leak frequencies and hole sizes, and the hazard distances produced by the calculational models. The zones defined by the policy under review would also be an input. Estimates would then be made of the distribution of ignition sources within and around the zones which could be iterated to give the historical ignition and explosion frequencies associated with the use of the particular zoning policy. Having modelled the current situation, the effect of changing zoning policies could then be explored. Such an approach would lead to revised definitions of Zone 1 and Zone 2 areas, which are currently viewed as being relevant only to the installation of electrical equipment.

It was concluded from the formulation of the fire and explosion model that whereas the present source-of-hazard philosophy deals with leak sources in isolation, a better measure of the fire risk on process plant is obtained by considering the totality of leak sources. This is more
a refinement of the blanket zoning technique. The uncertainties in the data used to formulate the model preclude any major extension of the source-of-hazard technique until more reliable data are gathered.

To conclude for outdoor plants, two interim proposals were made subject to the future progress of this work. Firstly, it was proposed that setting zone distances should take particular account of the phase of the release, since flashing liquids present an extraordinary hazard. Secondly, the notion of limiting zone distances should be considered in special cases where the strict adherence to fixed distances on a plant leads to severe operational difficulties. This would be based on the principle that the incremental increase in risk involved in decreasing the zone distance may be negligible compared to the overall fire risk presented by the plant.

15.5 Indoor plants

The problems of HAC of indoor plant were also considered. It was concluded that whereas for outdoor plant the leak sources are relatively easy to identify, indoor releases may occur in quite diverse ways which are mainly as a result of human activity. It was proposed that work should be directed at developing a HAZOP-style hazard identification method for controlling leak and ignition sources rather than simply setting hazard distances as for outdoor plant. Such a procedure may also include the setting of a minimum inventory of flammable substance for a particular situation below which a hazard is judged exceedingly unlikely. Where the setting of zone distances was necessary it was surmised that to be able to model indoor releases in a similar fashion to outdoor releases then predictable air conditions were required. In particular this entailed sufficient ventilation being available to dilute the release. It was realised that these were stringent conditions that were seldom encountered in actual situations. It was therefore proposed that a more practical study of the dispersion of indoor releases should be conducted to provide a firm basis for zoning guidelines. Notwithstanding the need for future work an attempt was made to characterise the range of different general emission scenarios. It was thought that these could then be used as a basis to define sets of assumptions from which to produce a modelling approach for indoor plant HAC.
15.6 Summary of recommendations

To summarise, the following are areas of work which need to be addressed before a rigorous method for HAC may be formulated.

1. The FAR criterion suggested here should be compared with others, and the selected criterion should be subsequently refined into a form applicable to HAC. This would probably entail the fixing of a cut-off point to distinguish hazardous from non-hazardous areas.

2. The fire and explosion model requires further refinement. In particular a study of leak self-ignition frequency is required. More event frequency data are required to enable better validation of the refined model.

3. A suitable plant layout design package must be located to which the fire and explosion model may be applied. This being done, the coupled model may then be used to investigate the variation of zoning policies and the possible relaxation of zone distances.

For indoor plant the following work is necessary:

4. A hazard identification procedure should be developed for the control of emission sources. This should also include the assessment of the minimum amount of flammable substance required to pose a hazard in a particular situation.

5. The dispersion of small releases indoors should be studied.

6. Further characterisation of indoor or enclosed situations is required where HAC may be applicable. These may include such locations as offshore plant modules.
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Appendix B. Incident and injury rates


Appendix C. The $\chi^2$ confidence test


Appendix E. Physical property data for use with hazardous area models


Appendix A. Relevant standards, codes of practice and guidance notes for hazardous area classification

A.1 UK National and industry documents

1. BS 5345: Code of practice for the selection, installation and maintenance of electrical apparatus for use in potentially explosive atmospheres (other than mining applications or explosive processing and manufacture). Part 2: Classification of hazardous areas. British Standards Institute (1983)


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Appendix B. Incident and injury rates

B.1 Fire and explosion study

An analysis was made of accident data collected by the Health and Safety Executive (HSE). The data formed a record of incidents involving the ignition of flammable atmospheres which were notifiable to the HSE under the Reporting of Injuries, Diseases and Dangerous Occurrences Regulations (RIDDOR) 1985 [124] during the period April 1986 to March 1987. Briefly, RIDDOR requires that the HSE be notified of an accident at the workplace if it results in injury or death. An analysis of such accidents therefore provides a measure of the effectiveness of HAC codes using the fatal accident rate as a risk criterion.

Table B.1 summarises the number of incidents, injuries and deaths occurring as a result of the ignition of a flammable substance during the period under review. Although they all occurred as a result of some sort of ignition they are not all relevant to hazardous area classification. The irrelevant incidents which were excluded from the analysis involved:

- Catering and baking
- Disposal of unwanted material by burning, crushing etc.
- Electricity generation, distribution and storage
- Transport of flammable material (e.g. by road or rail)
- Incinerators, furnaces, ovens and kilns
- Spontaneous ignitions, exothermic reaction runaways
- Lightning
- Flammable solids and dusts

Omission of these reduced the number of incidents from 968 to 335. Of these, 110 incidents were listed as being due to hotwork e.g. welding either as the source of ignition or as the process going on at the time of ignition. It was unclear in these cases whether hotwork was the cause of both the initial leak and its subsequent ignition or whether the leak occurred elsewhere and was
then ignited by the hotwork. It was assumed here that in all cases hotwork was the cause of the initial leak and therefore these cases were omitted from the analysis. Support for this assumption can be found in work by Forsth [96] concerning ignition incidents on offshore platforms in the Norwegian North Sea. He cited 80 incidents involving hotwork as the ignition source and in all cases the hotwork was also the initiating event for the flammable release.

Omission of incidents involving hotwork leaves a total of 225. These were then divided into two categories. The first group involved leaks from "closed" i.e. usually pressurised process plant such as vessels and pumps, whilst the second group involved incidents concerning open liquid surfaces which were probably due to spills, or at worst the uncontrolled emission of a spray. Dividing the incidents in this way enabled a better comparison with published data.

Tables B.2 and B.3 show the number of incidents for the two categories versus the origin of the fire or explosion as listed in the HSE data. Table B.4 then details injuries and fatalities resulting from the closed process plant incidents in Table B.2 whilst Table B.5 gives similar details for the open surface incidents in Table B.3.

B.2 Other employment/accident statistics

Table B.6 Sections A and B give further statistics provided by the HSE concerning fatalities and major injuries in the manufacturing industries during the period 1981-1985. These were reported according to the Notification of Accidents and Dangerous Occurrences Regulations (NADOR) and so were categorised slightly differently to those reported under RIDDOR. Another set of related statistics was published in the 1984 Annual Report of HM Chief Inspector of Factories. These were collected by the Chemical Industry National Group of the HSE [125] and are reproduced in Table B.6 Section C.

Using the data contained in Tables B.1 - B.6 an estimate was made of the fatal accident rate for ignitions on closed process plant. This type of plant was selected for analysis because published data were available to check the results, although a similar analysis may be done for plant involving open surfaces.

The fire and explosion incident survey for 1986-87 described in Section B.1 listed one fatality on closed process plant due to the ignition of a flammable leak. Further statistics made available
to this study by the Health and Safety Executive showed that for the chemical and oil industry between 1981 and 1987 inclusive there were six fatalities attributable to ignition of flammable fluids.

From brief descriptions of the incidents provided with the statistics it was possible to ascertain that the six fatalities occurred in five separate incidents, three of which were during maintenance work in refineries. The fourth incident occurred during switch loading of a tanker, where the ignition source was attributed to static electricity, and the fifth and last incident was as a result of a release from a condenser where the cooling had failed.

These data were crosschecked against the NADOR statistics for 1981 to 1985 given in Table B.6. The numbers of fatalities from all causes for these years were 11, 9, 11, 6 and 6 respectively. Also, the same set of statistics yielded the fact that out of 4743 injuries in manufacturing industries reported to the HSE in 1985 the number of injuries due to fire and explosion was 89, or about 2 percent. Applying this figure to the fatalities results in basic estimates of 0.12 - 0.22 fatalities per year due to fire and explosion. The estimates are particularly crude for these years since the figures include all types of incidents resulting in fire and explosion and not just leaks. Despite these considerations, this cursory check supports the conclusion made from the 1981-1987 HSE statistics given above that deaths due to ignition of flammable leaks are six in seven years i.e. about 0.9 per year.

B.3 Number of persons exposed to risk

An estimate of the number of employees in the manufacturing industries who are exposed to risk from flammable ignitions is necessary before an estimate of the FAR can be made. For closed process plant this can be narrowed down to those personnel working in the petrochemical and related industries. It was assumed that the bulk of the flammable risk in these industries is posed by closed plant of the type described in Table B.2 - B.3. Table B.6C gives summary details of these employment figures, which may be further categorised (to the nearest thousand):

B3
However, not all the employees who work in an industry are exposed to flammable risk. A tentative estimate was therefore required of the proportion of employees who are exposed. For the chemical industry this was assumed to be one quarter, whilst for the oil industry it was assumed to be one third. Justification for differentiating between the two industries came from the different injury rates listed in Table B.6C. There seemed no reason why the injury rate for oil processing and chemical processing should be significantly different except by reasoning that the proportion of people exposed is greater for the former. Assuming that there is negligible risk of flammable leak ignition in the soap industry the number of persons exposed was then estimated thus:

Chemical Industry  =  (249 000 + 38 000)/4  
= 71 750

Oil industry  =  18 000/3  
= 6 000

Total exposed  =  71 750 + 6 000  
= 80 000 (approx.)

B.4 Estimated fatal accident rate

Assuming that an exposed employee works a 40 hour week, 48 weeks of the year, then:

Number of exposed hours per employee per year  =  40 × 48  
= 80 000

Total number of exposed persons  =  1920

Therefore number of exposed hours of all employees per year  =  1920 × 80 000  
= 1.54 × 10⁸ hours
Using the figure of 0.9 fatalities per year,
FAR for ignition of flammable leaks

\[
\frac{0.9}{1.54 \times 10^8} = 0.6 \text{ fatalities per } 10^8 \text{ exposed hours}
\]

This estimate is only very approximate due to only one year’s detailed data being available and the simplifying assumptions made. Greater accuracy may be achieved as more data are collected.

### B.5 Incident fatality and injury rate

The data collected and assumptions made for the fire and explosion survey were further crosschecked by considering the incident fatality and injury rates. These are defined as the number of injuries or fatalities occurring per incident of a given type. Using the data in Tables B.2 - B.5:

For closed plant,

- Number of incidents = 86
- Number of injuries = 7
- Number of fatalities = 1

Therefore,

\[
\begin{align*}
\text{Incident injury rate} & = \frac{8}{86} = 0.093 \\
\text{Incident fatality rate} & = \frac{1}{86} = 0.012
\end{align*}
\]

These figures were compared with results in previously mentioned work by Forsth concerning platforms in the Norwegian North Sea and also with results in a similar study for the Gulf of Mexico [95]. The Forsth results can be summarised as:

<table>
<thead>
<tr>
<th></th>
<th>Gulf of Mexico</th>
<th>Norwegian North Sea</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fatalities/incident</td>
<td>0.24</td>
<td>0.04</td>
</tr>
<tr>
<td>Injuries/incident</td>
<td>0.69</td>
<td>0.04</td>
</tr>
</tbody>
</table>
Two points were noted concerning the above figures before conclusions were drawn. Firstly, Forsth indicated that the large difference between the Gulf and North Sea figures was principally due to different reporting procedures i.e. Gulf incidents were under-reported. Secondly, the North Sea fatality ratio is high because it was wholly derived from one incident involving five fatalities. The injury ratio gives a truer reflection, being derived from five separate incidents with one injury in each. It is this figure which is broadly comparable with the result obtained from the fire and explosion survey given above, and so provided a moderately good crosscheck.
Table B.1 Fire and explosion survey: summary of incidents

<table>
<thead>
<tr>
<th>Source of ignition</th>
<th>Number of incidents</th>
<th>Fatalities</th>
<th>Injuries</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flames: general*</td>
<td>237</td>
<td>2</td>
<td>222</td>
</tr>
<tr>
<td>LPG fired equipment</td>
<td>24</td>
<td>0</td>
<td>15</td>
</tr>
<tr>
<td>Hot surfaces</td>
<td>48</td>
<td>1</td>
<td>30</td>
</tr>
<tr>
<td>Friction</td>
<td>36</td>
<td>0</td>
<td>15</td>
</tr>
<tr>
<td>Electrical</td>
<td>70</td>
<td>1</td>
<td>53</td>
</tr>
<tr>
<td>Hot particles</td>
<td>20</td>
<td>0</td>
<td>17</td>
</tr>
<tr>
<td>Static electricity</td>
<td>19</td>
<td>0</td>
<td>10</td>
</tr>
<tr>
<td>Smoking</td>
<td>38</td>
<td>3</td>
<td>35</td>
</tr>
<tr>
<td>Autoignition</td>
<td>25</td>
<td>0</td>
<td>7</td>
</tr>
<tr>
<td>Spontaneous ignition</td>
<td>26</td>
<td>0</td>
<td>14</td>
</tr>
<tr>
<td>Hotwork</td>
<td>120</td>
<td>4</td>
<td>97</td>
</tr>
<tr>
<td>Other</td>
<td>5</td>
<td>0</td>
<td>5</td>
</tr>
<tr>
<td>Unknown</td>
<td>300</td>
<td>5</td>
<td>210</td>
</tr>
<tr>
<td>Total</td>
<td>968</td>
<td>16</td>
<td>730</td>
</tr>
</tbody>
</table>

* excludes all other flame sources listed
Table B.2 Fire and explosion survey: incidents concerning closed process plant

<table>
<thead>
<tr>
<th>Source of ignition</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
<th>8</th>
<th>9</th>
<th>Total</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flame</td>
<td>3</td>
<td></td>
<td>1</td>
<td>1</td>
<td></td>
<td>1</td>
<td></td>
<td>2</td>
<td></td>
<td>8</td>
</tr>
<tr>
<td>LPG fired equipment</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>1</td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td>2</td>
</tr>
<tr>
<td>Hot surface</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td></td>
<td>1</td>
<td>3</td>
<td>1</td>
<td>10</td>
</tr>
<tr>
<td>Friction</td>
<td>3</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td></td>
<td>4</td>
</tr>
<tr>
<td>Electrical</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>5</td>
<td>3</td>
<td></td>
<td>8</td>
</tr>
<tr>
<td>Hot particle</td>
<td>1</td>
<td></td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td>3</td>
</tr>
<tr>
<td>Static electricity</td>
<td>1</td>
<td>2</td>
<td>2</td>
<td>1</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>6</td>
</tr>
<tr>
<td>Smoking</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Autoignition</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td></td>
<td>2</td>
<td></td>
<td>1</td>
<td>1</td>
<td></td>
<td>7</td>
</tr>
<tr>
<td>Unknown</td>
<td>9</td>
<td>4</td>
<td>5</td>
<td>3</td>
<td>1</td>
<td>4</td>
<td>7</td>
<td>5</td>
<td></td>
<td>38</td>
</tr>
</tbody>
</table>

Total 19 8 10 7 4 2 13 17 6 86

Column nomenclature:

1. Plant (unspecified)
2. Reactor
3. Vessel
4. Tank
5. Heat exchanger
6. Vaporiser
7. Pump
8. Pipework
9. Hose
Table B.3 Fire and explosion survey: injuries and fatalities on closed process plant

<table>
<thead>
<tr>
<th>Source of ignition</th>
<th>Number of injuries (column nomenclature below)</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>7</th>
<th>8</th>
<th>9</th>
<th>Total</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flame LPG fired equipment</td>
<td>1(1)</td>
<td></td>
<td>1</td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>3(1)</td>
</tr>
<tr>
<td>Hot surface</td>
<td>-</td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td></td>
<td>2</td>
</tr>
<tr>
<td>Friction</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Electrical</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Hot particle</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Static electricity</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Smoking</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Autoignition</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Unknown</td>
<td>-</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>1</td>
<td></td>
<td></td>
<td>2</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td><strong>1(1)</strong></td>
<td>1</td>
<td>1</td>
<td>2</td>
<td></td>
<td></td>
<td>2</td>
<td></td>
<td></td>
<td></td>
<td><strong>7(1)</strong></td>
</tr>
</tbody>
</table>

The injury figures exclude fatalities, which are shown in brackets.

Column nomenclature

1. Plant (unspecified)        6. Vaporiser
2. Reactor                    7. Pump
3. Vessel                     8. Pipework
4. Tank                       9. Hose
5. Heat exchanger             

B9
Table B.4 Fire and explosion survey: incidents concerning plant and activities with open surfaces

<table>
<thead>
<tr>
<th>Source of ignition</th>
<th>Total</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flames</td>
<td>27</td>
</tr>
<tr>
<td>LPG fired equipment</td>
<td>2</td>
</tr>
<tr>
<td>Hot surface</td>
<td>20</td>
</tr>
<tr>
<td>Friction</td>
<td>11</td>
</tr>
<tr>
<td>Electrical</td>
<td>29</td>
</tr>
<tr>
<td>Hot particle</td>
<td>0</td>
</tr>
<tr>
<td>Static electricity</td>
<td>10</td>
</tr>
<tr>
<td>Smoking</td>
<td>17</td>
</tr>
<tr>
<td>Autoignition</td>
<td>2</td>
</tr>
<tr>
<td>Unknown</td>
<td>21</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>139</td>
</tr>
</tbody>
</table>

Column nomenclature

1. Solvent evaporating oven
2. Spray booth
3. Small container
4. Cleaning/degreasing process
5. Tanker/mobile plant
6. Other
Table B.5 Fire and explosion survey: injuries and fatalities for plant and activities with open surfaces

<table>
<thead>
<tr>
<th>Source of ignition</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
<th>Total</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flames</td>
<td>1</td>
<td>-</td>
<td>6</td>
<td>2</td>
<td>3</td>
<td>1</td>
<td>13</td>
</tr>
<tr>
<td>LPG fired equipment</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>Hot surfaces</td>
<td>-</td>
<td>1</td>
<td>-</td>
<td>-</td>
<td>4(1)</td>
<td>1</td>
<td>6(1)</td>
</tr>
<tr>
<td>Friction</td>
<td>-</td>
<td>1</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>1</td>
<td>2</td>
</tr>
<tr>
<td>Electrical</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>3(1)</td>
<td>-</td>
<td>3(1)</td>
<td>0</td>
</tr>
<tr>
<td>Hot particle</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>Static electricity</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>Smoking</td>
<td>-</td>
<td>-</td>
<td>4</td>
<td>2</td>
<td>1</td>
<td>1</td>
<td>8</td>
</tr>
<tr>
<td>Autoignition</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>0</td>
</tr>
<tr>
<td>Unknown</td>
<td>-</td>
<td>-</td>
<td>2</td>
<td>-</td>
<td>4</td>
<td>-</td>
<td>6</td>
</tr>
</tbody>
</table>

Total               | 1 | 2 | 12 | 4 | 15(2) | 5 | 39(2) |

The injury figures exclude fatalities, which are shown in brackets.

Column nomenclature

1. Solvent evaporating oven  
2. Spray booth  
3. Small container  
4. Cleaning/degreasing process  
5. Tanker/mobile plant  
6. Other
Table B.6 Other HSE accident statistics for the chemical and oil industries

A. NADOR Fatal and major injuries; 1981 to 1985

<table>
<thead>
<tr>
<th>Industry</th>
<th>Number of fatal (F) and major injuries (M)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>F</td>
</tr>
<tr>
<td>Chemical</td>
<td>8</td>
</tr>
<tr>
<td>Mineral oil processing</td>
<td>3</td>
</tr>
<tr>
<td>Total</td>
<td>11</td>
</tr>
</tbody>
</table>

B. NADOR Fatal and major injury rates; 1981 to 1985

<table>
<thead>
<tr>
<th>Industry</th>
<th>Fatal and major injury rates (incidence per 100 000 employees)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chemical</td>
<td>89.4</td>
</tr>
<tr>
<td>Mineral oil processing</td>
<td>108.4</td>
</tr>
</tbody>
</table>

C. HM Chief Inspector’s Annual Report 1984; fatal and major injury rates

<table>
<thead>
<tr>
<th>Industry</th>
<th>Number employed</th>
<th>Fatalities</th>
<th>Major injuries</th>
<th>Fatalities and major injuries per 100 000 persons</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chemical industry</td>
<td>360 000</td>
<td>5</td>
<td>349</td>
<td>98.4</td>
</tr>
<tr>
<td>Other chemical processes</td>
<td>38 200</td>
<td>1</td>
<td>38</td>
<td>102.1</td>
</tr>
<tr>
<td>Mineral oil refining</td>
<td>18 200</td>
<td>1</td>
<td>14</td>
<td>82.4</td>
</tr>
</tbody>
</table>
Appendix C. The $\chi^2$ confidence test

In all the sampled failure frequencies given in Section 7 it was a point value of the component failure frequency which was given. However, to gain an impression of overall reliability the true failure frequency was required i.e. the failure frequency that would result if all the components in the sample were allowed to fail. The point value and the true value can be related by means of a confidence test e.g. a 90% confidence test on a sampled failure frequency gives the upper and lower limits around that frequency within which there is a 90% probability that the true failure frequency lies.

Such a test may be performed on a sampled failure frequency if it may be assumed that the failure frequency is constant with time, in other words that failures occur randomly. In these conditions Smith [33] stated that the expression

$$\frac{2k\bar{\theta}}{\theta}$$

follows a $\chi^2$ distribution with $2k$ degrees of freedom such that the test is truncated after the $k$th failure and where

- $\theta$ = overall mean time between failures
- $\bar{\theta}$ = observed mean time between failures
- $T$ = total test time
- $k$ = Number of failures

Smith further showed that:

$$\frac{2k\bar{\theta}}{\theta} = \frac{2kT}{k\theta} = \frac{2T}{\theta}$$

Since the mean time between failures is the reciprocal of the failure frequency then the failure frequency can also be assumed to follow the $\chi^2$ distribution. Table C.1 shows a table of
the $\chi^2$ distribution where $\alpha$ is (1 - the confidence level) and $n$ is the number of degrees of freedom. Confidence tests may be one or two sided depending on whether both the lower and upper limits are desired. For two sided tests such as the ones carried out in Section 7 the limits were calculated as follows:

\[
\begin{align*}
\text{Upper confidence limit} & = \frac{\chi^2\left(\frac{\alpha}{2}, 2k + 2\right)}{2T} \\
\text{Lower confidence limit} & = \frac{\chi^2\left(1 - \frac{\alpha}{2}, 2k\right)}{2T}
\end{align*}
\]

For the upper limit the number of degrees of freedom is $2k + 2$ as the tests in Section 7 were time truncated. 90% confidence limits are the most commonly used, so $\alpha = 0.1$.

Finally, for some of the samples in Section 7 it was necessary to estimate confidence limits where the number of degrees of freedom was greater than 100. In these cases a prediction given by Kreysig [126] was used for the 90% limits where:

\[
\begin{align*}
\text{Upper confidence limit} & = \frac{(h + 1.64)^2}{2} \\
\text{Lower confidence limit} & = \frac{(h - 1.64)^2}{2} \\
\text{and where} \quad h & = \sqrt{2k - 1}
\end{align*}
\]
Table C.1 Percentage points of the $\chi^2$ distribution taken from Smith [33]
Appendix D. Programmed emission and dispersion models used in the estimation of hazard distances

This appendix contains listings of computer programs written to enable the convenient calculation of hazard distances in Section 10. Details of which routine was used to generate any particular figure is given below (the program name is given on the first line of the listing):

Table D.1 Application of programmed models to hazard distance estimation

<table>
<thead>
<tr>
<th>Program name</th>
<th>Models contained</th>
<th>Applications in Section 10:</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Leak item</td>
</tr>
<tr>
<td>ewanjet</td>
<td>Sonic gas emission rate</td>
<td>Flange</td>
</tr>
<tr>
<td></td>
<td>Gas compressibility</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Axial jet concentration</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Compressor</td>
<td></td>
</tr>
<tr>
<td>resist</td>
<td>Liquid jet emission rate</td>
<td>Flange</td>
</tr>
<tr>
<td></td>
<td>Inclined liquid jet throw with air resistance</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Test for atomisation</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Maximum drop size in atomised spray</td>
<td>Pump</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>meckevap</td>
<td>Volatile pool evaporation rate</td>
<td>Flange</td>
</tr>
<tr>
<td></td>
<td>Pool spreading rate</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Pool radius</td>
<td>Pump</td>
</tr>
</tbody>
</table>

continued
<table>
<thead>
<tr>
<th>pasgiff</th>
<th>Axial plume concentration</th>
<th>Flange</th>
<th>Vaporising impacted liquid jet hazard distance</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Test for gravity-controlled dispersion</td>
<td>Pump</td>
<td>Liquid spray hazard distance</td>
</tr>
<tr>
<td></td>
<td>Distance to gravity/Gaussian dispersion transition point</td>
<td>Compressor</td>
<td>Bunded liquid vapour release hazard distance</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Drain/ Sample point</td>
<td>Propane hazard distance</td>
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<tr>
<td></td>
<td></td>
<td></td>
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</tr>
<tr>
<td>physical</td>
<td>Physical property correlations (see Appendix E)</td>
<td></td>
<td>Propane hazard distance</td>
</tr>
</tbody>
</table>

D2
Main Program

1 ccccccyellow.fortran
2 real lel,mass,massfr,mol,pi,po,prat,p1
3 character*12 emit
4 character*1 query,quer2,quer3
5 print","***************
6 print","******
7 print","** EMISSION ROUTINE FOR SONIC RELEASES OF GAS **
8 print","** USING EQUATIONS GIVEN BY EWAN AND MOODIE **
9 print","**
10 print","******
11 print","***************
12 print","***************
13 do 1000,i=1,10
14 print"," "
15 1000 continue
16 print","Gas emitted?"
17 read*,emit
18 print","Molecular weight of ",emit," (kg.kmol-l) ="
19 read*,mol
20 print","LEL of ",emit," (vol%) ="
21 read*,lel
22 print","Critical temperature of ",emit," (K) ="
23 read*,tcrit
24 print","Critical pressure of ",emit," (bar) =
25 print","Pressure in enclosure (bar) =
26 print","Temperature in enclosure (K) =
27 print","Ambient temperature (K) =
28 print","Adiabatic gas coefficient =
29 read*,gamma
30 print","Hole diameter (m) =
31 read*,dexit
32 print","Discharge coefficient known? y or n"
33 read*,quer2
34 if (quer2.eq."y") then
35 print","Discharge coefficient =
36 read*,cd
37 else if (quer2.eq."n") then
38 cd=.6
39 endif
40 quer3="0"
41 10 if ( quer3.eq."0") goto 20
42 print","pressure in enclosure (bar) =
43 read*,p1
print*,"Temperature in enclosure (K) = "
read*, tl
print*,"Hole diameter (m) = "
read*, dexit

r = 8214.4
pi = 3.14159
po = 1.01324
patm = po
prat = (po / pl)
comp = (2 / (1 + gamma)) ** (gamma / (gamma + 1))
if (prat .ge. comp) then
  print*, "Release is subsonic"
goto 90
else if (query, eq."y") goto 80
ccc ccccccccccc****FELICHI-KWONG ROUTINE FOR COMPRESSIBILITY*************

akl = 0.42748 * (r**2) * (tcrit**2.5) / pcrit
bakl = 0.08664 * r * tcrit / pcrit
ak = akl * pl * le05 / (r**2) * (tcrit**2.5)
bak = bakl * pl * le05 / (r * t1)
cak = ak - bak * (1 + bak)
dak = ak * bak
90 print*,"Guess compressibility"
read*, zo
kak = 1
70 fak = (zo**3) - (zo**2) + kak * zo - dak
gak = 5 * (zo**2) - 2 * zo + kak
z1 = zo - (fak / gak)
if (z1 <= 0) then
  print*, "Compressibility < 0"
goto 90
else if (z1 > 1) then
  print*, "Compressibility > 0"
goto 90
else if (abs(z1 - zo) .le. 0.001) then
  zcomp = 1
goto 80
else if (kak .gt. 50) then
  print*, "Too many iterations"
goto 90
else
  zo = z1
  kak = kak + 1
  goto 70
endif

ccc ccccccc****DENSITIES OF AMBIENT GAS,AIR**********************
80 rhoa = 29 / (22.414 * to / 273)
rhoamb = mol / (22.414 * to / 273)
t2 = 2 * t1 / (1 - gamma)
15 ccccccc***********EXIT PRESSURE**********************
14 pexit = pl * (2 / (1 + gamma)) ** (gamma / (gamma - 1))
15 ccccccc***********EXIT DENSITY**********************
16 rhot = le05 * pl * mol / (zcomp * r * t1)
17 rhor = rhot * ((2 / (gamma + 1)) ** (1 / (gamma - 1)))
18 ccccccc***********EQUIVALENT DIAMETER AND DENSITY*************
deq = dexit * sqrt(pexit / patm)
req = deq / 2
rheq = rhor * eq / patm / pexit
cccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc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Main Program

1 ccccccliquid throw model
2 real mass,kdrag
3 integer test
4 character*1 quer1, quer2, quer3
5 character*8 emit
6 print*, "*****************************************************************************"
7 print*, "*****************************************************************************"
8 print*, "** MODEL FOR THROW OF A LIQUID JET **"
9 print*, "** INCORPORATING AIR RESISTANCE ALLOWANCE **"
10 print*, "*****************************************************************************"
11 print*, "*****************************************************************************"
12 do 1000,i=1,5
13 print*, " "
14 1000 continue
15 1000,i=1,5
16 print*, "Liquid emitted ?"
17 read*, emit
18 print*, "Density of ",emit," (kg.m-3)"
19 read*, rhol
20 print*, "Viscosity of ",emit," (kg.m-1.s-1)"
21 read*, viscl
22 print*, "Surface tension of ",emit," (Pa.m-1)"
23 read*, surft
24 print*, "Upstream pressure? (bar)"
25 read*, po
26 po=1e05*po
27 print*, "Area of leak orifice? (m2)"
28 read*, area
29 print*, "Discharge coefficient of leak orifice?"
30 read*, cd
31 print*, "Drag coefficient of jet?"
32 read*, kdrag
33 print*, "Ambient temperature? (K)"
34 read*, tamb
35 print*, "Height of leak above ground? (m)"
36 read*, hite
37 print*, "Angle leak makes to horizontal? (deg)"
38 read*, thet
39 theta =2*3.14159*thet/360
40 print*, " "
41 cccccPHYSICAL CONSTANTS*****************************************************************************
42 patm=101324
43 pi=3.14159
44 rhoa=29*273/(22.4*tamb)
45 g=9.8066
46 diam=sqrt(4*area/pi)
47 ccccJET VELOCITY*****************************************************************************
48 vel=mass/(area*rhol)
49 \m
50 print*, "TEST FOR SPRAY FORMATION - REYNOLD/WEBER NO. COMPARISON***
51 print*, " "
52 test=0
53 Re=vel*diam*rhol)/viscl
Webber No.

\[
\text{Webber No.} = \frac{(\text{vel} \times \text{diam} \times \text{rho})}{\text{surf}}
\]

if \( \text{Webber No.} \leq 0.45 \times 10^6 \) then

\[
\text{test} = 1
\]

DROP DIAMETER USING DROP WEBER NO.

\[
\text{ddrop} = 20 \times \text{surf} / (\text{vel} \times \text{rho})
\]

endif

TIME TO MAX JET HEIGHT

\[
\text{tmax} = \left( \frac{1}{\text{drag}} \right) \log \left( \frac{(\text{ddrop} \times \text{vel} \times \text{sin}(\theta)) + g}{g} \right)
\]

DISTANCE TO MAX JET HEIGHT

\[
\text{xmax} = \left( \frac{(\text{ddrop} \times \text{vel} \times \text{sin}(\theta)) + g}{g} \right) \times (1 - \exp(-\text{drag} \times \text{tmax}))
\]

MAX JET HEIGHT

\[
\text{ymax} = \left( \frac{(\text{ddrop} \times \text{vel} \times \text{sin}(\theta)) + g}{g} \right) \times (1 - \exp(-\text{drag} \times \text{tmax}))
\]

CHANGE TO JET FALLING PROCEDURE

\[
\text{dxdt} = \text{vel} \times \text{cos}(\theta)
\]

ITERATE FOR FALLING TIME

\[
\text{tfall} = 2 \times \text{tmax}
\]

\[
\text{Ntot} = \text{Nmax} + \text{xfall}
\]

RESULTS

\[
\text{do } i = 1, 5
\]

\[
\text{print}, "1", i, \text{print} \quad \text{continue}
\]

\[
\text{print}, "\text{Liquid emission of } \text{emit}
\]

\[
\text{print}\,, "\text{Mass flow}\, = \text{\text{"mass\," \,kg\,s\,-\,1\"}
\]

\[
\text{print}\,, "\text{Velocity}\, = \text{\text{"vel\," \,m\,s\,-\,1\"}
\]

\[
\text{print}\,, "\text{Projected angle of emission}\, = \text{\text{"theta\," \,deg.\"}
\]

\[
\text{print}\,, "\text{Height of emission}\, = \text{\text{"hite\," \,m\"}
\]

\[
\text{print}\,, "\text{Projected angle of emission}\, = \text{\text{"theta\," \,deg.\"}
\]

if \( \text{test} = 1 \) then

\[
\text{print}, "\text{Warning - spray predicted}\"\]

\[
\text{print}\,, "\text{Jet Reynolds number}\, = \text{\text{"Re\"}
\]

\[
\text{print}\,, "\text{Jet Weber number}\, = \text{\text{"We\"}
\]

\[
\text{print}\,, "\text{Maximum drop diameter}\, = \text{\text{"ddrop\,1000\," mm\"}
\]

end if

\[
\text{print}\,, "\text{Maximum height of emission}\, = \text{\text{"(hite\,\text{ymax})\," m\"}
\]

\[
\text{print}\,, "\text{Horizontal distance from source}\, = \text{\text{"xmax\," m\"}
\]

\[
\text{print}\,, "\text{Horizontal distance}\, = \text{\text{"xmax\," m\"}
\]

\[
\text{print}\,, "\text{another run at a different angle? \,v or n\"}
\]
read*, quer1
if (quer1.eq."y") goto 10
print*,"Another run at a different height? y or n"
read*, quer2
if (quer2.eq."y") goto 20
goto 90
print*,"Another run at a different height? y or n"
read*, quer3
if (quer3.eq."n") goto 90
print*,"Choose a value for falling jet time (secs)"
read*, tfall
goto 49
90 stop
end
Main Program

1 cccccccyelpool.fortran
2  real mass,mas2,mw,mrate,mrat2
3  character*8 evap
4  character*1 quer1 ,quer2
5  print*,"******************************************************************
6  print*,"******************************************************************
7  print*,"** CALCULATION OF VOLATILE POOL EVAPORATION RATE ACCORDING TO MECKLENBURGH **
8  print*,"** 
9  print*,"******************************************************************
10  print*,"******************************************************************
11  print*,"******************************************************************
12  do 2000, i=1,10
13  print*, "
14  2000 continue
15  print*,"Evaporating fluid?"
16  read*,evap
17  print*,"Molecular weight of ",evap," (kg/kmol)"
18  read*,mw
19  print*,"Density of ",evap," (kg.m-3)"
20  read*,dens
21  print*,"Temperature of spill (K)"
22  read*,tpool
23  print*,"Vapour pressure of ",evap," (bar)"
24  read*,psat
25  psat=1e05*psat
26  print*,"Ambient partial pressure of ",evap," (bar)"
27  read*,ppamb
28  ppamb=1e05*ppamb
29  print*,"Stability constant ne"
30  read*,stabc
31  print*,"Stability constant lambda"
32  read*,stabk
33  print*,"Minimum height of pool (m)"
34  read*,hmin
35  print*, "
36  cccccccPROGRAM CALCULATES THE EVAPORATION RATE OF A SPREADING
37  cccccccPOOL OF VOLATILE LIQUID ACCORDING TO THE SUTTON-PASQUILL
38  cccccccEVAPORATION MODEL
39  cccccccPHYSICAL CONSTANTS*******************************
40  pi=3.14159
41  r=6314.4
42  patm=101324
43  vdens=mw*273/(22.4136*tpool)
44  print*,"Wind speed (m.s-1)"
45  read*,wind
46  print*,"Spill rate (kg.s-1)"
47  read*,spill
48  print*,"Time increment (s)"
49  read*,tinc
50  print*,"Total time (s)"
51  read*,ttot
52  do 3000, i=1,5
53  print*, "
54  3000 continue
print*,"*************************************************
print*,"Evaporation of \"evap
print*,"********************************************************
print*,"Spill rate = ",spill,\" (kg.s-1)\"
print*,"********************************************************
print*,"Time Pool Evaporation Total"
print*,"(s) radius (m) rate (kg.s-1) (kg)"
print*,"********************************************************
t=1
etot=0
mrat2=0
tlast=0
80 vol=(spill*t-etot)/ldens
rad=sqrt(vol/(pi*hmin))
ccc
**EVAPORATION RATE**********
parta=stabk*wind*rad**2*pi*(psat/patm)
partb=(1/(rad*wind**2))**stabh
mass=parta*partb*vdens
etot=etot+mass*t-tlast)
print 201, t, rad, mass, etot
201 format (h ,f6.0,f7.4,f7.4,f7.4,f9.4)
tlast=t
if (abs(mass-mas2).le.1e-07.or.t.gt.ttot) goto 50
mas2=mass
10 t=tinc+t
20 goto 80
50 print*,"Another run at different windspeeds? y or n"
read*,quer2
if (quer2.eq."y") goto 90
stop
end
Main Program

1 cccccc PASQUILL-GIFFORD DISPERSION CALCULATION
2 real m,mw,lel,uw,h,z,a,b,c,d,l0,tav,xinc,x,dx,pi,c,ctav,sigz,
3 \*giff,g1,g2,cc,vper
4 integer i,k
5 character*8 disp
6 character*15 stab
7 print*, "************************************************************
8 print*, "PASQUILL-GIFFORD DISPERSION CALCULATION BASED ON"
9 print*, "THE PROCEDURE GIVEN IN THE TNO YELLOW BOOK"
10 print*, "CONTINUOUS POINT SOURCE"
11 print*, "************************************************************
12 do 1000, i=1,10
13 print*, 
14 1000 continue
15 print*, "Dispersing gas?"
16 read*, disp
17 print*, "Molecular weight of ",disp," (kg.kmol-l) = 
18 read*, mw
19 print*, "Source strength of ",disp," (kg.s-l) = 
20 read*, m
21 print*, "LEL of ",disp," (vol%) = 
22 read*, lel
23 print*, "Wind speed 10m above ground (m.s-l) ="
24 read*, uw
25 print*, "Height of release point above ground level (m) ="
26 read*, h
27 print*, "Cross-wind diameter of release point (m) ="
28 read*, ds
29 print*, "Height of desired concentration axis above ground ="
30 read*, z
31 print*, "Atmospheric stability class: a,b,c,d,e or f"
32 read*, stab
33 if (stab.eq."a") then
34 a=.527
35 b=.865
36 c=.28
37 d=.9
38 else if (stab.eq."b") then
39 a=.371
40 b=.966
41 c=.23
42 d=.85
43 else if (stab.eq."c") then
44 a=.219
45 b=.997
46 c=.22
47 d=.80
48 else if (stab.eq."d") then
49 a=.128
50 b=.905
51 c=.20
52 d=.76
53 else if (stab.eq."e") then
54 a=.098
55 b=.902
else if (stab.eq."f") then
a=.065
b=.902
c=.12
d=.67
endif
print*"," "
print*,"Ambient temperature (K) ="
read*,ambt
print*"," "
print*,"Yellow book roughness factor ZO (m) ="
read*,zo
print*"," "
print*,"Time interval for average concentration ="
print*,"(minimum 20 seconds) ="
read*,tav
print*"," "
print*"," "
print*,"Incremental distance for printout (m) ="
read*,xinc
x=xinc/20
d=x
volm=22.414*ambt/273
rhoa=29*273/(22.414*ambt)
g=9.8066
rhog=mw*273/(22.414*ambt)
do 2000 i=1,5
print*"," 
2000 continue
print*"," ***********************************************
print*,"PASQUILL-GIFFORD DISPERSION CALCULATION FOR A "
print*,"RELEASE OF GAS ACCORDING TO THE TNO YELLOW BOOK"
print*," (CONTINUOUS POINT SOURCE)"
print*"," ***********************************************
print*"," "
print*,"Gas dispersing = ",disp
print*,"Source strength = ",m," kg.s-1"
print*,"Height of source = ",h," m"
print*,"Height of x=axis = ",z," m"
print*"," 
print*,"Pasquill stability class = ",stab
print*,"Wind speed = ",uw," m.s-1"
print*,"TNO roughness factor = ",zo," m"
print*"," 
print*"," ***********************************************
print*,"Distance Concentration Volume ",km-
print*,"from along x-axis percent "
print*,"source(m) (kg,m-3) (\\%)
print*"," ***********************************************
ccccccCALCULATION OF S.D.Y & Z
vperl=0
pi=3.14159
k=1
30 czo=(10*zo)**(.53*x**(-.22))
c tav=(tav/b00)**.2
if (x.ge.100) then
sigz=czo*c*x**d
sigy=ctav*a**b
 goto 10
else if (x.lt.100) then
sigz=(((czo*c#100**d)/100)*x
sigy=((ctav*a#100**b)/100)*x
calculation of conc. over distance

g1 = m / (2 * pi * uw * sigz * sigz)
g2 = exp((-(z-h)**2) / (2 * sigz**2))
g3 = exp((-(z+h)**2) / (2 * sigz**2))
cc = g1 * (g2 + g3)

vper = 100 * (volm * cc / mw) / (1 + (cc + volm / mw))

if (x = selected increment or if conc. is close to lel print

results

if (vper < lel and vper < vperl)
goto 10
endif

if (k = eq. 20) then
k = 0
endif

if (vper < 0,0) goto 20
endif

if (dim(vper, lel). and vper < 0,5 * lel) goto 30
endif

40 = x + dx
vper = vper
k = k + 1

print 200, k, cc, vper

format (ih, f8.2, 9k, f8.3, 10h, f8.3)

ccccc:***********HANNA-DRIVAS TEST FOR HEAVY GAS DISPERSION***********

ccccc:***********CALCULATE FRICTION VELOCITY******************************

ufric = 0.4 * uw / (log(z0 / 10))

ccccc:***********CALCULATE RICHARDSON NUMBER*******************************

rio = (g / ufric**2) * ((rhog - rhoa) / rhoa) * (m / (rhog * uw * ds))

if (rio > 10) then
print, "Warning - dispersion is initially gravity"
print, "Critical Richardson number = 10.0"
endif

print, "Source Richardson number = " , rio
print, "Critical Richardson number = 10.0"

print, "Distance to gravity-Gaussian transition point m"
print, "= " , xt , " m"

stop
Main Program

1 include physical.fortran
2 real mu,muref,mw,lhs,ldens,ldensr,k,j
3 character*8 emit
4 character*1 quer1,quer2
5 character*3 na
6 integer mfac
7 na=-1
8 print","*******************************************************************"
9 print","*******************************************************************"
10 print","**
11 print","** PHYSICAL PROPERTIES GENERATION PROGRAM FOR **
12 print","** HAZARDOUS AREA MODELS **
13 print","**
14 print","*******************************************************************"
15 print","*******************************************************************"
16 do 1000 i=1,5
17 print"," "
18 1000 continue
19 print","Emitted substance ?"
20 read*,emit
21 print","Molecular weight (kg.kmol-1)"
22 read*,mw
23 print","Normal boiling point (K)"
24 read*,tnbp
25 print","Wagner vapour pressure coefficients a,b,c,d"
26 read*,vpa,vpb,vpc,vpd
27 print","Cpg correlation coefficients"
28 read*,cpa,cpb,cpc,cpd
29 print","Critical temperature (K)"
30 read*,tcrit
31 print","Critical pressure (bar)"
32 read*,pcrit
33 print","Critical volume (cm3.mol-1)"
34 read*,vcrit
35 print","Acentric factor w"
36 read*,w
37 print","Dipole movement (debye)"
38 read*,mu
39 print","Polar correction factor (0 if unknown)"
40 read*,k
41 print","Liquid viscosity correlation coefficients a,b,c,d"
42 read*,visca,viscb,viscc,viscd
43 r=8.31441
44 print","Experimental liquid density known ? y or n"
45 read*,quer2
46 if (quer2.eq."y") then
47 print","Experimental density (kg.m-3)"
48 read*,ldensr
49 var=1000*mw/ldensr
50 print","Reference temperature (K)"
51 read*,tref
52 endif
53 40 print","Starting temperature (K)"
54 read*,tstart

D14
55  print*,"Final temperature (K)"
56  read*,tfin
57  print*,"Incremental temperature (K)"
58  read*,tinc
59  do 2000 i=1,5
60   print* " "
61 2000 continue
62  print* "Physical properties of \( \text{emit} \)\( \text{pr} \)\*
63  print*, "\( \text{Temp} \) Psat dPsat dens Cpg Cpl \text{Visc Surf}"
64  print*, "(K) \( \text{bar} \) dT (kg. \( \text{J.g}^{-1. \text{K}^{-1}} \) g \( \text{l tens} \)"
65  print*, "m=3) \( \text{uF cp dyn/cm} \)"
66  print* " "
67  print*, "\( \text{t} = \text{start} \)
68  print* " "
69  t=tstart
70  cccccc************VAPOUR PRESSURE***************
71  30 tredEt/tcrit .......
72  tfac=1-tred
73  if (tfac.le.0) goto 10
74  cccccc************ALTERNATIVE DP/DT **************
75  35 psat=pcrit*exp ((vp*a+vp*b+tfac**1.5+vp*c+tfac**2+vdp*tfac**6)
76   /t(red)
77  cccccc************dPsat/dT***********************
78  ududt=-vpa/t-(1.5*vpb*sqrt(tfac))/t-(3*vp*c+tfac**2/t)
79  uvdvdt=(-tcrit/t+2)*(vp*a+vp*b+tfac**1.5+vp*c+tfac**2+vdp*tfac**6)
80  dpdt=psat*(ududt+uvdvdt)
81  cccccc*************LIQUID DENSITY***************
82  zra=0.29056-0.08775*w
83  zrafac=zra**(1+tfac**(2.0/7.0))
84  vs=1.0**t+tcrit*zrafac/pcrit
85  ldens=1e03*mw/vs
86  if (quer2.ne."y") gete 50
87  cccccc************LIQUID SPECIFIC HEAT*************
88  lhs=1.45+0.45*tfac+0.25*w*(17.11+5.2*tfac**(1.0/3.0))
89  *+1.742/tfac)
90  cpl=r*lhs+cpg
91  cp1m=cpl/mw
92  cccccc************LIQUID VISCOSITY***************
93  etal=exp(visca+viscb*v+viscc*t+viscd*t**2)
94  cccccc************GAS SPECIFIC HEAT***************
95  10 cp=g+cpg+cpp+cpp*c*t**2+cpd*t**3
96  cpg=cpg/mw
97  cccccc************GAS VISCOSITY***************
98  muref=131.3*mu/(sqrt(vcrit*tcrit))
99  ffac=1.0-0.2756*w*0.059035*muref**4+k
100 tstar=1.2893*tred
101 omega=1.16145*tstar**0.14874+0.52847*(exp(-0.7732*tstar))
102   *(exp(-2.4587*tstar))
103 etag=40.785*ffac*sqrt(mw*t)/(vcrit**2.0/3.0)*omega)
104 cccccc************SURFACE TENSION OF LIQUID**********
105 tbred=tnbp/tcrit
106 cccccc************TENSION SET TO 9000ETC TO PREVENT VALUE BEING PRINTED
107 cccccc************CALCULATION ONLY VALID FOR LIQUIDS AT ATMOSPHERIC PRESSURE
108 q=tbred*log(pcrit/1.01325)/(1-tbred)
109 q=0.1196*(1-q)+0.279
110 surf=pcrit**2.0/3.0*tcrit*(1.0/3.0)*q*(1-tbred)**(11.0/9.0)
111 if (t.ge.tnbp) then
112 surf=9000000000
113 endif

D15
if (t.gt.tcrit) then
  print 201, t, na, na, na, cpgm, na, etag, na, na
else
  print 202, t, psat, ldens, cplm, etal
  print 203, dpdt, cpgm, etag, surf
endif
201 format (lh, f5.1, x, f5.0, x, f3.0, x, f3.0, x, f7.3, x, f3.0, x, x)
202 format (lh, f5.1, x, f6.3, x, f6.3, x, f8.3, x, f8.3, x, f8.3, x, f8.3)
203 format (lh, f7.5, x, f7.5, x, f8.3, x, f8.3, x, f8.3, x, f8.3, x, x)
t=tinc+t
  if (t.gt.tfin) goto 20
  goto 30
20 print*, "Another temp range? y or n"
  read*, querl
  if (querl.eq."y") goto 40
  stop
end
Appendix E. Physical property data for use with hazardous area models

This appendix lists physical property data correlations used to estimate the extent of hazardous zones in Section 10. The empirical constants etc. for each of the six hazardous substances considered in Section 10 are given in Section E.8.

E.1 Vapour pressure

Calculation of vapour pressure at a given temperature was done using the Wagner equation [127]:

\[
\ln P_{vr} = \frac{At + Bt^{1.5} + Ct^2 + Dt^6}{T_r}
\]  

(E.1)

where

\[
P_{vr} = \text{reduced vapour pressure} = \frac{P_{vp}}{P_c}
\]

\[
t = (1 - T_r)
\]

\[
T_r = \text{reduced temperature} = \frac{T}{T_c}
\]  

(E.2)

and where

\[
P_{vp} = \text{vapour pressure (bar)}
\]

\[
T = \text{temperature (K)}
\]

\[
P_c = \text{critical pressure (bar)}
\]

\[
T_c = \text{critical temperature (K)}
\]

Constants A, B, C and D have been experimentally determined and tabulated by Reid et al. [128].
E.2 Heat capacity of pure gases and vapours

Reid et al. have also given tabulated coefficients for use in the following third-order polynomial to estimate $C_{PG}$:

$$C_{PG} = A + BT + CT^2 + DT^3$$  \hspace{0.5cm} (E.3)

and where

- $C_{PG} = \text{heat capacity of the ideal gas at constant pressure (kJ.kmol}^{-1}.\text{K}^{-1})$
- $T = \text{temperature (K)}$

For the sonic gas emission model (Equation (9.3)) non-ideality of the emitted gas was partly compensated for by using the Redlich-Kwong equation of state to estimate gas compressibility (see Section 9.1). The sonic gas emission rate equation was considered sufficiently insensitive to changes in the adiabatic expansion coefficient $\gamma$ to enable the use of the ideal gas expression in estimating it:

$$\gamma = \frac{C_p}{C_p - R}$$  \hspace{0.5cm} (E.4)

where $R = 8.314 \text{ kJ.kmol}^{-1}.\text{K}^{-1}$.

E.3 Heat capacity of liquids

A representative value of $C_{PL}$ at 298 K for each of the six substances (where applicable) is given in the list of empirical constants at the end of this appendix. For heat capacities at other temperatures the Bondi-Rowlinson method [129][130] was used which predicts $C_{PL}$ using the corresponding value of $C_{PG}$:

$$\frac{(C_{PL} - C_{PG})}{R} = 1.45 + 0.45(1-T_r)^{-1} + 0.25\omega[17.11 + 25.2(1-T_r)^{12}T_r^{-1} + 1.742(1-T_r)^{-1}]$$  \hspace{0.5cm} (E.5)

$$\frac{(C_{PL} - C_{dl})}{R} = \exp(20.1T_r - 17.9)$$  \hspace{0.5cm} (E.6)

$$\frac{(C_{dl} - C_{sat,l})}{R} = \exp(8.655T_r - 8.385)$$  \hspace{0.5cm} (E.7)
where

\[ T_r = \text{reduced temperature from (E.2)} \]
\[ C_{PG} = \text{heat capacity of the ideal gas at } T_r \text{ (kJ.kmol}^{-1}.\text{K}^{-1}) \]
\[ C_{PL} = \text{change of liquid enthalpy with temperature at constant pressure (kJ.kmol}^{-1}.\text{K}^{-1}) \]
\[ C_{dl} = \text{variation in enthalpy of a saturated liquid with temperature (kJ.kmol}^{-1}.\text{K}^{-1}) \]
\[ C_{SAT.L} = \text{energy required to effect a temperature change whilst keeping the liquid saturated (kJ.kmol}^{-1}.\text{K}^{-1}) \]
\[ \omega = \text{acentric factor} \]

If \( T_r \) is below approximately 0.8 then \( C_{PL}, C_{dl} \) and \( C_{SAT.L} \) are effectively the same.

E.4 Gas viscosity

For the examples in Section 10 the gas viscosities where required were estimated using the following equation by Chung et al. [131]:

\[ \mu = 40.785 \left[ \frac{F_c(MT)^{1/2}}{V_c^2 \Omega_v} \right] \quad (E.8) \]

where

\[ \Omega_v = [A(T^*)^{-8}] + C[\exp(-DT^*)] + E[\exp(-FT^*)] \quad (E.9) \]
\[ F_c = 1 - 0.2756\omega + 0.059035\phi_k + \kappa \quad (E.11) \]
\[ \phi_k = 131.3 \left( \frac{\phi}{(V_c T_c)^{1/2}} \right) \quad (E.11) \]

and

\[ T^* = 1.2593 T_r \quad (E.12) \]

where

\[ A = 1.16145 \quad D = 0.77320 \]
\[ B = 0.14874 \quad E = 2.16178 \]
\[ C = 0.52487 \quad F = 2.43787 \]
and where

\[
\begin{align*}
\mu & \quad \text{viscosity (µP)} \\
F_c & \quad \text{shape and polarity factor} \\
M & \quad \text{molecular weight (kg.kmol}^{-1}) \\
T & \quad \text{temperature (K)} \\
V_c & \quad \text{critical volume (cm}^3\text{.mol}^{-1} = 0.001 \text{ m}^3\text{.kmol}^{-1}) \\
T_c & \quad \text{critical temperature (K)} \\
\Omega & \quad \text{viscosity collision integral} \\
\omega & \quad \text{acentric factor} \\
\phi & \quad \text{dipole movement (debye)} \\
\kappa & \quad \text{polar correction factor for highly polar molecules e.g. alcohols} \\
\rho & \quad \text{density (mol.cm}^{-3})
\end{align*}
\]

E.5 Liquid viscosity

Tabulated constants for pure substances have been given by Reid et al. [128] which were inserted into the following equation to predict liquid viscosities:

\[
\ln \mu = A + (B/T) + CT + DT^2
\]  
(E.19)

(for certain substances C and D are not given.)

\[
\begin{align*}
\mu & = \text{viscosity (cP)} \\
T & = \text{temperature (K)}
\end{align*}
\]

E.6 Liquid density

Rackett’s method was used to predict liquid densities [132]:

\[
V_s = \left( \frac{R T_c}{P_c} \right) Z_{RA}^{1+0(T/T_c)^{3/2}}
\]  
(E.20)

where 
\[
Z_{RA} = 0.29056 - 0.08775\omega
\]

Where an experimental value of liquid density was known for a given temperature then the density at other temperatures was predicted using:

\[
V_s = V_s^{0.5} Z'_{RA}
\]  
(E.21)
where

\[ J = (1 - T_r)^{\frac{37}{2}} - (1 - T_r^{\frac{57}{8}})^{\frac{27}{2}} \]  \hspace{1cm} (E.22)

and where

- \( V_s \) = saturated liquid volume (cm\(^3\).mol\(^{-1}\))
- \( V_{SR} \) = saturated liquid volume at reference temperature (cm\(^3\).mol\(^{-1}\))
- \( P_c \) = critical pressure (bar)
- \( T_c \) = critical temperature (K)
- \( T_{R'} \) = reduced reference temperature
- \( R' \) = gas constant = 83.144 bar.cm\(^3\).mol\(^{-1}\).K\(^{-1}\)
- \( \omega \) = acentric factor

### E.7 Liquid surface tension

The surface tension correlation used was that of Brock and Bird [133] and Miller [134] which is applicable to non-polar liquids only:

\[ \sigma = P_c^{23} T_c^{-1/3} Q (1 - T_r)^{1199} \]  \hspace{1cm} (E.23)

\[ Q = 0.1196 \left[ 1 + \frac{T_{Br} \ln(P_c/1.01325)}{1 - T_{Br}} \right] - 0.279 \]  \hspace{1cm} (E.26)

where

- \( \sigma \) = surface tension (dyne.cm\(^{-1}\) = 10\(^{-3}\) N.m\(^{-1}\))
- \( P_c \) = critical pressure (bar)
- \( T_c \) = critical temperature (K)
- \( T_r \) = reduced temperature
- \( T_{Br} \) = reduced boiling point
E.8 Data and empirical coefficients for use with physical property correlations

The following lists were drawn up for the hazardous substances considered in Section 10. All were compiled from Reid et al. [128] except where indicated.

E.8.1 Hydrogen (H₂)

Molecular weight  2 kg.kmol⁻¹
Boiling point at atmospheric pressure  20.28 K
Liquid density (at boiling point)  70 kg.m⁻³

Wagner vapour pressure coefficients (bar, K)

<p>| | | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
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<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>-5.57929</td>
<td></td>
<td>Range: 14 K - T_c</td>
</tr>
<tr>
<td>B</td>
<td>2.60012</td>
<td></td>
<td></td>
</tr>
<tr>
<td>C</td>
<td>-0.85506</td>
<td></td>
<td></td>
</tr>
<tr>
<td>D</td>
<td>1.70503</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Coefficients for Cₚₒₘ correlation

<p>| | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>27.14</td>
<td></td>
</tr>
<tr>
<td>B</td>
<td>9.274 x 10⁻³</td>
<td></td>
</tr>
<tr>
<td>C</td>
<td>-1.381 x 10⁻⁵</td>
<td></td>
</tr>
<tr>
<td>D</td>
<td>7.645 x 10⁻⁹</td>
<td></td>
</tr>
</tbody>
</table>

Latent heat of vaporisation [135] 451.9 kJ.K⁻¹
LEL [4] 3.3 g.m⁻³ = 4.0% vol
UEL [4] 64.0 g.m⁻³ = 75.6% vol
Critical temperature T_c 33.25 K
Critical pressure P_c 12.9 bar
Critical volume V_c 64.3 cm³.mol⁻¹
Acentric factor ω 0.216
Dipole movement φ 0.0 debye
Polar correction factor κ n/a
E.8.2 Methane (CH₄)

Molecular weight 16 kg.kmol⁻¹
Boiling point at atmospheric pressure 112 K
Liquid density (at boiling point) 424 kg.m⁻³

Wagner vapour pressure coefficients (bar, K)
| A     | -6.00435 | Range: 91 K - Tₑ |
| B     | 1.18850  |
| C     | -0.83408 |
| D     | -1.22833 |

Coefficients for Cₚ₀ correlation
| A     | 19.25    |
| B     | 5.213 x 10⁻² |
| C     | 1.197 x 10⁻⁵ |
| D     | -1.132 x 10⁻⁹ |

Latent heat of vaporisation [135] 577.4 kJ.K⁻¹
LEL [4] 4.9% vol
UEL [4] 15.9% vol
Critical temperature Tₑ 190.65 K
Critical pressure Pₑ 46.3 bar
Critical volume Vₑ 99.2 cm³.mol⁻¹
Acentric factor ω 0.011
Dipole movement φ 0.0 debye
Polar correction factor κ n/a

Coefficients for liquid viscosity correlation
| A     | -26.87   | Range: 93 - 357 K |
| B     | 1150.0   | Reference value: |
| C     | 0.1871   | 0.14 cP at 103 K  |
| D     | -5.211 x 10⁻⁴ |            |
E.8.3 Ethylene (C₂H₄)

Molecular weight                       28.05 kg.kmol⁻¹
Boiling point at atmospheric pressure  169.29 K
Liquid density (at boiling point)      569.6 kg.m⁻³

Wagner vapour pressure coefficients (bar, K)
A  -6.32055
B   1.16819
C  -1.55935
D -1.83552

Coefficients for Cₚ₀ correlation
A     3.806
B    0.1566
C -8.348 × 10⁻⁵
D   1.755 × 10⁻⁸

Latent heat of vaporisation [135]       515.2 kJ.K⁻¹
LEL [4]                  31 g.m⁻³ = 2.7% vol
UEL [4]                     390 g.m⁻³ = 34% vol
Critical temperature Tᵣ         282.9 K
Critical pressure Pₑ             50.5 bar
Critical volume Vₑ             130.4 cm³.mol⁻¹
Acentric factor ω               0.089
Dipole movement φ                 0.0 debye
Polar correction factor κ           0.0

Coefficients for liquid viscosity correlation
A    -17.74
B     1078.0
C    0.08577
D -1.758 × 10⁻⁴

Range: 104 - 282 K
Reference value:
0.031 cP at 273 K
E.8.4 Propane (C₃H₈)

Molecular weight 44.08 kg.kmol⁻¹
Boiling point at atmospheric pressure 231.1 K
Liquid density (293 K) [135]
(231 K)
        500.5 kg.m⁻³
        582 kg.m⁻³

Wagner vapour pressure coefficients (bar, K)
A  -6.72219 Range: 145 K - Tₑ
B   1.33236
C  -2.13868
D  -1.38551

Coefficients for Cₚₑ correlation
A  -4.224
B   0.3063
C  -1.586 x 10⁻⁴
D   3.215 x 10⁻⁸

Liquid specific heat (200 K) [135]
(298 K) [135] 1.196 kJ.kg⁻¹.K⁻¹
Latent heat of vaporisation [135] 1.652 kJ.kg⁻¹.K⁻¹
LEL [4] 458 kJ.K⁻¹
UEL [4] 39 g.m⁻³ = 2.0% vol
Critical temperature Tₑ 369.6 K
Critical pressure Pₑ 42.5 bar
Critical volume Vₑ 203 cm³.mol⁻¹
Acentric factor ω 0.153
Dipole movement φ 0.0 debye
Polar correction factor κ n/a

Coefficients for liquid viscosity correlation
A  -7.764 C  2.381 x 10⁻² Range: 86 - 369 K
B   721.9 D  -4.665 x 10⁻⁴ Reference value: 0.091 cP at 298 K

E9
E.8.5 Acetone (C₃H₆O)

Molecular weight
Boiling point at atmospheric pressure
Liquid density (293 K)

Wagner vapour pressure coefficients (bar, K)
A -7.45514
B 1.20200
C -2.43926
D -3.35590

Coefficients for Cₚₒ correlation
A 6.301
B 0.2606
C -1.253 × 10⁻⁴
D 2.303 × 10⁻⁸

Liquid specific heat (298 K) [135] 2.176 kJ.kg⁻¹.K⁻¹
Latent heat of vaporisation [135] 550.5 kJ.K⁻¹
LEL [4] 60 g.m⁻³ = 2.15% vol
UEL [4] 310 g.m⁻³ = 13% vol
Critical temperature Tₑ 508.5 K
Critical pressure Pₑ 47 bar
Critical volume Vₑ 209 cm³.mol⁻¹
Acentric factor ω 0.304
Dipole movement φ 2.9 debye
Polar correction factor κ n/a

Coefficients for liquid viscosity correlation
A -4.0033
B 845.6

Range: 193 - 333 K
Reference value:
0.32 cP at 298 K
E.8.6 Methyl Ethyl Ketone (MEK) (C₄H₈O)

Molecular weight 72.107 kg.kmol⁻¹
Boiling point at atmospheric pressure 352.6 K
Liquid density (293 K) 805.4 kg.m⁻³

Wagner vapour pressure coefficients (bar, K)
A -7.71476 Range: 255 K - T_e
B 1.71061
C -3.68770
D -0.75169

Coefficients for \( C_P \) correlation
A 10.94
B 0.3559
C \(-1.9 \times 10^{-4}\)
D \(3.92 \times 10^{3}\)

Liquid specific heat (298 K) [135] 2.211 kJ.kg⁻¹.K⁻¹
Latent heat of vaporisation [135] 473.5 kJ.K⁻¹
LEL [4] 50 g.m⁻³ = 1.8% vol
UEL [4] 350 g.m⁻³ = 11.5% vol
Critical temperature \( T_e \) 537 K
Critical pressure \( P_e \) 41 bar
Critical volume \( V_e \) 267 cm³.mol⁻¹
Acentric factor \( \omega \) 0.320
Dipole movement \( \varphi \) 3.3 debye
Polar correction factor \( \kappa \) n/a

Coefficients for liquid viscosity correlation
A -4.213 Range: 273 - 353 K
B 975.9 Reference value:
0.42 cP at 294 K
Appendix F. Complete listing of fire and explosion model program IGNITION.
1,2589

COMPILATION LISTING OF ignition (<user_dir_dir>/ZMC/AWCox/firex/igniti
\con.fortran)

Compiled by: Multics New Fortran Compiler, Release 11.0
Compiled on: 03/16/89 2239.3 gmt Thu
Options: ansi77 binary_final_point table subscription string
\range round check_multiply card map

Main Program

1 cccccccigniter
2 character*1 quer1,quer2
3 character*12 type(3),ftype(2)
4 real d(4),len(4),f1a(4),val(4),gas(4),tph(4),liq(4),pfail(4,3),
5 *vfail(4,3),ffail(4,2),
6 *parea(4,3),varea(4,3),farea(4,2),pmass(4,3),plmass(4,3),
7 *vmass(4,3),vmaas(4,3),vmaas(4,3),vmass(4,3),vmaas(4,3),vmaas(4,3),vmass(4,3),
8 *fmass(4,3),pgmass(4,3),plmass(4,3),
9 *pignlp(4,3),vignlp(4,3),vigngp(4,3),vign2p(4,3),
10 *pexgp(4,3),vexgp(4,3),pex2p(4,3),
* \c2), ,
11 *vff(4,3),plff(4,3),vfff(4,3),pgff(4,3),vfff(4,3),pfff(4,3),
12 *gfff(4,3),vfff(4,3),vfff(4,3),vfff(4,3),vfff(4,3),
13 *pgfff(4,3),plfff(4,3),
14 *plfff(4,3),plfff(4,3),vfff(4,3),vfff(4,3),vfff(4,3),
15 *vfff(4,3),vfff(4,3),vfff(4,3),
16 *vfff(4,3),vfff(4,3),vfff(4,3),
\real
17 (4) ,lffmin,lffmed,lffmax,lff10min,lff10med,lff10max,
18 *lff10max,lff10med,lff10min,lffmax,lffmed,lffmin,
19 *real mw,lff10,lff10min,lffmax,lffmax,lff10max,lff10med,lff10min,
20 *real mov(4),mfailp(4,3),mgfailp(4,3),m2failp(4,3),
21 *mfailmin,mfailmax,mfailmed,mffmin,mffmax,mffmed,mffmed,
22 *mmax,mmfailp(4,3),mfailp(4,3),mfailp(4,3),mfailp(4,3),
23 *mfailmin,mfailmax,mfailmed,mffmin,mffmax,mmfailp(4,3),
24 *mffmin,mffmax,mffmed,mffmed,mffmed,mffmed,mffmed,mffmed,
25 *mffmed,mffmax,mffmed,mffmed,mffmed,mffmed,mffmed,mffmed,
26 *mffmed,mffmax,mffmed,mffmed,mffmed,mffmed,mffmed,mffmed,
27 *mmax,mincomin,mmfailp(4,3),mfailp(4,3),mfailp(4,3),
28 *mmax,mmfailmax,mfailmax,mfailmax,mfailmax,mfailmax,mfailmax,mfailmax,
29 *mmax,mmfailmax,mfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
30 *mmax,mmfailmax,mfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
31 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
32 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
33 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
34 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
35 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
36 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
37 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
38 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
39 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
40 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
41 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
42 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
43 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
44 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
45 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
46 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
47 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
48 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
49 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
50 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
51 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
52 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
53 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
54 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
55 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
56 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
57 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
58 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
59 *mmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,mmfailmax,
print*,"Leak ignition probability for gas flow = 100 kg.s-1 ?"
read*,aa2
ccccccA2, B2 ETC. ARE CONSTANTS FOR USE WITH IGNITION/EXPLOSION
ccccccPROBABILITY CALCULATION
b2=(log10(aa2/aa1))/(log10(200.0))
a2=aa2/(100**b2)
print*,"Leak ignition probability for liquid flow = 0.5 kg.s-1"
read*,aa3
print*,"Leak ignition probability for liquid flow = 100 kg.s-1"
read*,aa4
b1=(log10(aa4/aa3))/(log10(200.0))
a1=aa4/(100**b1)
print*,"Leak explosion probability for vapour flow = 0.5 kg.s-1"
read*,aa5
print*,"Leak explosion probability for vapour flow = 100 kg.s-1"
read*,aa6
b3=(log10(aa6/aa5))/(log10(200.0))
a3=aa6/(100**b3)
if (quer1.eq."n") goto 10
print*,"Summary results only ? y or n"
read*,quer1
if (quer1.eq."n") goto 10
print*,"Analyse plant ? y or n"
read*,quer1
print*,""
3900 continue
ccccccFLANT 1 PROFILE ARRAYS*******************************
ccccD(4)=PIPE DIAMETER,DSMA=SMALL CONNECTIONS DIAMETER,B(4)=NUMBER 0
ccccccBOLTS IN FLANGED JOINTS FOR MAJOR FLANGE LEAKS, LEN(4)=PIPE LEN
ccccccFLA(4)=FLANGES, VAL(4)=VALVES, MOV(4)=MOVERS I.E. PUMPS, GAS(4)
ccccccLIQ (4),TPH (4)=GAS: LIQUID:2-PHASE SPLIT IN PLANT. SPECIAL SPLITS
ccccccUSED FOR PUMPS (MGAS ETC) AND SMALL CONNECTIONS (SGAS ETC)
d(1)=0.025
d(2)=0.050
d(3)=0.100
d(4)=0.300
disma=0.01
b(1)=4
b(2)=8
b(3)=12
b(4)=20
len(1)=3750
len(2)=4200
len(3)=5400
len(4)=1650
fla(1)=900
fla(2)=1070
fla(3)=810
fla(4)=220
val(1)=720
val(2)=500
val(3)=240
val(4)=40
mov(1)=0
mov(2)=15
mov(3)=10
mov(4)=0
sma=700
gas(1)=0
gas(2) = .15
gas(3) = .15
gas(4) = .25
tph(1) = .5
tph(2) = .45
tph(3) = .45
tph(4) = .4
liq(1) = .5
liq(2) = .4
liq(3) = .4
liq(4) = .35
mgas = 0
mliq = 0
m2ph = 0
sliq = .4
s2ph = .45
sgas = .15

130 exception
131 exception
132 pfail(1,1) = 1e-06
133 pfail(1,2) = 1e-05
134 pfail(1,3) = 1e-04
135 pfail(2,1) = 1e-06
136 pfail(2,2) = 1e-05
137 pfail(2,3) = 1e-04
138 pfail(3,1) = 3e-07
139 pfail(3,2) = 3e-06
140 pfail(3,3) = 3e-05
141 pfail(4,1) = 1e-07
142 pfail(4,2) = 1e-06
143 pfail(4,3) = 1e-05
144 do 5200, i = 1, 4
145 do 5250, j = 1, 3
146 pfail(i, j) = 0.5 * pfail(i, j)
147 5250 continue
148 5200 continue
149 exception
150 exception
151 vfail(1,1) = 3e-06
152 vfail(1,2) = 3e-05
153 vfail(1,3) = 3e-04
154 vfail(2,1) = 3e-06
155 vfail(2,2) = 3e-05
156 vfail(2,3) = 3e-04
157 vfail(3,1) = 3e-06
158 vfail(3,2) = 3e-05
159 vfail(3,3) = 3e-04
160 vfail(4,1) = 3e-06
161 vfail(4,2) = 3e-05
162 vfail(4,3) = 3e-04
163 do 5350, j = 1, 3
164 vfail(1, j) = (1.0/3.0) * vfail(1, j)
165 vfail(2, j) = (1.0/3.0) * vfail(2, j)
166 vfail(3, j) = (1.0/3.0) * vfail(3, j)
167 vfail(4, j) = (1.0/6.0) * vfail(4, j)
168 5350 continue
169 exception
170 exception
171 ffail(1,1) = 1e-05
172 ffail(1,2) = 3e-05
173 ffail(2,1) = 1e-05
174 ffail(2,2) = 3e-05
175 ffail(3,1) = 1e-05
176 ffail(3,2) = 3e-05
177 do 5400, i = 1, 4
178 ffail(1, 1) = 3 * ffail(1, 1)
ffail(i,2) = 10 * ffail(i,2)

5400 continue
ccccccMOVERS

mfail(1,1) = 1e-04
mfail(1,2) = 1e-03
mfail(1,3) = 1e-02
mfail(2,1) = 1e-04
mfail(2,2) = 1e-03
mfail(2,3) = 1e-02
mfail(3,1) = 1e-04
mfail(3,2) = 1e-03
mfail(3,3) = 1e-02
mfail(4,1) = 1e-04
mfail(4,2) = 1e-03
mfail(4,3) = 1e-02

do 6500, i = 1, 4
  do 6550, j = 1, 3
    mfail(i, j) = 0.3 * mfail(i, j)
  enddo
6550 continue
6500 continue

ccccccCONNECTIONS

sfail(1) = 0.0001
sfail(2) = 0.003

cccccc**********CHARACTER TYPE ARRAY**********

type(1) = "Rupture leak"
type(2) = "Major leak"
type(3) = "Minor leak"
ftype(1) = "Major leak"
ftype(2) = "Minor leak"

cccccc**********PHYSICAL PROPERTIES OF GAS, LIQUID**********

gamma = 1.4
pw = 44.08
pl = 1.5e06
t = 300
rhol = 600
pi = 3.14159
r = 8314.4
cd = 0.6

kliq = sqrt(2 * (pi - 1e05) * rhol)
kgas = pi * sqrt((gamma * mw / (r * t * t)) * (2 / (gamma + 1))) / ((gamma + 1) / (gamma - 1))

cccccc**********COMPONENT FAILURE/PLANT/YEAR**********

ccccccARRAY ELEMENT LABELS ARE DERIVED AS FOR THE FOLLOWING EXAMPLE:

ccccccPGFAILP-PIPE-GAS-FAILURE FREQUENCY-PERPLANT

ccccccPIPEWORK

do 1500, i = 1, 4
  do 1550, j = 1, 3
    pgfailp(i, j) = len(i) * pfail(i, j) * gas(i)
  enddo
1550 continue
1500 continue
ccccccVALVES

do 1600, i = 1, 4
  do 1650, j = 1, 3
    vfailp(i, j) = val(i) * vfail(i, j) * gas(i)
  enddo
1650 continue
1600 continue
ccccccMOVERS

if (i.eq.2) then
  mliq = 0.733333
  m2ph = 0.266667
endif
if (i.eq.3) then
  mliq = 0.8
endif

F4
m2ph=0.4

endif

mmliq(i)=mliq*mov(i)

mm2ph(i)=m2ph*mov(i)

mgfailp(i,j)=mov(i)*mfail(i,j)*mgas

mlfailp(i,j)=mov(i)*mfail(i,j)*mliq

m2failp(i,j)=mov(i)*mfail(i,j)*m2ph

continue

continu

ccccccFLANGES

do 1700, i=1,4
do 1750, j=1,2
fgfailp(i,j)=fla(i)*ffail(i,j)*gas(i)
flfailp(i,j)=fla(i)*ffail(i,j)*liq(i)
f2failp(i,j)=fla(i)*ffail(i,j)*tph(i)

continue

continue

ccccccCONNECTIONS

do 6900, i=1,2

sgfailp(i)=sma*sfail(i)*sgas
 slfailp(i)=sma*sfail(i)*sliq
 s2failp(i)=sma*sfail(i)*s2ph

continue

cccccc************LEAK HOLE SIZE ARRAY***********************

ccccccPIES

do 1800, i=1,4

d0 1950, j=1,3

parea(i,1)=0.25*pi*d(i)**2
parea(i,2)=0.025*pi*d(i)**2
parea(i,3)=0.0025*pi*d(i)**2
continue

continue

ccccccFLANGE GASKETS

do 2000, i=1,4

farea(i,1)=pi*0.004*d(i)/b(i)
farea(i,2)=pi*0.004*d(i)/(10*b(i))

continue

ccccccCONNECTIONS

sarea(1)=0.25*pi*dsma**2
sarea(2)=0.140.25*pi*dsma**2

cccccc**********MASS FLOW THROUGH LEAK HOLES******************

ccccccVALVES

do 2100, i=1,4

varea(i,1)=v1*0.004*d(1)/b(1)
varea(i,2)=v1*0.004*d(1)/(10*b(1))

continue

continue

ccccccPIES

d0 2200, i=1,4

do 2250, j=1,3

vgmass(i,j)=cd*kgas*varea(i,j)
vlmass(i,j)=cd*klid*vearea(i,j)
v2mass(i,j)=cd*0.5*klid*vearea(i,j)

continue

continue

ccccccFOR GUILLOTINE FRACTURE PIPE Cd=4/3; ORIFICE Cd

if (j.eq.1) then

pgmass(i,j)=1.5*1.33333*pgmass(i,j)

end if
\[
\text{plmass}(i,j) = 1.5 \times 1.33333 \times \text{plmass}(i,j)
\]
\[
\text{p2mass}(i,j) = 1.5 \times 1.33333 \times \text{p2mass}(i,j)
\]
\[
\text{mlmass}(i,j) = \text{cd} \times \text{kgas} \times \text{marea}(i,j)
\]
\[
\text{m2mass}(i,j) = \text{cd} \times 0.8 \times \text{klq} \times \text{marea}(i,j)
\]
\[
\text{mgmass}(i,j) = \text{cd} \times \text{kgas} \times \text{marea}(i,j)
\]
\[
\text{mlmass}(i,j) = \text{cd} \times \text{klq} \times \text{marea}(i,j)
\]
\[
\text{m2mass}(i,j) = \text{cd} \times 0.8 \times \text{klq} \times \text{marea}(i,j)
\]
\[
\text{cc:ccc:cc:MOVERS}
\]
\[
\text{mgmass}(i,j) = \text{cd} \times \text{kgas} \times \text{marea}(i,j)
\]
\[
\text{mlmass}(i,j) = \text{cd} \times \text{klq} \times \text{marea}(i,j)
\]
\[
\text{m2mass}(i,j) = \text{cd} \times 0.8 \times \text{klq} \times \text{marea}(i,j)
\]
\[
\text{cc:ccc:cc:FLANGES}
\]
\[
\text{fgmass}(i,j) = \text{cd} \times \text{kgas} \times \text{farea}(i,j)
\]
\[
\text{flmass}(i,j) = \text{cd} \times \text{klq} \times \text{farea}(i,j)
\]
\[
\text{f2mass}(i,j) = \text{cd} \times 0.8 \times \text{klq} \times \text{farea}(i,j)
\]
\[
\text{cc:ccc:cc:CONNECTIONS}
\]
\[
\text{sgmass}(i) = \text{cd} \times \text{kgas} \times \text{sarea}(i)
\]
\[
\text{slmass}(i) = \text{cd} \times \text{klq} \times \text{sarea}(i)
\]
\[
\text{s2mass}(i) = 0.8 \times \text{cd} \times \text{klq} \times \text{sarea}(i)
\]
\[
\text{if (i.eq.1)} \text{then}
\]
\[
\text{sgmass}(i) = \frac{4.0}{3.0} \times \text{sgmass}(i)
\]
\[
\text{slmass}(i) = \frac{4.0}{3.0} \times \text{slmass}(i)
\]
\[
\text{s2mass}(i) = \frac{4.0}{3.0} \times \text{s2mass}(i)
\]
\[
\text{cc:ccc:cc:SMALL CONNECTION CD IS TAKEN TO BE 4/3 ORIFICE CD}
\]
\[
\text{cc:cccccc:ARRAY ELEMENT LABELS DERIVED AS FOR EXAMPLE:}
\]
\[
\text{fignlp}=\text{PIPE-IGNITION-FLUID FLOW-PROBABILITY}
\]
\[
\text{cc:cccccc:PIPES AND VALVES - LIQUID, GAS, 2-PHASE}
\]
\[
\text{do 2400, i=1,4}
\]
\[
\text{do 2450, j=1,3}
\]
\[
\text{pignlp}(i,j) = a1 \times \text{plmass}(i,j)^{b1}
\]
\[
\text{vignlp}(i,j) = a1 \times \text{vlmass}(i,j)^{b1}
\]
\[
\text{pigngp}(i,j) = a2 \times \text{pgmass}(i,j)^{b2}
\]
\[
\text{vigngp}(i,j) = a2 \times \text{vgmass}(i,j)^{b2}
\]
\[
\text{pign2p}(i,j) = a2 \times \text{p2mass}(i,j)^{b2}
\]
\[
\text{vign2p}(i,j) = a2 \times \text{v2mass}(i,j)^{b2}
\]
\[
\text{if (pignlp(i,j).gt.1) then}
\]
\[
\text{pignlp}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (vignlp(i,j).gt.1) then}
\]
\[
\text{vignlp}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (pigngp(i,j).gt.1) then}
\]
\[
\text{pigngp}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (vigngp(i,j).gt.1) then}
\]
\[
\text{vigngp}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (pign2p(i,j).gt.1) then}
\]
\[
\text{pign2p}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (vign2p(i,j).gt.1) then}
\]
\[
\text{vign2p}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{cc:cccccc:ALSO MOVERS}
\]
\[
\text{mignlp}(i,j) = a1 \times \text{mlmass}(i,j)^{b1}
\]
\[
\text{migngp}(i,j) = a2 \times \text{mgmass}(i,j)^{b2}
\]
\[
\text{mign2p}(i,j) = a2 \times \text{m2mass}(i,j)^{b2}
\]
\[
\text{if (mignlp(i,j).gt.1) then}
\]
\[
\text{mignlp}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (migngp(i,j).gt.1) then}
\]
\[
\text{migngp}(i,j) = 1
\]
\[
\text{endif}
\]
\[
\text{if (mign2p(i,j).gt.1) then}
\]
\[
\text{mign2p}(i,j) = 1
\]
\[
\text{endif}
\]

F6
if (migngp(i,j).gt.1) then
  migngp(i,j)=1
endif
if (mign2p(i,j).gt.1) then
  mign2p(i,j)=1
endif
continue
2450
continue
ccccccFLANGES - LIQUID, GAS, 2-PHASE
do 2500, i=1,4
do 2550, j=1,2
fignlp(i,j)=a1*flmass(i,j)**b1
figngp(i,j)=a2*fgmass(i,j)**b2
fign2p(i,j)=a2*f2mass(i,j)**b2
if (fignlp(i,j).gt.1) then
  fignlp(i,j)=1
endif
if (figngp(i,j).gt.1) then
  figngp(i,j)=1
endif
if (fign2p(i,j).gt.1) then
  fign2p(i,j)=1
endif
2550
continue
2500
continue
ccccccCONNECTIONS
do 7100, i=1,2
signlp(i)=a1*slmass(i)**b1
signgp(i)=a2*sgmass(i)**b2
sign2p(i)=a2*s2mass(i)**b2
7100
continue
cccccc**********GAS AND 2-PHASE EXPLOSION PROBABILITIES**********
ccccccARRAY ELEMENT LABELS DERIVED AS FOR EXAMPLE:
cccccP<EXG=PIPE-EXPLOSION-GAS-PROBABILITY
cccccPIF<EXG=PIPE-EXPLOSION-FRAGILE-PROBABILITY
cccccV<EXG=VALVE-EXPLOSION-FRAGILE-PROBABILITY
cccccM<EXG=MOVER-EXPLOSION-FRAGILE-PROBABILITY

cccc<ESTABLISH PIPE/VALVE/MOVER FIRE/EXPLOSION CUTOFFS**********

if (pigngp(i,j).lt.0.5) then
  pigngp(i,j)=aal
endif
if (pign2p(i,j).lt.0.5) then
  pign2p(i,j)=aal
endif
if (pignlp(i,j).lt.0.5) then
  pignlp(i,j)=aal
endif

if (vgrnass(i,j).lt.5) then
    vigngp(i,j)=aa1
endif
if (v2mass(i,j).lt.5) then
    vign2p(i,j)=aa1
endif
if (v1mass(i,j).lt.5) then
    vignlp(i,j)=aa3
endif
if (pgmass(i,j).gt.100) then
    pigngp(i,j)=aa2
    pexgp(i,j)=aa6
endif
if (p2mass(i,j).gt.100) then
    pign2p(i,j)=aa2
    peN2p(i,j)=aa6
endif
if (plmass(i,j).gt.1(0) then
    pignlp(i,j)=aa4
endif
if (vgrnass(i,j).gt.1(0) then
    vigngp(i,j)=aa2
    ve>:gp (i,j)=o.a6
endif
if (v2mass(i,j).gt.1(0) then
    vign2pCi,j)=aa2
    ve:{2p (i ,j) =a.a6
endif
if (vlmass(i,j).gt.1(0) then
    vignlp(i,j)=aa4
endif
if (mgmass<i,j).lt.O.5) then
    migngp(i,j)=o.a,l
endif
if (m2mass(i,j).lt.O.5) then
    mign12p(i,j) :=aa1
endif
if (mlmass(i,j).lt .. .0.5) then
    mignlp(i,j) =aa3
endif
if (mgmass(i,j).gt.100) then
    migngp(i,j)=aa2
    mexgp(i,j)=aa6
endif
if (m2mass(i,j).gt.100) then
    mign2p(i,j)=aa2
    me}: 2p (i ,j ) =aa6
endif
if (mlmass(i ,j) .gt.1(0) then
    mignlp<i,j)=aa4
endif
ccccccccEXPOSION PROBABILITY GIVEN LEAK = EXPLOSION PROBABILITY
cccccccccGIVEN IGNITION X IGNITION PROBABILITY GIVEN LEAK
pexgp(i,j)=pigngp(i,j)*pexgp(i,j)
pex2p(i,j)=pign2p(i,j)*pex2p(i,j)
pex1p(i,j)=vignlp(i,j)*pex2p(i,j)
mxgp(i,j)=migngp(i,j)*mxgp(i,j)
mx2p(i,j)=mign2p(i,j)*mx2p(i,j)
mx1p(i,j)=mignlp(i,j)*mx2p(i,j)

2650 continue
2600 continue
cccccccFLANGES
d0 2700, i=1,4
do 2750, j=1,2
fxgp(i,j)=a3*fgmass(i,j)**b3
fx2p(i,j)=a3*2mass(i,j)**b3
if (fexgp(i,j).gt.l) then
  fexgp(i,j)=1
endif

if (fex2p(i,j).gt.l) then
  fex2p(i,j)=1
endif

ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
ESTABLISH FLANGE FIRE/EXPLOSIONS CUTOFF
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
if (fgmass(i,j).lt.0.5) then
  figngp(i,j)=aa1
endif
if (fg2mass(i,j).lt.0.5) then
  fign2p(i,j)=aa1
endif
if (f1mass(i,j).lt.0.5) then
  fignlp(i,j)=aa3
endif
if (f2mass(i,j).lt.0.5) then
  figngp(i,j)=aa4
endif

fexgp(i,j)=figngp(i,j)*fexgp(i,j)
fex2p(i,j)=fign2p(i,j)*fex2p(i,j)

2750 continue

ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
CONNECTIONS
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
do 7200, i=1,2
sexgp(i)=a3*sgmass(i)**b3
sex2p(i)=a3*s2mass(i)**b3
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
CONNECTIONS FIRE/EXPLOSIONS CUTOFF
if (s1mass(i).lt.0.5) then
  signlp(i)=aa3
endif
if (s1mass(i).gt.100) then
  signlp(i)=aaA
endif
if (sgmass(i).lt.0.5) then
  signgp(i)=aa1
endif
if (sgmass(i).gt.100) then
  signgp(i)=aa2
endif
if (s2mass(i).lt.0.5) then
  sign2p(i)=aa1
endif
if (s2mass(i).gt.100) then
  sign2p(i)=aa2
endif
sexgp(i)=signgp(i)*sexgp(i)
sex2p(i)=sign2p(i)*sex2p(i)

7200 continue

ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
LEAK FIRE FREQUENCIES
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
ARRAY ELEMENT LABELLED DERIVED AS FOR EXAMPLE:
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
VGFF=VALVE-GAS-FIRE FREQUENCY
ccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
do 2800, i=1,4
do 2850, j=1,3
vlff(i,j)=vignlp(i,j)*vfailp(i,j)
\[
\begin{align*}
\text{f}(i, j) &= \text{sign}(p(i, j)) * p(i, j) \\
\text{vgff}(i, j) &= \text{vign}(p(i, j)) * \text{vgfail}(p(i, j)) \\
v2ff(i, j) &= \text{vign}(p(i, j)) * v2fail(p(i, j)) \\
p2ff(i, j) &= \text{vign}(p(i, j)) * p2fail(p(i, j)) \\
\text{mlff}(i, j) &= \text{mign}(p(i, j)) * \text{mfail}(p(i, j)) \\
\text{mgff}(i, j) &= \text{mign}(p(i, j)) * \text{mgfail}(p(i, j)) \\
m2ff(i, j) &= \text{mign}(p(i, j)) * m2fail(p(i, j)) \\
\text{flff}(i, j) &= \text{fign}(p(i, j)) * \text{flfail}(p(i, j)) \\
f2ff(i, j) &= \text{fign}(p(i, j)) * f2fail(p(i, j)) \\
m2ff(i, j) &= \text{mign}(p(i, j)) * m2fail(p(i, j)) \\
\text{sgff}(i) &= \text{slgn}(p(i)) * \text{sgfail}(p(i)) \\
\text{slff}(i) &= \text{slgn}(p(i)) * \text{slfail}(p(i)) \\
s2ff(i) &= \text{slgn}(p(i)) * s2fail(p(i)) \\
\text{vge}(i, j) &= \text{veg}(p(i, j)) * \text{vgfail}(p(i, j)) \\
v2exf(i, j) &= \text{veg}(p(i, j)) * v2fail(p(i, j)) \\
p2exf(i, j) &= \text{veg}(p(i, j)) * p2fail(p(i, j)) \\
m2exf(i, j) &= \text{veg}(p(i, j)) * m2fail(p(i, j)) \\
\text{flge}(i, j) &= \text{feg}(p(i, j)) * \text{flfail}(p(i, j)) \\
f2exf(i, j) &= \text{feg}(p(i, j)) * f2fail(p(i, j)) \\
m2exf(i, j) &= \text{feg}(p(i, j)) * m2fail(p(i, j)) \\
\text{sgexf}(i) &= \text{sex}(p(i)) * \text{sgfail}(p(i)) \\
s2exf(i) &= \text{sex}(p(i)) * s2fail(p(i)) \\
\text{m2failmin} &= 0 \\
\text{m2failmax} &= 0
\end{align*}
\]
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754  ccccc cov denotes overall totals. Firstly for phases, then for phases/ ccccc equipment, then finally for equipment
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758  vfailov=0
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805 flfov=0
806 fgefov=0
807 f2fov=0
808 mffov=0
809 m2ffov=0
810 sffov=0
811 s2ffov=0
812 slfov=0
813 s2ffov=0
814 ccccccPIPECs, VALVES AND MOVERS
815 ccccccLIQUID - FIRE
816 do 3200, i=1,4
817 do 3250, j=1,3
818 ccccccPIPECs
819 if (plmass(i,j).le.1.0) then
820 plffmin=plffmin+plff(i,j)
821 plfailmin=plfailmin+plfailp(i,j)
822 else if (plmass(i,j).gt.1.and.plmass(i,j).le.50) then
823 plffmed=plffmed+plff(i,j)
824 plfailmed=plfailmed+plfailp(i,j)
825 else if (plmass(i,j).gt.50) then
826 plffmax=plffmax+plff(i,j)
827 plfailmax=plfailmax+plfailp(i,j)
828 endif
829 ccccccVALVEs
830 if (v1mass(i,j).le.1.0) then
831 v1ffmin=v1ffmin+v1ff(i,j)
832 v1failmin=v1failmin+v1failp(i,j)
833 else if (v1mass(i,j).gt.1.and.v1mass(i,j).le.50) then
834 v1ffmed=v1ffmed+v1ff(i,j)

F13
endif

cccGOVERS

if (mlmass(i,j)<1.e-1) then
  mlffmin=mlffmin+mlff(i,j)
  mlfailmin=mlfailmin+mlfailp(i,j)
else if (mlmass(i,j)<1.and.mlmass(i,j)<1.e-50) then
  mlffmed=mlffmed+mlff(i,j)
  mlfailmed=mlfailmed+mlfailp(i,j)
else if (mlmass(i,j)<50) then
  mlffmax=mlffmax+mlff(i,j)
  mlfailmax=mlfailmax+mlfailp(i,j)
endif

cccGAS AND 2-PHASE - FIRE AND EXPLOSION

ccccVALUES

if (vgmass(i,j)<1.e-1) then
  vgffmin=vgffmin+vgff(i,j)
  vgexfmin=vgexfmin+vgexf(i,j)
  vgfailmin=vgfailmin+vgfailp(i,j)
else if (vgmass(i,j)<1.and.vgmass(i,j)<1.e-50) then
  vgffmed=vgffmed+vgff(i,j)
  vgexfmed=vgexfmed+vgexf(i,j)
  vgfailmed=vgfailmed+vgfailp(i,j)
else if (vgmass(i,j)<50) then
  vgffmax=vgffmax+vgff(i,j)
  vgexfmax=vgexfmax+vgexf(i,j)
  vgfailmax=vgfailmax+vgfailp(i,j)
endif

endf

if (v2mass(i,j)<1.e-1) then
  v2ffmin=v2ffmin+v2ff(i,j)
  v2exfmin=v2exfmin+v2exf(i,j)
  v2failmin=v2failmin+v2failp(i,j)
else if (v2mass(i,j)<1.and.v2mass(i,j)<1.e-50) then
  v2ffmed=v2ffmed+v2ff(i,j)
  v2exfmed=v2exfmed+v2exf(i,j)
  v2failmed=v2failmed+v2failp(i,j)
else if (v2mass(i,j)<50) then
  v2ffmax=v2ffmax+v2ff(i,j)
  v2exfmax=v2exfmax+v2exf(i,j)
  v2failmax=v2failmax+v2failp(i,j)
endif

endf

ccccPIPES

if (pgmass(i,j)<1.e-1) then
  pgffmin=pgffmin+pgff(i,j)
  pgexfmin=pgexfmin+pgexf(i,j)
  pgfailmin=pgfailmin+pgfailp(i,j)
else if (pgmass(i,j)<1.and.pgmass(i,j)<1.e-50) then
  pgffmed=pgffmed+pgff(i,j)
  pgexfmed=pgexfmed+pgexf(i,j)
  pgfailmed=pgfailmed+pgfailp(i,j)
else if (pgmass(i,j)<50) then
  pgffmax=pgffmax+pgff(i,j)
  pgexfmax=pgexfmax+pgexf(i,j)
  pgfailmax=pgfailmax+pgfailp(i,j)
endif

endf

endf
else if (p2mass(i,j) > 50) then
  p2ffmax = p2ffmax + p2ff(i,j)
  p2exfmax = p2exfmax + p2exf(i,j)
  p2failmax = p2failmax + p2failp(i,j)
endif

ccccccMOVERS
if (mgmass(i,j) <= 1.) then
  mgffmin = mgffmin + mgff(i,j)
  mgexfmin = mgexfmin + mgexf(i,j)
  mgfailmin = mgfailmin + mgfailp(i,j)
else if (mgmass(i,j) > 1. and mgmass(i,j) <= 50) then
  mgffmed = mgffmed + mgff(i,j)
  mgexfmed = mgexfmed + mgexf(i,j)
  mgfailmed = mgfailmed + mgfailp(i,j)
else if (mgmass(i,j) > 50) then
  mgffmax = mgffmax + mgff(i,j)
  mgexfmax = mgexfmax + mgexf(i,j)
  mgfailmax = mgfailmax + mgfailp(i,j)
endif

3250 continue
3200 continue
ccccccFLANGES
ccccliquid - fire
do 3300, i = 1, 4
  do 3350, j = 1, 2
    if (flmass(i,j) <= 1.) then
      flffmin = flffmin + flff(i,j)
      flfailmin = flfailmin + flfailp(i,j)
    else if (flmass(i,j) > 1. and flmass(i,j) <= 50) then
      flffmed = flffmed + flff(i,j)
      flfailmed = flfailmed + flfailp(i,j)
    else if (flmass(i,j) > 50) then
      flffmax = flffmax + flff(i,j)
      flfailmax = flfailmax + flfailp(i,j)
    endif
  enddo
  if (f2mass(i,j) <= 1.) then
    f2ffmin = f2ffmin + f2ff(i,j)
    f2exfmin = f2exfmin + f2exf(i,j)
    f2failmin = f2failmin + f2failp(i,j)
  endif
enddo
ccccejgas and 2-phase - fire and explosion
if (fgmass(i,j) <= 1.) then
  fffmin = fffmin + fff(i,j)
  fexfmin = fexfmin + fexf(i,j)
  ffailmin = ffailmin + ffailp(i,j)
else if (fgmass(i,j) > 1. and ffgmass(i,j) <= 50) then
  fffmed = fffmed + fff(i,j)
  fexfmed = fexfmed + fexf(i,j)
  ffailmed = ffailmed + ffailp(i,j)
else if (fgmass(i,j) > 50) then
  fffmax = fffmax + fff(i,j)
  fexfmax = fexfmax + fexf(i,j)
  ffailmax = ffailmax + ffailp(i,j)
endif
if (f2mass(i,j) <= 1.) then
  f2ffmin = f2ffmin + f2ff(i,j)
  f2exfmin = f2exfmin + f2exf(i,j)
  f2failmin = f2failmin + f2failp(i,j)
endif

else if (f2mass(i,j).gt.1.and.f2mass(i,j).le.50) then
  f2ffmed=f2ffmed+f2ff(i,j)
  f2exfmed=f2exfmed+f2exf(i,j)
  f2failmed=f2failmed+f2failp(i,j)
else if (f2mass(i,j).gt.50) then
  f2ffmax=f2ffmax+f2ff(i,j)
  f2exfmax=f2exfmax+f2exf(i,j)
  f2failmax=f2failmax+f2failp(i,j)
endif

if (s1mass(i).le.1) then
  slffmin=slffmin+slff(i)
  slfailmin=slfailmin+slfailp(i)
else if (s1mass(i).gt.1.and.s1mass(i).le.50) then
  slffmed=slffmed+slff(i)
  slfailmed=slfailmed+slfailp(i)
else if (s1mass(i).gt.50) then
  slffmax=slffmax+slff(i)
  slfailmax=slfailmax+slfailp(i)
endif
if (s2mass(i).le.1) then
  s2ffmin=s2ffmin+s2ff(i)
  s2exfmin=s2exfmin+s2exf(i)
  s2failmin=s2failmin+s2failp(i)
else if (s2mass(i).gt.1.and.s2mass(i).le.50) then
  s2ffmed=s2ffmed+s2ff(i)
  s2exfmed=s2exfmed+s2exf(i)
  s2failmed=s2failmed+s2failp(i)
else if (s2mass(i).gt.50) then
  s2ffmax=s2ffmax+s2ff(i)
  s2exfmax=s2exfmax+s2exf(i)
  s2failmax=s2failmax+s2failp(i)
endif
if (s2mass(i).le.1) then
  s2ffmin=s2ffmin+s2ff(i)
  s2exfmin=s2exfmin+s2exf(i)
  s2failmin=s2failmin+s2failp(i)
else if (s2mass(i).gt.1.and.s2mass(i).le.50) then
  s2ffmed=s2ffmed+s2ff(i)
  s2exfmed=s2exfmed+s2exf(i)
  s2failmed=s2failmed+s2failp(i)
else if (s2mass(i).gt.50) then
  s2ffmax=s2ffmax+s2ff(i)
  s2exfmax=s2exfmax+s2exf(i)
  s2failmax=s2failmax+s2failp(i)
endif

print*,"Ignition analysis program - Initial conditions"
do 6000, i=1,5
print*,"Plant operating conditions"
do 6000, i=1,5
print*,"Average plant pressure = ".,p1/1e05," bar
print*, "Average plant temperature = ", t1, " K"
print*, " Fluid properties ------------------
print*, "Molecular weight = ", mw, " kg.kmol-
print*, "Liquid density at plant temperature = ", rhol, " kg.m-3
print*, "Expansion coefficient = ", gamma
print*, "Leak orifice discharge coefficients"
print*, "Plain hole (gasket, seal etc.) Cd = ", cd
print*, "Full bore pipe leak Cd = ", cd*(4.0/3.0)
print*, "Ignition/Explosion probabilities"
print*, "Gas ignition probability at leak rate of 0.5 kg.s-1 = ", aa1
print*, "Gas ignition probability at leak rate of 100 kg.s-1 = ", aa2
print*, "Liquid ignition probability at leak rate of 0.5 kg.s-1 = ", aa3
print*, "Liquid ignition probability at leak rate of 100 kg.s-1 = ", aa4
print*, "Gas/Vapour explosion probability given ignition at leak rate of 0.5 kg.s-1 = ", aa5
print*, "Gas/Vapour explosion probability given leak rate of 100 kg.s-1 = ", aa6
do 6100, i=1,23
continue
if (quer2.eq."y") goto 26
**********************************PRINT PLANT PROFILES*************************************
print*, " Plant leak source and leak rate profile"

print*, " Pipe lengths (m), 5x,f7.1,3x,f7.1,3x,f7.1,3x,f7.1)
print 217, (val(i),i=1,4)
print 218, (fla(i),i=1,4)
print 219, (gas(i),i=1,4)
print 220, (liq(i),i=1,4)
print 221, (tph(i),i=1,4)

format (1h, "Two-phase fraction", 6x, f4.2, 6x, f4.2, 6x, f4.2, 6x, f4.2, 6x, f4.2, 6x, f4.2)

print*, "Pipe lengths (m)"

format (1h, "Gas ", 9x, f7.1, 3x, f7.1, 3x, f7.1, 3x, f7.1)

print*, "Gas

format (1h, "Liquid ", 6x, f7.1, 3x, f7.1, 3x, f7.1, 3x, f7.1)

print*, "Liquid

format (1h, "Two-phase ", 5x, f7.1, 3x, f7.1, 3x, f7.1, 3x, f7.1)

print*, "Two-phase

print*, "Valves"

print*, "Flanges"

print*, "Pumps"

print*, "Small-bore connections"

print*, "Failure frequency profile"
**Small-bore connection leak frequency**

(Leak/connection/year x 10000)

```
format(1h,a12,9x,f7.4)
do 7600, i=1,31
```

```
print*,
```

```
236, (typeCi), (1e04*sfail(i)) ,i=1,2)
```

```
format(lh,a12,9x,f7.4)
do 7600, i=1,31
```

```
print*,
```

```
236, format(lh,a12,9x,f7.4)
```

```
do 3400, i=1,29
```

```
cccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc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1221  print 204, (type(j), (100000*plff(i,j), i=1,4), j=1,3)
1222  do 4500, i=1,32
1223  print*, ""
1224  4500 continue
1225  print*, "Valves"
1226  print*, "--------
1227  print*, ""
1228  print*, "Leak frequency (leaks/plant/year)"
1229  print*, ""
1230  print 201, (d(i), i=1,4)
1231  print*, ""
1232  print 202, (type(j), (vlfailp(i,j), i=1,4), j=1,3)
1233  print*, ""
1234  print*, "Mass flow (kg/s)"
1235  print*, ""
1236  print 203, (type(j), (vlmass(i,j), i=1,4), j=1,3)
1237  print*, ""
1238  print*, "Ignition probability"
1239  print*, ""
1240  print 202, (type(j), (vignlp(i,j), i=1,4), j=1,3)
1241  print*, ""
1242  print*, "Fire frequency (fires/plant/year x 10000)"
1243  print*, ""
1244  print 204, (type(j), (100000*vlff(i,j), i=1,4), j=1,3)
1245  print*, ""
1246  print*, "Leak frequency (leaks/plant/year)"
1247  print*, "--------
1248  print*, ""
1249  print*, "Mass flow (kg/s)"
1250  print*, ""
1251  print 203, (type(j), (vlmass(i,j), i=1,4), j=1,2)
1252  print*, ""
1253  print*, "Ignition probability"
1254  print*, ""
1255  print 202, (type(j), (fignlp(i,j), i=1,4), j=1,2)
1256  print*, ""
1257  print*, "Fire frequency (fires/plant/year x 10000)"
1258  print*, ""
do 6400, i=1,10
1259  print*, ""
1260  6400 continue
1261  print*, "Pumps"
1262  print*, "--------
1263  print*, ""
1264  print 201, (d(i), i=1,4)
1265  print*, ""
1266  print*, "Leak frequency (leaks/plant/year)"
1267  print*, ""
1268  print 202, (type(j), (mlfailp(i,j), i=1,4), j=1,3)
1269  print*, ""
1270  print*, "Mass flow (kg/s)"
1271  print*, ""
1272  print 203, (type(j), (mlmass(i,j), i=1,4), j=1,3)
1273  print*, ""
1274  print*, "Ignition probability"
1275  print*, ""
1276  print 202, (type(j), (mignlp(i,j), i=1,4), j=1,3)
1277  print*, ""
1278  print*, "Fire frequency (fires/plant/year x 10000)"
1279  print*, ""
do 6400, i=1,10
1280  print*, ""
print," "
print,"Fire frequency (fires/plant/year x 10000)"
print," "
print 204, (type(j),(10000*mlff(i,j), i=1,4),j=1,3)
print," "
print,"Small-bore connections"
print," "
print,"Leak frequency (leaks/plant/year)"
print," "
print 237, (type(i),s1failp(i),i=1,2)
print," "
print,"Mass flow (kg.s-1)"
print," "
print 236, (type(i),smass(i),i=1,2)
print," "
print,"Ignition probability"
print," "
print 237, (type(i),signlp(i),i=1,2)
print," "
print,"Fire frequency (fires/plant/year x 10000)"
print," "
print 238, (type(i),slff(i),i=1,2)
print* 
237 format(lh ,a12,9x,f7.5)
238 format(lh ,a12,9%f7.3)
do 4600,i=1,20
print," "
4600 continue
print,"Gas leaks"
print,"=============
print," "
print,"Pipework"
print,"--------
print,"Leak frequency (leaks/plant/year)"
print," "
print 201, (d(i),i=1,4)
print," "
print 202, (type(j),(pgfailp(i,j),i=1,4),j=1,3)
print," "
print,"Mass flow (kg.s-1)"
print," "
print 204, (type(j),(pgmass(i,j),i=1,4),j=1,3)
204 format (lh ,a12,9%f7.3,4x,f7.3,4x,f7.3,4x,f7.3)
print," "
print,"Ignition probability"
print," "
print 202, (type(j),(pigngp(i,j),i=1,4),j=1,3)
print," "
print,"Explosion probability"
print," "
print 202, (type(j),(pexplp(i,j),i=1,4),j=1,3)
print," "
print,"Fire frequency (fires/plant/year - x 10000)"
print," "
print 204, (type(j),(1e04*pgff(i,j),i=1,4),j=1,3)
print," "
print,"Explosion frequency (explosions/plant/year - x 10000)"
print," "
print 204, (type(j),(1e04*pgexf(i,j),i=1,4),j=1,3)
do 4700,i=1,25
print," "
4700 continue
print,"Valves"
print*,-"
print*,
print*,'Leak frequency (leaks/plant/year)'
print*,
print 201, (d(i), i=1,4)
print*,
print 202, (type(j), (vgfailp(i,j), i=1,4), j=1,3)
print*,
print*,'Mass flow (kg.s⁻¹)
print*,
print 204, (type(j), (vgmass(i,j), i=1,4), j=1,3)
print*,
print*,'Ignition probability'
print*,
print 202, (type(j), (vgignp(i,j), i=1,4), j=1,3)
print*,
print*,'Explosion probability'
print*,
print 202, (type(j), (vgexp(i,j), i=1,4), j=1,3)
print*,
print*,'Mass flow (kg.s⁻¹)
print*,
print 204, (type(j), (vgmass(i,j), i=1,4), j=1,3)
print*,
print*,'Ignition probability'
print*,
print 202, (type(j), (vgignp(i,j), i=1,4), j=1,3)
print*,
print*,'Explosion probability'
print*,
print 202, (type(j), (vgexp(i,j), i=1,4), j=1,3)
print*,
print*,'Fire frequency (fires/plant/year - x 10000)'
print*,
print 204, (type(j), (vgff(i,j), i=1,4), j=1,3)
print*,
print*,'Explosion frequency (explosions/plant/year - x 10000)'
do 7900, i=1,22
print*,-"
print*,'Flanges'
print*,-"
print*,'Leak frequency (leaks/plant/year)'
print*,
print 201, (d(i), i=1,4)
print*,
print 202, (type(j), (fgfailp(i,j), i=1,4), j=1,2)
print*,
print*,'Mass flow (kg.s⁻¹)
print*,
print 204, (type(j), (fgmass(i,j), i=1,4), j=1,2)
print*,
print*,'Ignition probability'
print*,
print 202, (type(j), (fgignp(i,j), i=1,4), j=1,2)
print*,
print*,'Explosion probability'
print*,
print 202, (type(j), (fgexp(i,j), i=1,4), j=1,2)
print*,
print*,'Fire frequency (fires/plant/year - x 10000)'
print*,
print 204, (type(j), (fgff(i,j), i=1,4), j=1,2)
print*,
print*,'Explosion frequency (explosions/plant/year - x 10000)'
do 7900, i=1,26
print*,-"
print*,'Small-bore connections'
print*,-"
print*,'Leak frequency (leaks/plant/year)'
F22
print *, "Mass flow (kg.s-1)"
print *, "Ignition probability"
print *, "Explosion probability"
print *, "Fire frequency (fires/plant/year x 10000)"
print *, "Explosion frequency (explosions/plant/year x 10000)"
print *, "Two-phase releases"  
print *, "Leak frequency (leaks/plant/year)"
print *, "Mass flow (kg.s-1)"
print *, "Ignition probability"
print *, "Explosion probability"
print *, "Fire frequency (fires/plant/year x 10000)"
print *, "Explosion frequency (explosions/plant/year x 10000)"

print *, "Leak frequency (leaks/plant/year)"
print *, "Mass flow (kg.s-1)"
print *, "Ignition probability"
print *, "Explosion probability"
print *, "Fire frequency (fires/plant/year x 10000)"
print *, "Explosion frequency (explosions/plant/year x 10000)"

continue
print *, "Valves"
print *, "Leak frequency (leaks/plant/year)"
print *, ""  
print 202, (type(j),(v2failp(i,j),i=1,4),j=1,3)  
print *, "Mass flow (kg.s⁻¹)"  
print *, ""  
print 203, (type(j),(v2mass(i,j),i=1,4),j=1,3)  
print *, ""  
print*, "Ignition probability"  
print *, ""  
print 202, (type(j),(vign2p(i,j),i=1,4),j=1,3)  
print *, ""  
print*, "Explosion probability"  
print *, ""  
print 202, (type(j),(v2xp(i,j),i=1,4),j=1,3)  
print*, ""  
print*, "Explosion frequency (explosions/plant/year - x 10000)"  
print *, ""  
print 204, (type(j),(1e04*v2expf(i,j),i=1,4),j=1,3)  
print*, ""  
do 7700, i=1,26  
print *, ""  
5100 continue  
print*, ""  
print*, "Flanges"  
print *, ""  
print*, ""  
print*, ""  
print*, "Leak frequency (leaks/plant/year)"  
print *, ""  
print 201, (d(i),i=1,4)  
print *, ""  
print 202, (ftype(j),(f2failp(i,j),i=1,4),j=1,2)  
print *, ""  
print*, "Mass flow (kg.s⁻¹)"  
print *, ""  
print 204, (ftype(j),(f2mass(i,j),i=1,4),j=1,2)  
print*, ""  
print*, "Ignition probability"  
print *, ""  
print 202, (ftype(j),(fign2p(i,j),i=1,4),j=1,2)  
print *, ""  
print*, "Explosion probability"  
print *, ""  
print 202, (ftype(j),(fxp(i,j),i=1,4),j=1,2)  
print*, ""  
print*, "Fire frequency (fires/plant/year - x 10000)"  
print *, ""  
print 204, (ftype(j),(1e04*f2ff(i,j),i=1,4),j=1,2)  
print*, ""  
print*, ""  
print*, "Explosion frequency (explosions/plant/year - x 10000)"  
print *, ""  
print 204, (ftype(j),(1e04*f2expf(i,j),i=1,4),j=1,2)  
do 7700, i=1,34  
print *, ""  
7700 continue  
print*, "Small-bore connections"  
print *, ""  
print*, ""  
print*, ""  
print*, "Leak frequency (leaks/plant/year)"  
print *, ""  
print 237, (type(i),s2failp(i),i=1,2)  
print *, ""  
print*, "Mass flow (kg.s⁻¹)"  
print *, ""  
print 236, (type(i),s2mass(i),i=1,2)
1551 print*, " "
1552 print*, "Ignition probability"
1553 print*, " "
1554 print 237, (type(i),sign2p(i),i=1,2)
1555 print*, " "
1556 print*, "Explosion probability"
1557 print*, " "
1558 print 237, (type(i),sign2p(i),i=1,2)
1559 print*, " "
1560 print*, "Fire frequency (fires/plant/year \times 10000)"
1561 print*, " "
1562 print 238, (type(i),(1e04*sf(i)),i=1,2)
1563 print*, " "
1564 print*, "Explosion frequency (explosions/plant/year \times 10000)"
1565 print*, " "
1566 print 238, (type(i),(1e04*se(i)),i=1,2)
1567 do 6300,i=1,32
1568 print*, " "
1569 6300 continue
1570 print*, "Pumps"
1571 print*, "-----
1572 print*, " "
1573 print 201, (d(i),i=1,4)
1574 print*, " "
1575 print*, "Leak frequency (leaks/plant/year)"
1576 print*, " "
1577 print 202, (type(j), (m2failp(i,j),i=1,4),j=1,3)
1578 print*, " "
1579 print*, "Mass flow (kg.s-1)"
1580 print*, " "
1581 print 203, (type(j), (m2mass(i,j),i=1,4),j=1,3)
1582 print*, " "
1583 print*, "Ignition probability"
1584 print*, " "
1585 print 202, (type(j), (mign2p(i,j),i=1,4),j=1,3)
1586 print*, " "
1587 print*, "Explosion probability"
1588 print*, " "
1589 print 202, (type(j), (mex2p(i,j),i=1,4),j=1,3)
1590 print*, " "
1591 print*, "Fire frequency (fires/plant/year \times 10000)"
1592 print*, " "
1593 print 204, (type(j), (1e04*m2ff(i,j),i=1,4),j=1,3)
1594 print*, " "
1595 print*, "Explosion frequency (explosions/plant/year \times 10000)"
1596 print*, " "
1597 print 204, (type(j), (1e04*seff(i,j),i=1,4),j=1,3)
1598 cccccc******************SUMMARY TABLES******************
1599 do 3500, i=1,3500
1600 3500 continue
1601 cccccc****************SUMMARISE RESULTS*****************
1602 cccccc******************LEAK FREQUENCIES*****************
1603 ccccccEXAMPLE: PFFAILMIN=TOTAL PIPE FAILURE FREQUENCY RESULTING IN FLOWS OF LESS THAN 1 KG.S-1
1604 ccccccFLOWS OF LESS THAN 1 KG.S-1
1605 do 20 ffailmin=1,1,20
1606 20 ffailmin=1,1,20
1607 vfailmin=ffailmin+vgfailmin+v2failmin
1608 ffailmin=ffailmin+fgfailmin+f2failmin
1609 mfailmin=mfailmin+mgfailmin+m2failmin
1610 sfailmin=sfailmin+sgfailmin+s2failmin
1611 pfailmin=pfailmin+pgfailmin+p2failmin
1612 vfailmed=ffailmed+fgfailmed+f2failmed
1613 mfailmed=mfailmed+mgfailmed+m2failmed
1614 sfailmed=sfailmed+sgfailmed+s2failmed
1615 pfailmax=pfailmax+pgfailmax+p2failmax
1616
1617 \( v_{\text{failmax}} = v_{\text{fail1max}} + v_{\text{fail2max}} + v_{\text{fail3max}} \)
1618 \( f_{\text{failmax}} = f_{\text{fail1max}} + f_{\text{fail2max}} + f_{\text{fail3max}} \)
1619 \( m_{\text{failmax}} = m_{\text{fail1max}} + m_{\text{fail2max}} + m_{\text{fail3max}} \)
1620 \( s_{\text{failmax}} = s_{\text{fail1max}} + s_{\text{fail2max}} + s_{\text{fail3max}} \)
1621 \[ \begin{align*}
1622 & \text{EQUIPMENT LEAK FREQUENCY TOTALS} \\
1623 & \text{failmin = fail1min + fail2min + fail3min} \\
1624 & \text{failmed = fail1med + fail2med + fail3med} \\
1625 & \text{failmax = fail1max + fail2max + fail3max} \\
1626 & \text{EQUIPMENT LEAK FREQUENCY TOTALS} \\
1627 & \text{failmin = fail1min + fail2min + fail3min} \\
1628 & \text{failmed = fail1med + fail2med + fail3med} \\
1629 & \text{failmax = fail1max + fail2max + fail3max} \\
1630 & \text{EXAMPLE VFFMAX = FIRE FREQUENCY RESULTING FROM VALVE LEAKS} \\
1631 & \text{WITH A MASS FLOW GREATER THAN 50 KG.S-1} \\
1632 & \text{EXFMIN = EXPLOSION FREQUENCY RESULTING FROM PIPE LEAKS WITH} \\
1633 & \text{A MASS FLOW BELOW 1 KG.S-1} \\
1634 & \text{EXFMAD = EXPLOSION FREQUENCY RESULTING FROM PIPE LEAKS WITH} \\
1635 & \text{A MASS FLOW BELOW 1 KG.S-1} \\
1636 & \text{PERCENTAGE POINTS} \\
1637 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1638 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1639 & \text{failmt = fail1min + fail2min + fail3min} \\
1640 & \text{failmed = fail1med + fail2med + fail3med} \\
1641 & \text{failmax = fail1max + fail2max + fail3max} \\
1642 & \text{EXFMAD = EXPLOSION FREQUENCY RESULTING FROM PIPE LEAKS} \\
1643 & \text{WITH A MASS FLOW BELOW 1 KG.S-1} \\
1644 & \text{EXFMAD = EXPLOSION FREQUENCY RESULTING FROM PIPE LEAKS} \\
1645 & \text{WITH A MASS FLOW BELOW 1 KG.S-1} \\
1646 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1647 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1648 & \text{failmt = fail1min + fail2min + fail3min} \\
1649 & \text{failmed = fail1med + fail2med + fail3med} \\
1650 & \text{failmax = fail1max + fail2max + fail3max} \\
1651 & \text{PERCENTAGE POINTS} \\
1652 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1653 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1654 & \text{failmt = fail1min + fail2min + fail3min} \\
1655 & \text{failmed = fail1med + fail2med + fail3med} \\
1656 & \text{failmax = fail1max + fail2max + fail3max} \\
1657 & \text{EXFMAD = EXPLOSION FREQUENCY RESULTING FROM PIPE LEAKS} \\
1658 & \text{WITH A MASS FLOW BELOW 1 KG.S-1} \\
1659 & \text{EXFMAD = EXPLOSION FREQUENCY RESULTING FROM PIPE LEAKS} \\
1660 & \text{WITH A MASS FLOW BELOW 1 KG.S-1} \\
1661 & \text{PERCENTAGE POINTS} \\
1662 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1663 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1664 & \text{failmt = fail1min + fail2min + fail3min} \\
1665 & \text{failmed = fail1med + fail2med + fail3med} \\
1666 & \text{failmax = fail1max + fail2max + fail3max} \\
1667 & \text{PERCENTAGE POINTS} \\
1668 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1669 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1670 & \text{failmt = fail1min + fail2min + fail3min} \\
1671 & \text{failmed = fail1med + fail2med + fail3med} \\
1672 & \text{failmax = fail1max + fail2max + fail3max} \\
1673 & \text{PERCENTAGE POINTS} \\
1674 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1675 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1676 & \text{failmt = fail1min + fail2min + fail3min} \\
1677 & \text{failmed = fail1med + fail2med + fail3med} \\
1678 & \text{failmax = fail1max + fail2max + fail3max} \\
1679 & \text{PERCENTAGE POINTS} \\
1680 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1681 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1682 & \text{failmt = fail1min + fail2min + fail3min} \\
1683 & \text{failmed = fail1med + fail2med + fail3med} \\
1684 & \text{failmax = fail1max + fail2max + fail3max} \\
1685 & \text{PERCENTAGE POINTS} \\
1686 & \text{TOTAL FIRE FREQUENCIES BY EQUIPMENT SOURCE TYPE} \\
1687 & \text{CONVERTED TO PERCENTAGE OF OVERALL FREQUENCIES} \\
1688 & \text{failmt = fail1min + fail2min + fail3min} \\
1689 & \text{failmed = fail1med + fail2med + fail3med} \\
1690 & \text{failmax = fail1max + fail2max + fail3max} \\
1691 & \text{PERCENTAGE POINTS}
SUMMARIZE RESULTS = PHASE BREAKDOWN

EXAMPLE. **FNOTE** = FIRE FREQUENCY RESULTING FROM 2-PHASE SOURCE LEAK FREQUENCY

\[ \text{lfail} = \text{lfailmin} + \text{lfailmed} + \text{lfailmax} \]
\[ \text{gfail} = \text{gfailmin} + \text{gfailmed} + \text{gfailmax} \]
\[ \text{tfail} = \text{tfailmin} + \text{tfailmed} + \text{tfailmax} \]
\[ \text{lff} = \text{lffmin} + \text{lffmed} + \text{lffmax} \]
\[ \text{gff} = \text{gffmin} + \text{gffmed} + \text{gffmax} \]
\[ \text{tff} = \text{tffmin} + \text{tffmed} + \text{tffmax} \]
\[ \text{gexf} = \text{gexfmin} + \text{gexfmed} + \text{gexfmax} \]
\[ \text{texf} = \text{texfmin} + \text{texfmed} + \text{texfmax} \]

PERCENTAGE POINTS FOR EACH TOTAL PHASE FREQUENCY AS PART OF OVERALL PHASE FREQUENCY

\[ \text{ifaill0min} = 100 \times \frac{\text{lfailmin}}{\text{lfail}} \]
\[ \text{ifaill0med} = 100 \times \frac{\text{lfailmed}}{\text{lfail}} \]
\[ \text{ifaill0max} = 100 \times \frac{\text{lfailmax}}{\text{lfail}} \]
\[ \text{gfaill0min} = 100 \times \frac{\text{gfailmin}}{\text{gfail}} \]
\[ \text{gfaill0med} = 100 \times \frac{\text{gfailmed}}{\text{gfail}} \]
\[ \text{gfaill0max} = 100 \times \frac{\text{gfailmax}}{\text{gfail}} \]
\[ \text{tfaill0min} = 100 \times \frac{\text{tfailmin}}{\text{tfail}} \]
\[ \text{tfaill0med} = 100 \times \frac{\text{tfailmed}}{\text{tfail}} \]
\[ \text{tfaill0max} = 100 \times \frac{\text{tfailmax}}{\text{tfail}} \]
\[ \text{gexf10min} = 100 \times \frac{\text{gexfmin}}{\text{gexf}} \]
\[ \text{gexf10med} = 100 \times \frac{\text{gexfmed}}{\text{gexf}} \]
\[ \text{gexf10max} = 100 \times \frac{\text{gexfmax}}{\text{gexf}} \]
\[ \text{texf10min} = 100 \times \frac{\text{texfmin}}{\text{texf}} \]
\[ \text{texf10med} = 100 \times \frac{\text{texfmed}}{\text{texf}} \]

**SUMMARY RESULTS = PHASE BREAKDOWN**

FREQUENCY TOTALS FOR EACH PHASE EXAMPLE. **FNOTE** = TOTAL 2-PHASE SOURCE LEAK FREQUENCY

\[ \text{lfail} = \text{lfailmin} + \text{lfailmed} + \text{lfailmax} \]
\[ \text{gfail} = \text{gfailmin} + \text{gfailmed} + \text{gfailmax} \]
\[ \text{tfail} = \text{tfailmin} + \text{tfailmed} + \text{tfailmax} \]
\[ \text{lff} = \text{lffmin} + \text{lffmed} + \text{lffmax} \]
\[ \text{gff} = \text{gffmin} + \text{gffmed} + \text{gffmax} \]
\[ \text{tff} = \text{tffmin} + \text{tffmed} + \text{tffmax} \]
\[ \text{gexf} = \text{gexfmin} + \text{gexfmed} + \text{gexfmax} \]
\[ \text{texf} = \text{texfmin} + \text{texfmed} + \text{texfmax} \]

**SUMMARY RESULTS = PHASE BREAKDOWN**

OVERALL PHASE FREQUENCY EXAMPLE. **FNOTE** = EXPLOSION FREQUENCY

\[ \text{lfail} = \text{lfailmin} + \text{lfailmed} + \text{lfailmax} \]
\[ \text{gfail} = \text{gfailmin} + \text{gfailmed} + \text{gfailmax} \]
\[ \text{tfail} = \text{tfailmin} + \text{tfailmed} + \text{tfailmax} \]
\[ \text{lff} = \text{lffmin} + \text{lffmed} + \text{lffmax} \]
\[ \text{gff} = \text{gffmin} + \text{gffmed} + \text{gffmax} \]
\[ \text{tff} = \text{tffmin} + \text{tffmed} + \text{tffmax} \]
\[ \text{gexf} = \text{gexfmin} + \text{gexfmed} + \text{gexfmax} \]
\[ \text{texf} = \text{texfmin} + \text{texfmed} + \text{texfmax} \]

GAS LEAKS BELOW 1 KG.S-1 AS PERCENTAGE OF TOTAL GAS LEAK FREQUENCY

\[ \text{lfail} = \text{lfailmin} + \text{lfailmed} + \text{lfailmax} \]
\[ \text{gfail} = \text{gfailmin} + \text{gfailmed} + \text{gfailmax} \]
\[ \text{tfail} = \text{tfailmin} + \text{tfailmed} + \text{tfailmax} \]
\[ \text{lff} = \text{lffmin} + \text{lffmed} + \text{lffmax} \]
\[ \text{gff} = \text{gffmin} + \text{gffmed} + \text{gffmax} \]
\[ \text{tff} = \text{tffmin} + \text{tffmed} + \text{tffmax} \]
\[ \text{gexf} = \text{gexfmin} + \text{gexfmed} + \text{gexfmax} \]
\[ \text{texf} = \text{texfmin} + \text{texfmed} + \text{texfmax} \]
1747 \text{x10max}=100*\text{xfmax}/\text{xf}
1748 \text{ccccc}***************\text{TOTALS BY PHASE}***************
\text{c}****
1749 \text{failmin}=\text{lfailmin}+\text{gfailmin}+\text{tfailmin}
1750 \text{failmed}=\text{lfailmed}+\text{gfailmed}+\text{tfailmed}
1751 \text{failmax}=\text{lfailmax}+\text{gfailmax}+\text{tfailmax}
1752 \text{ffmin}=\text{lfmin}+\text{gffmin}+\text{tffmin}
1753 \text{ffmed}=\text{lfmed}+\text{gffmed}+\text{tffmed}
1754 \text{ffmax}=\text{lfmax}+\text{gffmax}+\text{tffmax}
1755 \text{exfmin}=\text{gexfmin}+\text{texfmin}
1756 \text{exfmed}=\text{gexfmed}+\text{texfmed}
1757 \text{exfmax}=\text{gexfmax}+\text{texfmax}
1758 \text{fail}=\text{failmin}+\text{failmed}+\text{failmax}
1759 \text{ff}=\text{ffmin}+\text{ffmed}+\text{ffmax}
1760 \text{exf}=\text{exfmin}+\text{exfmed}+\text{exfmax}
1761 \text{faill0min}=100*\text{failmin}/\text{fail}
1762 \text{faill0med}=100*\text{failmed}/\text{fail}
1763 \text{faill0max}=100*\text{failmax}/\text{fail}
1764 \text{ffl0min}=100*\text{ffmin}/\text{ff}
1765 \text{ffl0med}=100*\text{ffmed}/\text{ff}
1766 \text{ffl0max}=100*\text{ffmax}/\text{ff}
1767 \text{exfl0min}=100*\text{exfmin}/\text{exf}
1768 \text{exfl0med}=100*\text{exfmed}/\text{exf}
1769 \text{exfl0max}=100*\text{exfmax}/\text{exf}
1770 \text{ccccc}***************\text{CROSS-CHECK = MAX+MIN+MED FOR EACH}**********
1771 \text{ccccc}EXAMPLE \text{PFAILOV=OVERALL PIPE LEAK FREQUENCY CROSS-CHECK}
1772 \text{ccccc}BY SUMMING ALL PIPE FAILURE ARRAY ELEMENTS
1773 \text{do 4100 } i=1,4
1774 \text{do 4200 } j=1,3
1775 \text{pfailov}=\text{pfailov}+\text{plfailp}(i,j)+\text{pgfailp}(i,j)+\text{p2failp}(i,j)
1776 \text{vfailov}=\text{vfailov}+\text{v1failp}(i,j)+\text{vgfailp}(i,j)+\text{v2failp}(i,j)
1777 \text{plfailov}=\text{plfailov}+\text{plfailp}(i,j)
1778 \text{v1failov}=\text{v1failov}+\text{v1failp}(i,j)
1779 \text{pgfailov}=\text{pgfailov}+\text{pgfailp}(i,j)
1780 \text{vgfailov}=\text{vgfailov}+\text{vgfailp}(i,j)
1781 \text{p2failov}=\text{p2failov}+\text{p2failp}(i,j)
1782 \text{v2failov}=\text{v2failov}+\text{v2failp}(i,j)
1783 \text{ccccc}FIRES***************
1784 \text{pffov}=\text{pffov}+\text{plff}(i,j)+\text{pgff}(i,j)+\text{p2ff}(i,j)
1785 \text{vffov}=\text{vffov}+\text{v1ff}(i,j)+\text{vgff}(i,j)+\text{v2ff}(i,j)
1786 \text{plffov}=\text{plffov}+\text{plff}(i,j)
1787 \text{v1ffov}=\text{v1ffov}+\text{v1ff}(i,j)
1788 \text{pgffov}=\text{pgffov}+\text{pgff}(i,j)
1789 \text{vgffov}=\text{vgffov}+\text{vgff}(i,j)
1790 \text{p2ffov}=\text{p2ffov}+\text{p2ff}(i,j)
1791 \text{v2ffov}=\text{v2ffov}+\text{v2ff}(i,j)
1792 \text{ccccc}EXPLOSIONS***************
1793 \text{pexfov}=\text{pexfov}+\text{pgexf}(i,j)+\text{p2exf}(i,j)
1794 \text{vexfov}=\text{vexfov}+\text{vge:xf}(i,j)+\text{v2exf}(i,j)
1795 \text{pgexfov}=\text{pgexfov}+\text{pgexf}(i,j)
1796 \text{vge:xfov}=\text{vge:xfov}+\text{vge:xf}(i,j)
1797 \text{p2exfov}=\text{p2exfov}+\text{p2exf}(i,j)
1798 \text{v2exfov}=\text{v2exfov}+\text{v2exf}(i,j)
1799 \text{cccccAND MOVERS}
1800 \text{mfailov}=\text{mfailov}+\text{mlfailp}(i,j)+\text{mgfailp}(i,j)+\text{m2failp}(i,j)
1801 \text{mlfailov}=\text{mlfailov}+\text{mlfailp}(i,j)
1802 \text{mgfailov}=\text{mgfailov}+\text{mgfailp}(i,j)
1803 \text{m2failov}=\text{m2failov}+\text{m2failp}(i,j)
1804 \text{mffov}=\text{mffov}+\text{mlff}(i,j)+\text{mgff}(i,j)+\text{m2ff}(i,j)
1805 \text{mlffov}=\text{mlffov}+\text{mlff}(i,j)
1806 \text{mgffov}=\text{mgffov}+\text{mgff}(i,j)
1807 \text{m2ffov}=\text{m2ffov}+\text{m2ff}(i,j)
1808 \text{maxfov}=\text{maxfov}+\text{maxen}(i,j)+\text{m2exf}(i,j)
1809 \text{maxenfov}=\text{maxenfov}+\text{maxen}(i,j)
1810 \text{m2exfov}=\text{m2exfov}+\text{m2exf}(i,j)
1811 \text{4200 continue}
do 4300, i=1,4
        do 4350, j=1,2
            ffailov=ffailov+ffailp(i,j)+ffailp(i,j)+2ffailp(i,j)
            fffov=fffov+ff(i,j)+fgf(i,j)+f2ff(i,j)
            fflov=fflov+ff(i,j)
            fggfv=fggf+fgf(i,j)
            f2fglov=f2ffov+f2ff(i,j)
            fexfov=fexfov+fgef(i,j)+f2exf(i,j)
            f2exfov=f2exfov+f2exf(i,j)
        end
4350 continue
4300 continue
cccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccccc
1876 \[ mg_{fail}100_{ov} = 100 \times mg_{fail}ov/gfail \]
1877 \[ sg_{fail}100_{ov} = 100 \times sg_{fail}ov/gfail \]
1878 \[ p2_{fail}100_{ov} = 100 \times p2_{fail}ov/tfail \]
1879 \[ v2_{fail}100_{ov} = 100 \times v2_{fail}ov/tfail \]
1880 \[ f2_{fail}100_{ov} = 100 \times f2_{fail}ov/tfail \]
1881 \[ m2_{fail}100_{ov} = 100 \times m2_{fail}ov/tfail \]
1882 \[ s2_{fail}100_{ov} = 100 \times s2_{fail}ov/tfail \]
1883 ccceccccFIRE FREQUENCIES
1884 \[ pl_{ff}100_{ov} = 100 \times pl_{ff}ov/lff \]
1885 \[ vl_{ff}100_{ov} = 100 \times vl_{ff}ov/lff \]
1886 \[ fl_{ff}100_{ov} = 100 \times fl_{ff}ov/lff \]
1887 \[ ml_{ff}100_{ov} = 100 \times ml_{ff}ov/lff \]
1888 \[ sl_{ff}100_{ov} = 100 \times sl_{ff}ov/lff \]
1889 \[ pg_{ff}100_{ov} = 100 \times pg_{ff}ov/gff \]
1890 \[ vg_{ff}100_{ov} = 100 \times vg_{ff}ov/gff \]
1891 \[ fg_{ff}100_{ov} = 100 \times fg_{ff}ov/gff \]
1892 \[ mg_{ff}100_{ov} = 100 \times mg_{ff}ov/gff \]
1893 \[ sg_{ff}100_{ov} = 100 \times sg_{ff}ov/gff \]
1894 \[ p2_{ff}100_{ov} = 100 \times p2_{ff}ov/tff \]
1895 \[ v2_{ff}100_{ov} = 100 \times v2_{ff}ov/tff \]
1896 \[ f2_{ff}100_{ov} = 100 \times f2_{ff}ov/tff \]
1897 \[ m2_{ff}100_{ov} = 100 \times m2_{ff}ov/tff \]
1898 \[ s2_{ff}100_{ov} = 100 \times s2_{ff}ov/tff \]
1899 ccceccccEXPLOSION FREQUENCIES
1900 \[ pg_{ex}100_{ov} = 100 \times pg_{ex}ov/gexf \]
1901 \[ vg_{ex}100_{ov} = 100 \times vg_{ex}ov/gexf \]
1902 \[ fg_{ex}100_{ov} = 100 \times fg_{ex}ov/gexf \]
1903 \[ mg_{ex}100_{ov} = 100 \times mg_{ex}ov/gexf \]
1904 \[ sg_{ex}100_{ov} = 100 \times sg_{ex}ov/gexf \]
1905 \[ p2_{ex}100_{ov} = 100 \times p2_{ex}ov/texf \]
1906 \[ v2_{ex}100_{ov} = 100 \times v2_{ex}ov/texf \]
1907 \[ f2_{ex}100_{ov} = 100 \times f2_{ex}ov/texf \]
1908 \[ m2_{ex}100_{ov} = 100 \times m2_{ex}ov/texf \]
1909 \[ s2_{ex}100_{ov} = 100 \times s2_{ex}ov/texf \]
1910 cccecccc********TOTALS AND PERCENTAGES FOR EQUIPMENT BREAKDOWN********
1911 ccceccccWHERE 11 APPEARS IN PERCENTAGE POINTS THE FOLLOWING APPLIES:
1912 ccceccccEXAMPLE: PLFAIL110V=LIQUID PIPE FAILURE FREQUENCY
1913 ccceccccAS PERCENTAGE OF OVERALL PIPE FAILURE FREQUENCY
1914 ccceccccPLFAIL100V=LIQUID PIPE FAILURE FREQUENCY AS PERCENTAGE OF
1915 ccceccccOVERALL LIQUID FAILURE FREQUENCY
1916 p1fail11lov=100*p1faillov/pfaillov
1917 p2fail11lov=100*p2faillov/pfaillov
1918 pfail10min=100*pfailmin/pfaillov
1919 pfail10med=100*pfailmed/pfaillov
1920 pfail10max=100*pfailmax/pfaillov
1921 pl_{ff}10lov=100*pl_{ff}ov/pfflov
1922 pl_{ff}10min=100*pl_{ff}min/pfflov
1923 pl_{ff}10med=100*pl_{ff}med/pfflov
1924 pl_{ff}10max=100*pl_{ff}max/pfflov
1925 pg_{ff}10lov=100*pg_{ff}ov/pfflov
1926 pg_{ff}10min=100*pg_{ff}min/pfflov
1927 pg_{ff}10med=100*pg_{ff}med/pfflov
1928 pg_{ff}10max=100*pg_{ff}max/pfflov
1929 p2_{ex}11lov=100*p2_{ex}ov/pexfov
1930 p2_{ex}10min=100*p2_{ex}min/pexfov
1931 p2_{ex}10med=100*p2_{ex}med/pexfov
1932 p2_{ex}10max=100*p2_{ex}max/pexfov
1933 ccceccccVALVES
1934 v1fail11lov=100*v1faillov/vfaillov
1935 v2fail11lov=100*v2faillov/vfaillov
1936 vfail10min=100*vfailmin/vfaillov
1937 vfail10med=100*vfailmed/vfaillov
1938 vfail10max=100*vfailmax/vfaillov
1939 v1_{ff}10lov=100*v1_{ff}ov/vfflov
1940 vg_{ff}10lov=100*vg_{ff}ov/vfflov
1941 vg_{ff}10min=100*vg_{ff}min/vfflov
1942 vg_{ff}10med=100*vg_{ff}med/vfflov
1943 vg_{ff}10max=100*vg_{ff}max/vfflov
1944 cccecccc--
"Summary tables - equipment breakdown"
|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|------|
prin*,*
print*,*
print*,*"Fire frequencies by flow (fires/plant/year x 10000)"
print*,*-----------------------------------------------
print*,*
print 299
print*,*
print *,1e04*v1ffmin,1e04*v1ffmed,1e04*v1ffmax
*,1e04*v1ffov,v1ff110v
print 226, 1e04*v1ffmin,1e04*v1ffmed,1e04*v1ffmax
*,1e04*v1ffov,v1ff110v
print 227, 1e04*vgffmin,1e04*vgffmed,1e04*vgffmax
*,1e04*vgffov,vgff110v
print 228, 1e04*v2ffmin,1e04*v2ffmed,1e04*v2ffmax
*,1e04*v2ffov,v2ff110v
print*,*
print 208, 1e04*vffmin,1e04*vffmed,1e04*vffmax,1e04*vffov
print*,*
print 226, le04*vlffmin,le04*vlffmed,le04*vlffmax
*,le04*vlffov,vlff110v
print 227, le04*vgffmin,le04*vgffmed,le04*vgffmax
*,le04*vgffov,vgff110v
print 228, le04*v2ffmin,le04*v2ffmed,le04*v2ffmax
*,le04*v2ffov,v2ff110v
print*,*
print*,1I
print*,1I 11
print*,1I 11
print 299
print*,1I 11
print 226, le04*vlffmin,le04*vlffmed,le04*vlffmax
*,le04*vlffov,vlff110v
print 227, le04*vgffmin,le04*vgffmed,le04*vgffmax
*,le04*vgffov,vgff110v
print 228, le04*v2ffmin,le04*v2ffmed,le04*v2ffmax
*,le04*v2ffov,v2ff110v
print*,*
print 208, le04*vffmin,le04*vffmed,le04*vffmax,le04*vffov
print*,*
print*,1I
print 226, le04*vlffmin,le04*vlffmed,le04*vlffmax
*,le04*vlffov,vlff110v
print 227, le04*vgffmin,le04*vgffmed,le04*vgffmax
*,le04*vgffov,vgff110v
print 228, le04*v2ffmin,le04*v2ffmed,le04*v2ffmax
*,le04*v2ffov,v2ff110v
print*,*
print*,1I
print*,1I
print*,1I
print*,1I 11
print*,1I 11
print 226, le04*vlffmin,le04*vlffmed,le04*vlffmax
*,le04*vlffov,vlff110v
print 227, le04*vgffmin,le04*vgffmed,le04*vgffmax
*,le04*vgffov,vgff110v
print 228, le04*v2ffmin,le04*v2ffmed,le04*v2ffmax
*,le04*v2ffov,v2ff110v
print*,*
print 226, le04*vlffmin,le04*vlffmed,le04*vlffmax
*,le04*vlffov,vlff110v
print 227, le04*vgffmin,le04*vgffmed,le04*vgffmax
*,le04*vgffov,vgff110v
print 228, le04*v2ffmin,le04*v2ffmed,le04*v2ffmax
*,le04*v2ffov,v2ff110v
print*,*
print 226, le04*vlffmin,le04*vlffmed,le04*vlffmax
*,le04*vlffov,vlff110v
print 227, le04*vgffmin,le04*vgffmed,le04*vgffmax
*,le04*vgffov,vgff110v
print 228, le04*v2ffmin,le04*v2ffmed,le04*v2ffmax
*,le04*v2ffov,v2ff110v
print*,*
print 208, le04*vffmin,le04*vffmed,le04*vffmax,le04*vffov
print*,*

F33
print 209, fff10min,fff10med,fff10max
print*, " "
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, " "
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
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print*, "Explosion frequencies (explosions/plant/year x 10000)"
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print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
print*, "Explosion frequencies (explosions/plant/year x 10000)"
Leak frequencies by flow (leaks/plant/year)

---

Fire frequencies by flow (fires/plant/year x 10000)

Explosion frequencies (explosions/plant/year x 10000)

---

Summary tables - equipment totals
print 213, failmin, failmed, failmax, fail
print*, " 
print 214, fail10min, fail10med, fail10max
print*, " 
print*, " 
print*, "Fire frequencies by flow (fires/plant/year × 10000)"
print*, "-------------------- ___________________________________
print*, " 
print*, " 
print*, 
print*,*Connections"
print 214, exfl0min, exfl0med, exfl0max
214 format (1h, "Per cent", 2x, f7.3, 4x, f7.3, 4x, f7.3)
do 3600, i=1,16
print*, ""
3600 continue

ccccc**************************SUMMARISE RESULTS BY PHASE**************************
c**
print*, "Summary tables - phase breakdown"
c**
print*, "Liquid phase"
c**
print*, "Leak frequency by flow (Leaks/plant/year)"
c**
print*, "Fire frequency by flow (Fires/plant/year x 10000)"
c**
do 3800, i=1,21
3800 continue

ccccc**************************SUMMARISE RESULTS BY PHASE**************************
c**
print*, "Gas phase"
c**
print*, "Leak frequencies by flow (Leaks/plant/year)"
c**
F37
print 299
print*, "  
print 210, pgf failmin, pgf failmed, pgf failmax, pgf failov, pgf fail100ov
print 211, vgfailmin, vgfailmed, vgfailmax, vgfailov, vgfail100ov
print 212, fgfailmin, fgfailmed, fgfailmax, fgfailov, fgfail100ov
print 234, sgfailmin, sgfailmed, sgfailmax, sgfailov, sgfail100ov
print*, "connections"
print*, "  
print 213, gfailmin, gfailmed, gfailmax, gfail
print*, "  
print 214, gfail10min, gfail10med, gfail10max
print*, "  
print*, "Fire frequencies by flow (Fires/plant/year x 10000)"
print*, "-----------------------------
print*, "  
print 205, le04*pgffmin, le04*pgffmed, le04*pgffmax
print*, le04*pgffav, pgffl0ov
print 206, le04*vgffmin, le04*vgffmed, le04*vgffmax
print*, le04*vgffov, vgffl0ov
print 207, le04*fgffmin, le04*fgffmed, le04*fgffmax
print*, le04*fgffov, fgffl0ov
print 235, le04*sgffmin, le04*sgffmed, le04*sgffmax
print*, le04*sgffov, sgffl0ov
print*, "connections"
print*, "  
print 208, le04*gffmin, le04*gffmed, le04*gffmax, le04*gff
print*, "  
print 209, gff10min, gff10med, gff10max
print*, "  
print*, "Explosion frequencies (Explosions/plant/year x 10000)"
print*, "-----------------------------
print*, "  
print 205, le04*pgexfmin, le04*pgexfmed, le04*pgexfmax
print*, le04*pgexfav, pgexf10ov
print 206, le04*vgexfmin, le04*vgexfmed, le04*vgexfmax
print*, le04*vgexfov, vgexf10ov
print 207, le04*fgexfmin, le04*fgexfmed, le04*fgexfmax
print*, le04*fgexfov, fgexf10ov
print 235, le04*sgexfmin, le04*sgexfmed, le04*sgexfmax
print*, le04*sgexfov, sgexf10ov
print*, "connections"
print*, "  
print 208, le04*gexfmin, le04*gexfmed, le04*gexfmax, le04*gexf
print*, "  
print 209, gexf10min, gexf10med, gexf10max
print*, "  
print*, "Explosion frequencies (Explosions/plant/year x 10000)"
print*, "-----------------------------
print*, "  
print 299
print*, "  
print 210, p2failmin, p2failmed, p2failmax, p2failov, p2fail100ov
print 211, v2failmin, v2failmed, v2failmax, v2failov, v2fail100ov
print 212, f2failmin, f2failmed, f2failmax, f2failov, f2fail100ov
print*, "connections"
print*, "  
print 213, gfailmin, gfailmed, gfailmax, gfail
print*, "  
print 214, gfail10min, gfail10med, gfail10max
print*, "  
print*, "Explosion frequencies (Explosions/plant/year x 10000)"
print*, "-----------------------------
print*, "  
print 205, le04*pgffmin, le04*pgffmed, le04*pgffmax
print*, le04*pgffav, pgffl0ov
print 206, le04*vgffmin, le04*vgffmed, le04*vgffmax
print*, le04*vgffov, vgffl0ov
print 207, le04*fgffmin, le04*fgffmed, le04*fgffmax
print*, le04*fgffov, fgffl0ov
print 235, le04*sgffmin, le04*sgffmed, le04*sgffmax
print*, le04*sgffov, sgffl0ov
print*, "connections"
print*, "  
print 208, le04*gffmin, le04*gffmed, le04*gffmax, le04*gff
print*, "  
print 209, gff10min, gff10med, gff10max
print*, "  
print*, "Explosion frequencies (Explosions/plant/year x 10000)"
print*, "-----------------------------
print*, "  
print 299
print*, "  
3700 continue
print*, "Two-phase leaks"
print 232, m2failmin, m2failmed, m2failmax, m2failcov, m2fail10ov
print 234, s2failmin, s2failmed, s2failmax, s2failcov, s2fail10ov
print*, "connections"
print*, "
print 213, tfailmin, tfailmed, tfailmax, tfail
print*, "
print 214, tfail10min, tfail10med, tfail10max
print*, "
print*, "Fire frequencies by flow (Fires/plant/year x 10000)"
print*, "-------------------------------------------------------------------"
print*, "
print 205, e04*p2ffmin, e04*p2ffmed, e04*p2ffmax
*, e04*p2ffcov, p2ff10ov
print 206, e04*v2ffmin, e04*v2ffmed, e04*v2ffmax
*, e04*v2ffcov, v2ff10ov
print 207, e04*f2ffmin, e04*f2ffmed, e04*f2ffmax
*, e04*f2ffcov, f2ff10ov
print 233, e04*m2ffmin, e04*m2ffmed, e04*m2ffmax
*, e04*m2ffcov, m2ff10ov
print 235, e04*s2ffmin, e04*s2ffmed, e04*s2ffmax
*, e04*s2ffcov, s2ff10ov
print*, "connections"
print*, "
print 208, e04*tffmin, e04*tffmed, e04*tffmax, e04*tff
print*, "
print 209, tff10min, tff10med, tff10max
print*, "
print*, "Explosion frequencies (Explosions/plant/year x 10000)"
print*, "-------------------------------------------------------------------"
print*, "
print 205, e04*p2exfmin, e04*p2exfmed, e04*p2exfmax
*, e04*p2exfcov, p2exf10ov
print 206, e04*v2exfmin, e04*v2exfmed, e04*v2exfmax
*, e04*v2exfcov, v2exf10ov
print 207, e04*f2exfmin, e04*f2exfmed, e04*f2exfmax
*, e04*f2exfcov, f2exf10ov
print 233, e04*m2exfmin, e04*m2exfmed, e04*m2exfmax
*, e04*m2exfcov, m2exf10ov
print 235, e04*s2exfmin, e04*s2exfmed, e04*s2exfmax
*, e04*s2exfcov, s2exf10ov
print*, "connections"
print*, "
print 213, e04*txxfmin, e04*txxfmed, e04*txxfmax, e04*txxf
print*, "
print 214, txxf10min, txxf10med, txxf10max
print*, "
do 4000, i=1,14
print*, 
4000 continue
print*, "**********************************************************************************" print*, "Summary tables - phase totals"
print*, "**********************************************************************************
print*, "**********************************************************************************
print*, "Leak frequencies (leaks/plant/year)"
print*, "**********************************************************************************
print*, "Explosion frequencies (Explosions/plant/year)"
print*, "**********************************************************************************
print*, "**********************************************************************************
print 223, lfailmin, lfailmed, lfailmax, lfail, lfail10
Main Program

print 224, gfailmin, gfailmed, gfailmax, gfail, gfail10
print 225, tfailmin, tfailmed, tfailmax, tfail, tfail10

format (1h, "Liquid", 4x, f9.6 2x, f9.6 2x, f9.6 2x, f9.6 2x, f5.1)

format (1h, "Gas", 7x, f9.6 2x, f9.6 2x, f9.6 2x, f9.6 2x, f5.1)

format (1h, "Two-phase", x, f9.6 2x, f9.6 2x, f9.6 2x, f9.6 2x, f5.1)

print *, "
print 213, failmin, failmed, failmax, fail

print *, "
print 214, fail10min, fail10med, fail10max

print *, "
print *, "Fire frequency (fires/plant/year x 10000)"

print *, "----------------------------------------

print *, "
print 299

print *, "
print 226, le04*lffmin, le04*lffmed, le04*lffmax
print *, 

*, le04*lff, lff10

print 227, le04*gffmin, le04*gffmed, le04*gffmax

*, le04*gff, gff10

print 228, le04*tffmin, le04*tffmed, le04*tffmax

*, le04*tff, tff10

format (1h, "Liquid", 4x, f7.3 4x, f7.3 4x, f7.3 4x, f5.1)

format (1h, "Gas", 7x, f7.3 4x, f7.3 4x, f7.3 4x, f5.1)

format (1h, "Two-phase", x, f7.3 4x, f7.3 4x, f7.3 4x, f5.1)

print *, "
print 208, le04*fmin, le04*fmed, le04*fmax, le04*f

print *, "
print 209, ffl0min, ffl0med, ffl0max

print *, 
print *, "Explosion frequency (explosions/plant/year x 10000)"

print *, "----------------------------------------

print *, "
print 208, le04*exfmin, le04*exfmed, le04*exfmax

print *, "
print 209, exf10min, exf10med, exf10max

print *, 
print *, "**************************************************

10 stop

end
Appendix G. Complete output from fire and explosion model program IGNITION
Ignition analysis program - Initial conditions

Plant operating conditions

Average plant pressure = 15.0 bar a
Average plant temperature = 300.0 K

Fluid properties

Molecular weight = 44.08 kg.kmol⁻¹
Liquid density at plant temperature = 600.0 kg.m⁻³
Expansion coefficient = 1.4

Leak orifice discharge coefficients

Plain hole (gasket, seal etc.) Cd = 0.6
Full bore pipe leak Cd = 0.8

Ignition/Explosion probabilities

Gas ignition probability at leak rate of 0.5 kg.s⁻¹ = 0.01
Gas ignition probability at leak rate of 100 kg.s⁻¹ = 0.3

Liquid ignition probability at leak rate of 0.5 kg.s⁻¹ = 0.01
Liquid ignition probability at leak rate of 100 kg.s⁻¹ = 0.08

Gas/Vapour explosion probability given ignition at leak rate of 0.5 kg.s⁻¹ = 0.025
Gas/Vapour explosion probability given ignition at leak rate of 100 kg.s⁻¹ = 0.25
Plant leak source and leak rate profile

Leak source profile

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipe lengths (m)</td>
<td>3750.0</td>
<td>4200.0</td>
<td>5400.0</td>
<td>1650.0</td>
</tr>
<tr>
<td>Valves</td>
<td>720.0</td>
<td>500.0</td>
<td>240.0</td>
<td>40.0</td>
</tr>
<tr>
<td>Flanges</td>
<td>900.0</td>
<td>1070.0</td>
<td>810.0</td>
<td>220.0</td>
</tr>
<tr>
<td>Gas fraction</td>
<td>0.00</td>
<td>0.15</td>
<td>0.15</td>
<td>0.25</td>
</tr>
<tr>
<td>Liquid fraction</td>
<td>0.50</td>
<td>0.40</td>
<td>0.40</td>
<td>0.35</td>
</tr>
<tr>
<td>Two-phase fraction</td>
<td>0.50</td>
<td>0.45</td>
<td>0.45</td>
<td>0.40</td>
</tr>
</tbody>
</table>

Pipe lengths (m)

<table>
<thead>
<tr>
<th>Gas</th>
<th>0.0</th>
<th>630.0</th>
<th>810.0</th>
<th>412.5</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>1875.0</td>
<td>1680.0</td>
<td>2160.0</td>
<td>577.5</td>
</tr>
<tr>
<td>Two-phase</td>
<td>1875.0</td>
<td>1890.0</td>
<td>2430.0</td>
<td>660.0</td>
</tr>
</tbody>
</table>

Valves

<table>
<thead>
<tr>
<th>Gas</th>
<th>0.0</th>
<th>75.0</th>
<th>36.0</th>
<th>10.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>360.0</td>
<td>200.0</td>
<td>96.0</td>
<td>14.0</td>
</tr>
<tr>
<td>Two-phase</td>
<td>360.0</td>
<td>225.0</td>
<td>108.0</td>
<td>16.0</td>
</tr>
</tbody>
</table>

Flanges

<table>
<thead>
<tr>
<th>Gas</th>
<th>0.0</th>
<th>160.5</th>
<th>121.5</th>
<th>55.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>450.0</td>
<td>428.0</td>
<td>324.0</td>
<td>77.0</td>
</tr>
<tr>
<td>Two-phase</td>
<td>450.0</td>
<td>481.5</td>
<td>364.5</td>
<td>88.0</td>
</tr>
</tbody>
</table>

Pumps

<table>
<thead>
<tr>
<th>Liquid</th>
<th>0.0</th>
<th>11.0</th>
<th>6.0</th>
<th>0.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Two-phase</td>
<td>0.0</td>
<td>4.0</td>
<td>4.0</td>
<td>0.0</td>
</tr>
</tbody>
</table>

Small-bore connections

<table>
<thead>
<tr>
<th>Gas</th>
<th>105.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>280.0</td>
</tr>
<tr>
<td>Two-phase</td>
<td>315.0</td>
</tr>
</tbody>
</table>
# Failure frequency profile

## Pipework leak frequency (leaks/metre/year x 10000)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.0050</td>
<td>0.0050</td>
<td>0.0150</td>
<td>0.0050</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.0500</td>
<td>0.0500</td>
<td>0.0150</td>
<td>0.0050</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.5000</td>
<td>0.5000</td>
<td>0.1500</td>
<td>0.0500</td>
</tr>
</tbody>
</table>

## Valve leak frequency (leaks/valve/year x 10000)

| Major leak | 0.0100 | 0.0100 | 0.0100 | 0.0050 |
| Minor leak | 1.0000 | 1.0000 | 1.0000 | 0.5000 |

## Flange leak frequency (leak/flange/year x 10000)

| Major leak | 0.3000 | 0.3000 | 0.3000 | 0.3000 |
| Minor leak | 3.0000 | 3.0000 | 3.0000 | 3.0000 |

## Pump leak frequency (leak/pump/year x 10000)

| Major leak | 0.3000 | 0.3000 | 0.3000 | 0.3000 |
| Minor leak | 3.0000 | 3.0000 | 3.0000 | 3.0000 |

## Small-bore connection leak frequency (leak/connection/year x 10000)

| Major leak | 1.0000 |
| Minor leak | 10.0000 |
Leak size distribution

---

Pipework leak area (m² x 1000)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.4909</td>
<td>1.9635</td>
<td>7.8540</td>
<td>70.6858</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.0491</td>
<td>0.1983</td>
<td>0.7854</td>
<td>7.0686</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.0049</td>
<td>0.0196</td>
<td>0.0785</td>
<td>0.7069</td>
</tr>
</tbody>
</table>

Valve leak area (m² x 1000)

<table>
<thead>
<tr>
<th>Valve leak area (m² x 1000)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
</tr>
<tr>
<td>Major leak</td>
</tr>
<tr>
<td>Minor leak</td>
</tr>
</tbody>
</table>

Flange leak area (m² x 1000)

<table>
<thead>
<tr>
<th>Flange leak area (m² x 1000)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
</tr>
<tr>
<td>Minor leak</td>
</tr>
</tbody>
</table>

Pump leak area (m² x 1000)

<table>
<thead>
<tr>
<th>Pump leak area (m² x 1000)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
</tr>
<tr>
<td>Major leak</td>
</tr>
<tr>
<td>Minor leak</td>
</tr>
</tbody>
</table>

Small-bore connection leak area (m² x 1000)

<table>
<thead>
<tr>
<th>Small-bore connection leak area (m² x 1000)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
</tr>
<tr>
<td>Major leak</td>
</tr>
</tbody>
</table>
Ignition analysis results

Liquid leaks

Pipework

Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00094</td>
<td>0.00084</td>
<td>0.00032</td>
<td>0.00003</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00938</td>
<td>0.00840</td>
<td>0.00324</td>
<td>0.00029</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.09375</td>
<td>0.08400</td>
<td>0.03240</td>
<td>0.00289</td>
</tr>
</tbody>
</table>

Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>24.14</th>
<th>96.57</th>
<th>386.30</th>
<th>3476.70</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>1.21</td>
<td>4.83</td>
<td>19.32</td>
<td>173.84</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.12</td>
<td>0.48</td>
<td>1.93</td>
<td>17.38</td>
</tr>
</tbody>
</table>

Ignition probability

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>0.04580</th>
<th>0.07891</th>
<th>0.08000</th>
<th>0.08000</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.01413</td>
<td>0.02435</td>
<td>0.04196</td>
<td>0.08000</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.01700</td>
<td>0.04026</td>
</tr>
</tbody>
</table>

Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>0.429</th>
<th>0.663</th>
<th>0.259</th>
<th>0.023</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>1.325</td>
<td>2.046</td>
<td>1.359</td>
<td>0.231</td>
</tr>
<tr>
<td>Minor leak</td>
<td>9.375</td>
<td>8.400</td>
<td>5.507</td>
<td>1.162</td>
</tr>
</tbody>
</table>
### Valves

#### Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00036</td>
<td>0.00020</td>
<td>0.00010</td>
<td>0.00001</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00360</td>
<td>0.00200</td>
<td>0.00096</td>
<td>0.00007</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.03600</td>
<td>0.02000</td>
<td>0.00960</td>
<td>0.00070</td>
</tr>
</tbody>
</table>

#### Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>12.07</td>
<td>48.29</td>
<td>193.15</td>
</tr>
<tr>
<td>Minor leak</td>
<td>1.21</td>
<td>4.83</td>
<td>19.32</td>
</tr>
</tbody>
</table>

#### Ignition probability

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.03489</td>
<td>0.06012</td>
<td>0.08000</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.01413</td>
<td>0.02433</td>
<td>0.04196</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.01700</td>
</tr>
</tbody>
</table>

#### Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.126</td>
<td>0.120</td>
<td>0.077</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.509</td>
<td>0.487</td>
<td>0.403</td>
</tr>
<tr>
<td>Minor leak</td>
<td>3.600</td>
<td>2.000</td>
<td>1.632</td>
</tr>
</tbody>
</table>

### Flanges

#### Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.01350</td>
<td>0.01284</td>
<td>0.00972</td>
<td>0.00231</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.13500</td>
<td>0.12840</td>
<td>0.09720</td>
<td>0.02310</td>
</tr>
</tbody>
</table>

#### Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th></th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>1.93</td>
<td>1.93</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.19</td>
<td>0.19</td>
</tr>
</tbody>
</table>

#### Ignition probability

<table>
<thead>
<tr>
<th></th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.01700</td>
<td>0.01700</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.01000</td>
<td>0.01000</td>
</tr>
</tbody>
</table>

#### Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th></th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>2.294</td>
<td>2.182</td>
</tr>
<tr>
<td>Pumps</td>
<td></td>
<td></td>
</tr>
<tr>
<td>-------</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pipe diameter (m)</td>
<td>0.025</td>
<td>0.050</td>
</tr>
</tbody>
</table>

Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.000000</td>
<td>0.00033</td>
<td>0.00018</td>
</tr>
<tr>
<td></td>
<td>0.000000</td>
<td>0.00330</td>
<td>0.00180</td>
</tr>
<tr>
<td></td>
<td>0.000000</td>
<td>0.00300</td>
<td>0.00180</td>
</tr>
</tbody>
</table>

Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
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<tbody>
<tr>
<td></td>
<td>12.07</td>
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<tr>
<td></td>
<td>1.21</td>
<td>4.83</td>
<td>19.32</td>
</tr>
<tr>
<td></td>
<td>0.12</td>
<td>0.48</td>
<td>1.93</td>
</tr>
</tbody>
</table>

Ignition probability

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.03489</td>
<td>0.06012</td>
<td>0.08000</td>
</tr>
<tr>
<td></td>
<td>0.01413</td>
<td>0.02435</td>
<td>0.04196</td>
</tr>
<tr>
<td></td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.01700</td>
</tr>
</tbody>
</table>

Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.000</td>
<td>0.198</td>
<td>0.144</td>
</tr>
<tr>
<td></td>
<td>0.000</td>
<td>0.804</td>
<td>0.755</td>
</tr>
<tr>
<td></td>
<td>0.000</td>
<td>3.300</td>
<td>3.059</td>
</tr>
</tbody>
</table>

Small-bore connections

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.02800</td>
<td>0.28000</td>
<td>0.28000</td>
</tr>
</tbody>
</table>

Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>2.5753</td>
<td>0.1932</td>
<td>0.1932</td>
</tr>
</tbody>
</table>

Ignition probability

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.01903</td>
<td>0.01000</td>
<td>0.01000</td>
</tr>
</tbody>
</table>

Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.001</td>
<td>0.003</td>
<td>0.003</td>
</tr>
</tbody>
</table>
### Gas leaks

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### Pipework

---

### Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00000</td>
<td>0.00031</td>
<td>0.00012</td>
<td>0.00002</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00000</td>
<td>0.00315</td>
<td>0.00121</td>
<td>0.00021</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.00000</td>
<td>0.03150</td>
<td>0.01215</td>
<td>0.00206</td>
</tr>
</tbody>
</table>

### Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>2.543</th>
<th>10.173</th>
<th>40.694</th>
<th>366.242</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.127</td>
<td>0.509</td>
<td>2.035</td>
<td>18.312</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.013</td>
<td>0.051</td>
<td>0.203</td>
<td>1.831</td>
</tr>
</tbody>
</table>

### Ignition probability

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>0.02841</th>
<th>0.06918</th>
<th>0.16845</th>
<th>0.30000</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.01006</td>
<td>0.01011</td>
<td>0.02462</td>
<td>0.10089</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.02301</td>
</tr>
</tbody>
</table>

### Explosion probability

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>0.00144</th>
<th>0.00641</th>
<th>0.02849</th>
<th>0.07500</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.00014</td>
<td>0.00025</td>
<td>0.00113</td>
<td>0.01206</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.00005</td>
<td>0.00009</td>
<td>0.00017</td>
<td>0.00101</td>
</tr>
</tbody>
</table>

### Fire frequency (fires/plant/year - x 10000)

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>0.000</th>
<th>0.218</th>
<th>0.205</th>
<th>0.062</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.000</td>
<td>0.318</td>
<td>0.299</td>
<td>0.208</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.000</td>
<td>3.150</td>
<td>1.215</td>
<td>0.475</td>
</tr>
</tbody>
</table>

### Explosion frequency (explosions/plant/year - x 10000)

<table>
<thead>
<tr>
<th>Rupture leak</th>
<th>0.000</th>
<th>0.020</th>
<th>0.035</th>
<th>0.015</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.000</td>
<td>0.008</td>
<td>0.014</td>
<td>0.025</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.000</td>
<td>0.029</td>
<td>0.021</td>
<td>0.021</td>
</tr>
</tbody>
</table>
Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00000</td>
<td>0.00008</td>
<td>0.00004</td>
<td>0.00000</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00000</td>
<td>0.00035</td>
<td>0.00036</td>
<td>0.00035</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.00000</td>
<td>0.00075</td>
<td>0.00070</td>
<td>0.00005</td>
</tr>
</tbody>
</table>

Mass flow (kg.s⁻¹)

| Rupture leak     | 1.272 | 5.087 | 20.347 | 183.122 |
| Major leak       | 0.127 | 0.509 | 2.035  | 18.312  |
| Minor leak       | 0.013 | 0.051 | 0.203  | 1.831   |

Ignition probability

| Rupture leak     | 0.01821 | 0.04433 | 0.10795 | 0.30000 |
| Major leak       | 0.01000 | 0.01011 | 0.02462 | 0.10089 |
| Minor leak       | 0.01000 | 0.01000 | 0.01000 | 0.02301 |

Explosion probability

| Rupture leak     | 0.00068 | 0.00304 | 0.01351 | 0.07500 |
| Major leak       | 0.00014 | 0.00025 | 0.00113 | 0.01206 |
| Minor leak       | 0.00005 | 0.00009 | 0.00017 | 0.00101 |

Fire frequency (fires/plant/year - x 10000)

| Rupture leak     | 0.000 | 0.033 | 0.039 | 0.015 |
| Major leak       | 0.000 | 0.076 | 0.089 | 0.050 |
| Minor leak       | 0.000 | 0.750 | 0.360 | 0.115 |

Explosion frequency (explosions/plant/year - x 10000)

| Rupture leak     | 0.000 | 0.002 | 0.005 | 0.004 |
| Major leak       | 0.000 | 0.002 | 0.004 | 0.006 |
| Minor leak       | 0.000 | 0.007 | 0.006 | 0.005 |
### Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.000</td>
<td>0.00481</td>
<td>0.00364</td>
<td>0.00165</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.000</td>
<td>0.04815</td>
<td>0.03645</td>
<td>0.01650</td>
</tr>
</tbody>
</table>

### Mass flow (kg.s⁻¹)

| Major leak | 0.203 | 0.203 | 0.271 | 0.488 |
| Minor leak | 0.020 | 0.020 | 0.027 | 0.049 |

### Ignition probability

| Major leak | 0.01000 | 0.01000 | 0.01000 | 0.01000 |
| Minor leak | 0.01000 | 0.01000 | 0.01000 | 0.01000 |

### Explosion probability

| Major leak | 0.00017 | 0.00017 | 0.00019 | 0.00025 |
| Minor leak | 0.00006 | 0.00006 | 0.00007 | 0.00009 |

### Fire frequency (fires/plant/year - x 10000)

| Major leak | 0.000 | 0.481 | 0.365 | 0.165 |
| Minor leak | 0.000 | 4.815 | 3.645 | 1.650 |

### Explosion frequency (explosions/plant/year - x 10000)

| Major leak | 0.000 | 0.008 | 0.007 | 0.004 |
| Minor leak | 0.000 | 0.030 | 0.026 | 0.015 |
Small-bore connections

Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Condition</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.01050</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.10500</td>
</tr>
</tbody>
</table>

Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th>Condition</th>
<th>Mass Flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.2713</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.0203</td>
</tr>
</tbody>
</table>

Ignition probability

<table>
<thead>
<tr>
<th>Condition</th>
<th>Probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.01000</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.01000</td>
</tr>
</tbody>
</table>

Explosion probability

<table>
<thead>
<tr>
<th>Condition</th>
<th>Probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00019</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00006</td>
</tr>
</tbody>
</table>

Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Condition</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>1.050</td>
</tr>
<tr>
<td>Major leak</td>
<td>10.500</td>
</tr>
</tbody>
</table>

Explosion frequency (explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Condition</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.020</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.065</td>
</tr>
</tbody>
</table>
Two-phase releases

Pipework

Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00094</td>
<td>0.00094</td>
<td>0.00036</td>
<td>0.00003</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00938</td>
<td>0.00945</td>
<td>0.00364</td>
<td>0.00033</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.09375</td>
<td>0.09450</td>
<td>0.03645</td>
<td>0.00330</td>
</tr>
</tbody>
</table>

Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>19.31</td>
<td>77.26</td>
<td>309.04</td>
<td>2781.36</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.97</td>
<td>3.86</td>
<td>15.45</td>
<td>139.07</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.10</td>
<td>0.39</td>
<td>1.55</td>
<td>13.91</td>
</tr>
</tbody>
</table>

Ignition probability

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.10440</td>
<td>0.25421</td>
<td>0.30000</td>
<td>0.30000</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.01526</td>
<td>0.03715</td>
<td>0.09047</td>
<td>0.30000</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.02063</td>
<td>0.08455</td>
</tr>
</tbody>
</table>

Explosion probability

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.01277</td>
<td>0.05681</td>
<td>0.07500</td>
<td>0.07500</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00051</td>
<td>0.00226</td>
<td>0.01005</td>
<td>0.07500</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.00012</td>
<td>0.00022</td>
<td>0.00084</td>
<td>0.00897</td>
</tr>
</tbody>
</table>

Fire frequency (fires/plant/year - x 10000)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.979</td>
<td>2.402</td>
<td>1.094</td>
<td>0.099</td>
</tr>
<tr>
<td>Major leak</td>
<td>1.431</td>
<td>3.511</td>
<td>3.298</td>
<td>0.990</td>
</tr>
<tr>
<td>Minor leak</td>
<td>9.375</td>
<td>9.450</td>
<td>7.521</td>
<td>2.790</td>
</tr>
</tbody>
</table>

Explosion frequency (explosions/plant/year - x 10000)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.120</td>
<td>0.537</td>
<td>0.273</td>
<td>0.025</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.048</td>
<td>0.213</td>
<td>0.366</td>
<td>0.247</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.115</td>
<td>0.211</td>
<td>0.307</td>
<td>0.296</td>
</tr>
</tbody>
</table>
### Valves

**Leak frequency (leaks/plant/year)**

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00036</td>
<td>0.00022</td>
<td>0.00011</td>
<td>0.00001</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00360</td>
<td>0.00225</td>
<td>0.00108</td>
<td>0.00008</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.03600</td>
<td>0.02250</td>
<td>0.01080</td>
<td>0.00080</td>
</tr>
</tbody>
</table>

**Mass flow (kg.s⁻¹)**

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>9.66</td>
<td>38.63</td>
<td>154.52</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.97</td>
<td>3.86</td>
<td>15.45</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.10</td>
<td>0.39</td>
<td>1.55</td>
</tr>
</tbody>
</table>

**Ignition probability**

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.06691</td>
<td>0.16291</td>
<td>0.30000</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.01526</td>
<td>0.03715</td>
<td>0.09047</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.01000</td>
<td>0.01000</td>
<td>0.02063</td>
</tr>
</tbody>
</table>

**Explosion probability**

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00606</td>
<td>0.02694</td>
<td>0.07500</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00051</td>
<td>0.00226</td>
<td>0.01005</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.00012</td>
<td>0.00022</td>
<td>0.00084</td>
</tr>
</tbody>
</table>

**Fire frequency (fires/plant/year - x 10000)**

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.241</td>
<td>0.367</td>
<td>0.324</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.549</td>
<td>0.856</td>
<td>0.977</td>
</tr>
<tr>
<td>Minor leak</td>
<td>3.600</td>
<td>2.250</td>
<td>2.228</td>
</tr>
</tbody>
</table>

**Explosion frequency (explosions/plant/year - x 10000)**

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.022</td>
<td>0.061</td>
<td>0.081</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.018</td>
<td>0.051</td>
<td>0.108</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.044</td>
<td>0.050</td>
<td>0.091</td>
</tr>
</tbody>
</table>
### Flanges

#### Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
<tbody>
<tr>
<td>Major leak</td>
<td>0.01350</td>
<td>0.01444</td>
<td>0.01093</td>
<td>0.00264</td>
</tr>
<tr>
<td>Minor leak</td>
<td>0.13500</td>
<td>0.14445</td>
<td>0.10935</td>
<td>0.02640</td>
</tr>
</tbody>
</table>

#### Mass flow (kg.s⁻¹)

| Major leak | 1.545 | 1.545 | 2.060 | 3.708 |
| Minor leak | 0.155 | 0.155 | 0.206 | 0.371 |

#### Ignition probability

| Major leak         | 0.02063 | 0.02063 | 0.02482 | 0.03619 |
| Minor leak         | 0.01000 | 0.01000 | 0.01000 | 0.01000 |

#### Explosion probability

| Major leak         | 0.00084 | 0.00084 | 0.00115 | 0.00216 |
| Minor leak         | 0.00015 | 0.00015 | 0.00017 | 0.00022 |

#### Fire frequency (fires/plant/year - x 10000)

| Major leak | 2.785 | 2.980 | 2.714 | 0.956 |
| Minor leak | 13.300 | 14.445 | 10.935 | 2.640 |

#### Explosion frequency (explosions/plant/year - x 10000)

| Major leak | 0.114 | 0.122 | 0.126 | 0.057 |
| Minor leak | 0.203 | 0.217 | 0.186 | 0.058 |
Small-bore connections

---

**Leak frequency (leaks/plant/year)**

<table>
<thead>
<tr>
<th>Type</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.03150</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.31500</td>
</tr>
</tbody>
</table>

**Mass flow (kg.s⁻¹)**

<table>
<thead>
<tr>
<th>Type</th>
<th>Mass flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>2.0603</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.1545</td>
</tr>
</tbody>
</table>

**Ignition probability**

<table>
<thead>
<tr>
<th>Type</th>
<th>Probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.02482</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.01000</td>
</tr>
</tbody>
</table>

**Explosion probability**

<table>
<thead>
<tr>
<th>Type</th>
<th>Probability</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.00115</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.00015</td>
</tr>
</tbody>
</table>

**Fire frequency (fires/plant/year x 10000)**

<table>
<thead>
<tr>
<th>Type</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>7.818</td>
</tr>
<tr>
<td>Major leak</td>
<td>31.500</td>
</tr>
</tbody>
</table>

**Explosion frequency (explosions/plant/year x 10000)**

<table>
<thead>
<tr>
<th>Type</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rupture leak</td>
<td>0.020</td>
</tr>
<tr>
<td>Major leak</td>
<td>0.065</td>
</tr>
</tbody>
</table>
### Pumps

<table>
<thead>
<tr>
<th>Pipe diameter (m)</th>
<th>0.025</th>
<th>0.050</th>
<th>0.100</th>
<th>0.300</th>
</tr>
</thead>
</table>

#### Leak frequency (leaks/plant/year)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.00000</td>
<td>0.00120</td>
<td>0.01200</td>
<td>0.00012</td>
<td>0.00120</td>
<td>0.01200</td>
</tr>
</tbody>
</table>

#### Mass flow (kg.s⁻¹)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>9.66</td>
<td>38.63</td>
<td>154.52</td>
<td>1390.68</td>
<td>139.07</td>
<td>13.91</td>
</tr>
</tbody>
</table>

#### Ignition probability

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.06691</td>
<td>0.16291</td>
<td>0.30000</td>
<td>0.30000</td>
<td>0.30000</td>
<td>0.30000</td>
</tr>
</tbody>
</table>

#### Explosion probability

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.00606</td>
<td>0.02694</td>
<td>0.07500</td>
<td>0.07500</td>
<td>0.07500</td>
<td>0.07500</td>
</tr>
</tbody>
</table>

#### Fire frequency (fires/plant/year - x 10000)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.000</td>
<td>0.195</td>
<td>0.360</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
</tr>
</tbody>
</table>

#### Explosion frequency (explosions/plant/year - x 10000)

<table>
<thead>
<tr>
<th></th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
<th>Rupture leak</th>
<th>Major leak</th>
<th>Minor leak</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.000</td>
<td>0.032</td>
<td>0.090</td>
<td>0.000</td>
<td>0.000</td>
<td>0.000</td>
</tr>
</tbody>
</table>
### Summary tables - equipment breakdown

#### Pipework

**Leak frequencies by flow (Leaks/plant/year)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>0.177750</td>
<td>0.057240</td>
<td>0.001482</td>
<td>0.236472</td>
<td>43.8</td>
</tr>
<tr>
<td>Gas</td>
<td>0.046800</td>
<td>0.003920</td>
<td>0.000021</td>
<td>0.050741</td>
<td>9.4</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.197625</td>
<td>0.053783</td>
<td>0.001672</td>
<td>0.253080</td>
<td>46.8</td>
</tr>
<tr>
<td>Total</td>
<td>0.422175</td>
<td>0.114943</td>
<td>0.003175</td>
<td>0.540293</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>78.138</td>
<td>21.274</td>
<td>0.588</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Fire frequencies by flow (fires/plant/year x 10000)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>17.775</td>
<td>11.829</td>
<td>1.176</td>
<td>30.780</td>
<td>38.5</td>
</tr>
<tr>
<td>Gas</td>
<td>4.683</td>
<td>1.404</td>
<td>0.062</td>
<td>6.150</td>
<td>7.7</td>
</tr>
<tr>
<td>Two-phase</td>
<td>20.256</td>
<td>18.098</td>
<td>4.585</td>
<td>42.939</td>
<td>53.8</td>
</tr>
<tr>
<td>Total</td>
<td>42.714</td>
<td>31.331</td>
<td>5.823</td>
<td>79.868</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>53.481</td>
<td>39.229</td>
<td>7.291</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Explosion frequencies (explosions/plant/year x 10000)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>0.058</td>
<td>0.114</td>
<td>0.015</td>
<td>0.187</td>
<td>6.4</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.374</td>
<td>1.302</td>
<td>1.082</td>
<td>2.758</td>
<td>93.6</td>
</tr>
<tr>
<td>Total</td>
<td>0.431</td>
<td>1.417</td>
<td>1.098</td>
<td>2.946</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>14.639</td>
<td>48.089</td>
<td>37.272</td>
<td>6.4</td>
<td></td>
</tr>
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</table>
Valves

Leak frequencies by flow (leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Liquid</th>
<th>Gas</th>
<th>Two-phase</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.056000</td>
<td>0.011850</td>
<td>0.062100</td>
<td>0.129950</td>
<td>44.8</td>
</tr>
<tr>
<td>1 - 50</td>
<td>0.017420</td>
<td>0.010215</td>
<td>0.015315</td>
<td>0.033956</td>
<td>12.3</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>0.000173</td>
<td>0.000006</td>
<td>0.000196</td>
<td>0.000374</td>
<td>0.12</td>
</tr>
<tr>
<td>Total</td>
<td>0.073593</td>
<td>0.012876</td>
<td>0.077811</td>
<td>0.164280</td>
<td>7.8</td>
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</table>

Fire frequencies by flow (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Liquid</th>
<th>Gas</th>
<th>Two-phase</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>5.600</td>
<td>1.186</td>
<td>6.399</td>
<td>13.185</td>
<td>40.2</td>
</tr>
<tr>
<td>1 - 50</td>
<td>3.558</td>
<td>0.326</td>
<td>5.325</td>
<td>9.209</td>
<td>6.6</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>0.138</td>
<td>0.015</td>
<td>0.588</td>
<td>0.741</td>
<td>53.2</td>
</tr>
<tr>
<td>Total</td>
<td>9.296</td>
<td>1.527</td>
<td>12.313</td>
<td>23.136</td>
<td></td>
</tr>
</tbody>
</table>

Explosion frequencies (explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Gas</th>
<th>Two-phase</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.015</td>
<td>0.113</td>
<td>0.128</td>
<td>18.091</td>
</tr>
<tr>
<td>1 - 50</td>
<td>0.022</td>
<td>0.404</td>
<td>0.427</td>
<td>60.527</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>0.004</td>
<td>0.147</td>
<td>0.151</td>
<td>21.381</td>
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</table>
Leak frequencies by flow (leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>0.383700</td>
<td>0.038370</td>
<td>0.000000</td>
<td>0.422070</td>
<td>42.6</td>
</tr>
<tr>
<td>Gas</td>
<td>0.111210</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.111210</td>
<td>11.2</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.415200</td>
<td>0.041520</td>
<td>0.000000</td>
<td>0.456720</td>
<td>46.1</td>
</tr>
<tr>
<td>Total</td>
<td>0.910110</td>
<td>0.079890</td>
<td>0.000000</td>
<td>0.990000</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>91.930</td>
<td>8.070</td>
<td>0.000</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Fire frequencies by flow (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>38.370</td>
<td>6.880</td>
<td>0.000</td>
<td>45.250</td>
<td>42.2</td>
</tr>
<tr>
<td>Gas</td>
<td>11.121</td>
<td>0.000</td>
<td>0.000</td>
<td>11.121</td>
<td>10.4</td>
</tr>
<tr>
<td>Two-phase</td>
<td>41.520</td>
<td>9.435</td>
<td>0.000</td>
<td>50.955</td>
<td>47.5</td>
</tr>
<tr>
<td>Total</td>
<td>91.011</td>
<td>16.315</td>
<td>0.000</td>
<td>107.326</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>84.799</td>
<td>15.201</td>
<td>0.000</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Explosion frequencies (explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>0.090</td>
<td>0.000</td>
<td>0.000</td>
<td>0.090</td>
<td>7.7</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.663</td>
<td>0.418</td>
<td>0.000</td>
<td>1.081</td>
<td>92.3</td>
</tr>
<tr>
<td>Total</td>
<td>0.753</td>
<td>0.418</td>
<td>0.000</td>
<td>1.171</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>64.310</td>
<td>35.690</td>
<td>0.000</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Pumps

Leak frequencies by flow (leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s(^{-1}))</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>0.033000</td>
<td>0.023430</td>
<td>0.000180</td>
<td>0.056610</td>
<td>68.0</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.012000</td>
<td>0.014520</td>
<td>0.000120</td>
<td>0.026640</td>
<td>32.0</td>
</tr>
<tr>
<td>Total</td>
<td>0.045000</td>
<td>0.037950</td>
<td>0.000300</td>
<td>0.083250</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>54.054</td>
<td>45.586</td>
<td>0.360</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Fire frequencies by flow (fires/plant/year \(\times 10000\))

<table>
<thead>
<tr>
<th>Flow (kg.s(^{-1}))</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>3.300</td>
<td>4.817</td>
<td>0.144</td>
<td>8.261</td>
<td>58.9</td>
</tr>
<tr>
<td>Two-phase</td>
<td>1.200</td>
<td>4.203</td>
<td>0.360</td>
<td>5.763</td>
<td>41.1</td>
</tr>
<tr>
<td>Total</td>
<td>4.500</td>
<td>9.020</td>
<td>0.504</td>
<td>14.024</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>32.089</td>
<td>64.317</td>
<td>3.594</td>
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<td></td>
</tr>
</tbody>
</table>

Explosion frequencies (explosions/plant/year \(\times 10000\))

<table>
<thead>
<tr>
<th>Flow (kg.s(^{-1}))</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Two-phase</td>
<td>0.027</td>
<td>0.281</td>
<td>0.090</td>
<td>0.398</td>
<td>100.0</td>
</tr>
<tr>
<td>Total</td>
<td>0.027</td>
<td>0.281</td>
<td>0.090</td>
<td>0.398</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>6.740</td>
<td>70.639</td>
<td>22.620</td>
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</tr>
</tbody>
</table>
Small-bore connections

Leak frequencies by flow (leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>0.280000</td>
<td>0.028000</td>
<td>0.000000</td>
<td>0.308000</td>
<td>40.0</td>
</tr>
<tr>
<td>Gas</td>
<td>0.115500</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.115500</td>
<td>15.0</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.315000</td>
<td>0.031500</td>
<td>0.000000</td>
<td>0.346500</td>
<td>45.0</td>
</tr>
<tr>
<td>Total</td>
<td>0.710500</td>
<td>0.059500</td>
<td>0.000000</td>
<td>0.770000</td>
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</tr>
<tr>
<td>Per cent</td>
<td>92.273</td>
<td>7.727</td>
<td>0.000</td>
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<td></td>
</tr>
</tbody>
</table>

Fire frequencies by flow (fires/plant/year x 100000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>28.000</td>
<td>5.328</td>
<td>0.000</td>
<td>33.328</td>
<td>39.6</td>
</tr>
<tr>
<td>Gas</td>
<td>11.550</td>
<td>0.000</td>
<td>0.000</td>
<td>11.550</td>
<td>13.7</td>
</tr>
<tr>
<td>Two-phase</td>
<td>31.500</td>
<td>7.818</td>
<td>0.000</td>
<td>39.318</td>
<td>46.7</td>
</tr>
<tr>
<td>Total</td>
<td>71.050</td>
<td>13.145</td>
<td>0.000</td>
<td>84.195</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>84.387</td>
<td>15.613</td>
<td>0.000</td>
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<td></td>
</tr>
</tbody>
</table>

Explosion frequencies (explosions/plant/year x 100000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>0.085</td>
<td>0.000</td>
<td>0.000</td>
<td>0.085</td>
<td>9.3</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.473</td>
<td>0.362</td>
<td>0.000</td>
<td>0.834</td>
<td>90.7</td>
</tr>
<tr>
<td>Total</td>
<td>0.558</td>
<td>0.362</td>
<td>0.000</td>
<td>0.920</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>60.683</td>
<td>39.317</td>
<td>0.000</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
### Leak frequencies by flow (Leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.422175</td>
<td>0.129950</td>
<td>0.910110</td>
<td>0.045000</td>
<td>0.710500</td>
<td>2.217735</td>
<td>21.2</td>
</tr>
<tr>
<td>1 - 50</td>
<td>0.114943</td>
<td>0.033956</td>
<td>0.079890</td>
<td>0.037950</td>
<td>0.059500</td>
<td>0.326239</td>
<td>30.2</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>0.003175</td>
<td>0.000374</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.0003849</td>
<td>0.151</td>
</tr>
</tbody>
</table>

### Fire frequencies by flow (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>42.714</td>
<td>13.185</td>
<td>91.011</td>
<td>4.500</td>
<td>71.050</td>
<td>222.460</td>
<td>25.9</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>5.823</td>
<td>0.741</td>
<td>0.000</td>
<td>0.504</td>
<td>0.000</td>
<td>7.068</td>
<td>0.816</td>
</tr>
</tbody>
</table>

### Explosion frequencies (Explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.431</td>
<td>0.128</td>
<td>0.735</td>
<td>0.027</td>
<td>0.538</td>
<td>1.897</td>
<td>48.0</td>
</tr>
<tr>
<td>1 - 50</td>
<td>1.417</td>
<td>0.427</td>
<td>0.418</td>
<td>0.281</td>
<td>0.362</td>
<td>2.904</td>
<td>11.5</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>1.098</td>
<td>0.151</td>
<td>0.000</td>
<td>0.090</td>
<td>0.000</td>
<td>1.339</td>
<td>6.5</td>
</tr>
</tbody>
</table>

---

**Summary tables - equipment totals**

Leak frequencies by flow (Leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.422175</td>
<td>0.129950</td>
<td>0.910110</td>
<td>0.045000</td>
<td>0.710500</td>
<td>2.217735</td>
<td>21.2</td>
</tr>
<tr>
<td>1 - 50</td>
<td>0.114943</td>
<td>0.033956</td>
<td>0.079890</td>
<td>0.037950</td>
<td>0.059500</td>
<td>0.326239</td>
<td>30.2</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>0.003175</td>
<td>0.000374</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.0003849</td>
<td>0.151</td>
</tr>
</tbody>
</table>

Fire frequencies by flow (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>42.714</td>
<td>13.185</td>
<td>91.011</td>
<td>4.500</td>
<td>71.050</td>
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<td>25.9</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>5.823</td>
<td>0.741</td>
<td>0.000</td>
<td>0.504</td>
<td>0.000</td>
<td>7.068</td>
<td>0.816</td>
</tr>
</tbody>
</table>

Explosion frequencies (Explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.431</td>
<td>0.128</td>
<td>0.735</td>
<td>0.027</td>
<td>0.538</td>
<td>1.897</td>
<td>48.0</td>
</tr>
<tr>
<td>1 - 50</td>
<td>1.417</td>
<td>0.427</td>
<td>0.418</td>
<td>0.281</td>
<td>0.362</td>
<td>2.904</td>
<td>11.5</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>1.098</td>
<td>0.151</td>
<td>0.000</td>
<td>0.090</td>
<td>0.000</td>
<td>1.339</td>
<td>6.5</td>
</tr>
</tbody>
</table>

---

**Summary tables - equipment totals**

Leak frequencies by flow (Leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.422175</td>
<td>0.129950</td>
<td>0.910110</td>
<td>0.045000</td>
<td>0.710500</td>
<td>2.217735</td>
<td>21.2</td>
</tr>
<tr>
<td>1 - 50</td>
<td>0.114943</td>
<td>0.033956</td>
<td>0.079890</td>
<td>0.037950</td>
<td>0.059500</td>
<td>0.326239</td>
<td>30.2</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>0.003175</td>
<td>0.000374</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.0003849</td>
<td>0.151</td>
</tr>
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</table>

Fire frequencies by flow (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>42.714</td>
<td>13.185</td>
<td>91.011</td>
<td>4.500</td>
<td>71.050</td>
<td>222.460</td>
<td>25.9</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>5.823</td>
<td>0.741</td>
<td>0.000</td>
<td>0.504</td>
<td>0.000</td>
<td>7.068</td>
<td>0.816</td>
</tr>
</tbody>
</table>

Explosion frequencies (Explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>Pipes</th>
<th>Valves</th>
<th>Flanges</th>
<th>Pumps</th>
<th>Small connections</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>&lt; 1</td>
<td>0.431</td>
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<td>0.735</td>
<td>0.027</td>
<td>0.538</td>
<td>1.897</td>
<td>48.0</td>
</tr>
<tr>
<td>1 - 50</td>
<td>1.417</td>
<td>0.427</td>
<td>0.418</td>
<td>0.281</td>
<td>0.362</td>
<td>2.904</td>
<td>11.5</td>
</tr>
<tr>
<td>&gt; 50</td>
<td>1.098</td>
<td>0.151</td>
<td>0.000</td>
<td>0.090</td>
<td>0.000</td>
<td>1.339</td>
<td>6.5</td>
</tr>
</tbody>
</table>
### Summary tables - phase breakdown

#### Liquid phase

**Leak frequency by flow (Leaks/plant/year)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.17775</td>
<td>0.057240</td>
<td>0.001482</td>
<td>0.23472</td>
<td>21.6</td>
</tr>
<tr>
<td>Valves</td>
<td>0.05600</td>
<td>0.017420</td>
<td>0.000173</td>
<td>0.07359</td>
<td>6.7</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.38370</td>
<td>0.038370</td>
<td>0.000000</td>
<td>0.42207</td>
<td>38.5</td>
</tr>
<tr>
<td>Pumps</td>
<td>0.03300</td>
<td>0.023430</td>
<td>0.000180</td>
<td>0.05610</td>
<td>5.2</td>
</tr>
<tr>
<td>Small</td>
<td>0.28000</td>
<td>0.028000</td>
<td>0.000000</td>
<td>0.30800</td>
<td>28.1</td>
</tr>
<tr>
<td>Total</td>
<td>0.93045</td>
<td>0.164460</td>
<td>0.001835</td>
<td>1.09674</td>
<td>100.0</td>
</tr>
<tr>
<td>Percent</td>
<td>84.837</td>
<td>14.995</td>
<td>0.167</td>
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<td></td>
</tr>
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</table>

**Fire frequency by flow (Fires/plant/year x 10000)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>17.775</td>
<td>11.829</td>
<td>1.176</td>
<td>30.780</td>
<td>24.3</td>
</tr>
<tr>
<td>Valves</td>
<td>5.600</td>
<td>3.558</td>
<td>0.138</td>
<td>9.276</td>
<td>7.3</td>
</tr>
<tr>
<td>Flanges</td>
<td>38.370</td>
<td>6.880</td>
<td>0.000</td>
<td>45.250</td>
<td>35.7</td>
</tr>
<tr>
<td>Pumps</td>
<td>3.300</td>
<td>4.817</td>
<td>0.144</td>
<td>8.261</td>
<td>6.5</td>
</tr>
<tr>
<td>Small</td>
<td>28.000</td>
<td>5.328</td>
<td>0.000</td>
<td>33.328</td>
<td>26.3</td>
</tr>
<tr>
<td>Total</td>
<td>93.045</td>
<td>32.411</td>
<td>1.459</td>
<td>126.914</td>
<td></td>
</tr>
<tr>
<td>Percent</td>
<td>73.313</td>
<td>25.538</td>
<td>1.149</td>
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<td></td>
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</tbody>
</table>
Gas phase

Leak frequencies by flow (Leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.0468</td>
<td>0.00392</td>
<td>0.000021</td>
<td>0.050741</td>
<td>17.5</td>
</tr>
<tr>
<td>Valves</td>
<td>0.01185</td>
<td>0.001021</td>
<td>0.000005</td>
<td>0.012876</td>
<td>4.4</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.111210</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.111210</td>
<td>38.3</td>
</tr>
<tr>
<td>Small connections</td>
<td>0.115500</td>
<td>0.000000</td>
<td>0.000000</td>
<td>0.115500</td>
<td>39.8</td>
</tr>
<tr>
<td>Total</td>
<td>0.285360</td>
<td>0.004941</td>
<td>0.000026</td>
<td>0.290327</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>98.289</td>
<td>1.702</td>
<td>0.009</td>
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<td></td>
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</table>

Fire frequencies by flow (Fires/plant/year x 100000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>4.683</td>
<td>1.404</td>
<td>0.062</td>
<td>6.150</td>
<td>20.3</td>
</tr>
<tr>
<td>Valves</td>
<td>1.186</td>
<td>0.326</td>
<td>0.015</td>
<td>1.527</td>
<td>5.0</td>
</tr>
<tr>
<td>Flanges</td>
<td>11.121</td>
<td>0.000</td>
<td>0.000</td>
<td>11.121</td>
<td>36.6</td>
</tr>
<tr>
<td>Small connections</td>
<td>11.5500</td>
<td>0.000</td>
<td>0.000</td>
<td>11.550</td>
<td>38.1</td>
</tr>
<tr>
<td>Total</td>
<td>28.540</td>
<td>1.731</td>
<td>0.077</td>
<td>30.348</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>94.044</td>
<td>5.703</td>
<td>0.253</td>
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</table>

Explosion frequencies (Explosions/plant/year x 100000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.058</td>
<td>0.114</td>
<td>0.015</td>
<td>0.187</td>
<td>46.4</td>
</tr>
<tr>
<td>Valves</td>
<td>0.015</td>
<td>0.022</td>
<td>0.004</td>
<td>0.041</td>
<td>10.2</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.090</td>
<td>0.000</td>
<td>0.000</td>
<td>0.090</td>
<td>22.3</td>
</tr>
<tr>
<td>Small connections</td>
<td>0.085</td>
<td>0.000</td>
<td>0.000</td>
<td>0.085</td>
<td>21.2</td>
</tr>
<tr>
<td>Total</td>
<td>0.247946</td>
<td>0.136595</td>
<td>0.019219</td>
<td>0.403759</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>61.409</td>
<td>33.831</td>
<td>4.760</td>
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<td></td>
</tr>
</tbody>
</table>
### Two-phase leaks

- Leak frequencies by flow (Leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.197625</td>
<td>0.053783</td>
<td>0.001672</td>
<td>0.253080</td>
<td>21.8</td>
</tr>
<tr>
<td>Valves</td>
<td>0.062100</td>
<td>0.015515</td>
<td>0.000196</td>
<td>0.077811</td>
<td>6.7</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.415200</td>
<td>0.041520</td>
<td>0.000000</td>
<td>0.456720</td>
<td>39.3</td>
</tr>
<tr>
<td>Pumps</td>
<td>0.012000</td>
<td>0.014520</td>
<td>0.000120</td>
<td>0.026640</td>
<td>2.3</td>
</tr>
<tr>
<td>Small</td>
<td>0.315000</td>
<td>0.051500</td>
<td>0.000000</td>
<td>0.366500</td>
<td>29.9</td>
</tr>
<tr>
<td>Connections</td>
<td></td>
<td></td>
<td></td>
<td>1.001925</td>
<td></td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td></td>
<td></td>
<td>1.160751</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>86.317</td>
<td>13.512</td>
<td></td>
<td>0.171</td>
<td></td>
</tr>
</tbody>
</table>

- Fire frequencies by flow (Fires/plant/year x 10000)

<table>
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<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>20.256</td>
<td>18.098</td>
<td>4.583</td>
<td>42.939</td>
<td>28.4</td>
</tr>
<tr>
<td>Valves</td>
<td>6.399</td>
<td>5.325</td>
<td>0.588</td>
<td>12.313</td>
<td>8.1</td>
</tr>
<tr>
<td>Flanges</td>
<td>41.520</td>
<td>9.435</td>
<td>0.000</td>
<td>50.955</td>
<td>33.7</td>
</tr>
<tr>
<td>Pumps</td>
<td>1.200</td>
<td>4.203</td>
<td>0.360</td>
<td>5.763</td>
<td>3.8</td>
</tr>
<tr>
<td>Small</td>
<td>31.500</td>
<td>7.818</td>
<td>0.000</td>
<td>39.318</td>
<td>26.0</td>
</tr>
<tr>
<td>Connections</td>
<td></td>
<td></td>
<td></td>
<td>100.875</td>
<td></td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td></td>
<td></td>
<td>151.287</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>66.678</td>
<td>29.663</td>
<td>3.657</td>
<td></td>
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</tr>
</tbody>
</table>

- Explosion frequencies (Explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.374</td>
<td>1.302</td>
<td>1.082</td>
<td>2.758</td>
<td>48.1</td>
</tr>
<tr>
<td>Valves</td>
<td>0.113</td>
<td>0.404</td>
<td>0.147</td>
<td>0.664</td>
<td>11.6</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.663</td>
<td>0.418</td>
<td>0.000</td>
<td>1.081</td>
<td>19.9</td>
</tr>
<tr>
<td>Pumps</td>
<td>0.027</td>
<td>0.281</td>
<td>0.090</td>
<td>0.398</td>
<td>6.9</td>
</tr>
<tr>
<td>Small</td>
<td>0.473</td>
<td>0.362</td>
<td>0.000</td>
<td>0.834</td>
<td>14.5</td>
</tr>
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<td>Connections</td>
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<td></td>
<td></td>
<td>1.648973</td>
<td></td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td></td>
<td></td>
<td>5.735909</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>28.748</td>
<td>48.248</td>
<td>23.004</td>
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</tr>
</tbody>
</table>

G25
Summary tables - phase totals

Leak frequencies (leaks/plant/year)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>0.93045</td>
<td>0.164460</td>
<td>0.001835</td>
<td>1.096745</td>
<td>43.0</td>
</tr>
<tr>
<td>Gas</td>
<td>0.285360</td>
<td>0.004941</td>
<td>0.000026</td>
<td>0.290327</td>
<td>11.4</td>
</tr>
<tr>
<td>Two-phase</td>
<td>1.001925</td>
<td>0.156838</td>
<td>0.001988</td>
<td>1.160751</td>
<td>45.6</td>
</tr>
<tr>
<td>Total</td>
<td>2.217735</td>
<td>0.326239</td>
<td>0.003849</td>
<td>2.547823</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>87.044</td>
<td>12.805</td>
<td>0.151</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Fire frequency (fires/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>93.045</td>
<td>32.411</td>
<td>1.459</td>
<td>126.914</td>
<td>41.1</td>
</tr>
<tr>
<td>Gas</td>
<td>26.540</td>
<td>1.731</td>
<td>0.077</td>
<td>30.348</td>
<td>9.8</td>
</tr>
<tr>
<td>Two-phase</td>
<td>100.875</td>
<td>44.879</td>
<td>5.533</td>
<td>151.287</td>
<td>49.0</td>
</tr>
<tr>
<td>Total</td>
<td>222.460</td>
<td>79.021</td>
<td>7.068</td>
<td>308.549</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>72.099</td>
<td>25.610</td>
<td>2.291</td>
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<td></td>
</tr>
</tbody>
</table>

Explosion frequency (explosions/plant/year x 10000)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>0.248</td>
<td>0.137</td>
<td>0.019</td>
<td>0.404</td>
<td>6.6</td>
</tr>
<tr>
<td>Two-phase</td>
<td>1.649</td>
<td>2.767</td>
<td>1.319</td>
<td>5.736</td>
<td>93.4</td>
</tr>
<tr>
<td>Total</td>
<td>1.897</td>
<td>2.904</td>
<td>1.339</td>
<td>6.140</td>
<td></td>
</tr>
<tr>
<td>Per cent</td>
<td>30.896</td>
<td>47.299</td>
<td>21.804</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

STOP
r 23:53 7.21562
Appendix H. Criteria for the definition of zone categories

A typical method of distinguishing between areas of varying flammable risk is to categorise the areas according to the number of hours per year that a flammable atmosphere may be expected to exist in each such that:

<table>
<thead>
<tr>
<th>Zone</th>
<th>Time of occupancy for flammable atmosphere per year</th>
</tr>
</thead>
<tbody>
<tr>
<td>Zone 0</td>
<td>&gt; 1000 hours</td>
</tr>
<tr>
<td>Zone 1</td>
<td>10 - 1000 hours</td>
</tr>
<tr>
<td>Zone 2</td>
<td>1 - 10 hours</td>
</tr>
</tbody>
</table>

A non-hazardous area is defined as not being Zone 0, 1 or 2. This categorisation is used by both the ICI/RoSPA code and the British Gas Engineering Standard PS/SHA 1 which were both listed in Appendix A. The method by which the above figures were arrived at was described by Parvin [120] and formulated with regard to the criteria listed below.

H.1 Target hazard rate

Parvin set a target FAR of $3.0 \times 10^{-5}$ fatalities per exposed man.year for each main class of hazard applied to the personnel at greatest risk, which in this case was the risk of ignition of flammable materials. For the each of the personnel at greatest risk Parvin defined $P_p$ as the probability of being in the area affected by the possible ignition of flammable materials and $P_K$ as the probability of being killed if present in the area affected by the ignition. The target hazard rate, in other words the minimum allowable frequency for the ignition of flammable materials was then calculated as:

$$3.0 \times 10^{-5} \times \frac{1}{P_p} \frac{1}{P_K} \text{ occurences.year}^{-1}$$ (H.1)
Parvin assumed that the personnel at greatest risk from flammable ignition would spend their entire time within the affected area and also assumed a typical value of 0.3 for $P_k$. From (H.1) the target hazard rate was calculated by Parvin as:

$$3.0 \times 10^{-5} \times \frac{1}{0.3} = 1.0 \times 10^{-4} \text{ occurrences.year}^{-1}.$$

H.2 Frequency of ignition

Parvin defined ignition frequency $F_{SA}$ as the frequency of coincidence of a spark and a flammable atmosphere and quantified it thus:

$$F_{SA} = F_S \cdot F_A (D_S + D_A) 
\quad \text{(H.2)}$$

where:

- $F_S = \text{Average frequency of sparking (occurrences.year}^{-1})$
- $D_S = \text{Average effective duration of spark (years.occurrence}^{-1})$
- $F_A = \text{Average frequency for a flammable atmosphere extending into an area in which there is a potential ignition source (occurrences.year}^{-1)}$
- $D_A = \text{Average duration for the flammable atmosphere remaining at the ignition source (years.occurrence}^{-1)}$

$D_S$ is virtually instantaneous and consequently small compared to $D_A$ therefore Parvin reduced equation (H.2) for all practical purposes to:

$$F_{SA} = F_S \times F_A \cdot D_A 
\quad \text{(H.3)}$$

where $F_A D_A$ is the probability of occurrence of a flammable atmosphere.

With these two criteria Parvin defined the zones in terms of the ignition hazard from electrical equipment as follows:

H.3 Non-hazardous areas

Parvin assumed that all non-protected electrical equipment should be regarded as a continuous ignition source. Since the ignition probability was therefore unity, the probability
of a flammable release occurring around the ignition source should not exceed the calculated
hazard rate of $1 \times 10^{-4}$ occurrences.year$^{-1}$ i.e. a flammable atmosphere should be present no
longer than approximately 1 hour per year. Parvin concluded that since this was a very small
frequency it implied that no flammable atmosphere could be tolerated in non-hazardous areas.

H.4 Zone 2 areas

Parvin assumed that in each zone around a leak source there were on average 10 items of
type N electrical equipment (type N is the minimum protection currently required for equipment
operating in a Zone 2 area). The sparking frequency was taken to be once in $10^6$ hours or once
in $10^2$ years. With 10 items this became $0.1$ occurrences.year$^{-1}$. To obtain the target hazard rate
of $1 \times 10^{-4}$ occurrences.year$^{-1}$ the amount of time that a flammable atmosphere is permitted
within a Zone 2 area was calculated as:

$$1.0 \times 10^{-4} \times \frac{1}{10^{-3}} = 10^{-3} \text{ years.year}^{-1}$$

i.e. about 10 hours per year.

H.5 Zone 1 areas

Using the same approach as above the frequency of sparking in flameproof electrical
equipment was taken to be once in $10^8$ hours or $10^4$ occurrences.year$^{-1}$. With 10 items this
became $10^{-3}$ occurrences.year$^{-1}$ for typical process plant. The amount of time that a flammable
atmosphere may be permitted in a Zone 1 area to achieve the target hazard rate was then calculated
as:

$$1.0 \times 10^{-4} \times \frac{1}{10^{-3}} = 10^{-1} \text{ years.year}^{-1}$$

i.e. about 1000 hours per year.
H.6 Zone 0 areas

Parvin stipulated that in areas where flammable atmospheres were likely to occur more than 1000 hours per year then electrical equipment with a sparking frequency considerably less than $10^{-4}$ occurrences.year$^{-1}$ must be used. Parvin concluded that in practice this implied that no electrical equipment should be allowed in a Zone 0 area.
Appendix I. Case study of process plant using fire and explosion model

I.1 Objective

Section 12 described the construction of an ignition model to estimate the ignition and explosion frequency of a section of process plant handling flammable materials and containing both leak and ignition sources. In an initial study the fire and explosion model was used to adjust basic estimates of leak frequency and ignition probability by applying these to a "standard plant". This plant was composed of leak sources in characteristic proportions obtained from inventory data given in Section 5. It was intended to be of average size, so that the overall and intermediate fire and explosion frequencies per plant generated using the model could be multiplied by the number of plants at risk from fire and explosion nationally, to give numbers of ignitions per country and cross-checked against statistical event data. Although reasonable agreement was obtained, it was felt worthwhile to explore the use of the model further, by applying it to the leak source inventory of an actual plant, for the following reasons:

1. It would indicate the ease with which leak source inventory data suitable for use with the fire and explosion model could be generated.

2. It may be expected to highlight which leak sources make the biggest contribution to the expected ignition frequency, and hence suggest how this frequency may be improved upon.

3. By applying the model to an actual plant, there should be no unexpected or peculiar results compared to engineering judgement. This would serve to support the results of the initial study given in Section 12.
Whilst the fire and explosion model is being used here to consider leak source behaviour, its use as a tool for HAC depends on further research into the nature of ignition probability. The requirements of future work are discussed in Section 14.

1.2 Choice of leak source inventory

The leak source inventory data analysed in this appendix were taken from the medium-sized hydrocarbon plant described in Section 5 as Material take-off 1. This plant was selected for the case study for two principal reasons. Firstly, the task of extracting the inventory data (discussed further in Section 1.3) was largely complete, providing a source of data which was relatively easy to use. Note that although the "standard plant" input used in Section 12 was derived from the hydrocarbon plant, the two data sets are not exactly alike; for example, the standard plant was postulated to contain more large diameter components. The intention was to reflect the idea that it may represent a section of a much larger plant e.g. a refinery, rather than the self-contained process plant described by Material take-off 1. For this reason it was also posited that the standard plant would handle a higher proportion of flashing liquids than was evident from Material take-off 1.

The second reason for selecting the hydrocarbon plant as the subject for a case study was because it was necessary to use a plant with a similar ignition source density to the standard plant. It was clear from Section 11 that the ignition probability of a release is correlated to the ignition source density around the release source as well as the leak flow rate. The fire and explosion model does not yet account for differences in this variable, instead it assumes an ignition source density corresponding to the average of a mix of zoning practices. It would thus be meaningless to apply the model as given in Appendix F to a plant that was very different to the standard plant. The hydrocarbon plant was therefore convenient, as the plant was in fact zoned in accordance with a proprietary method that was in effect a mixture of the UK HAC codes of practice listed in Appendix A.

1.3 Formulation of inventory input to fire and explosion model

One of the reasons for performing the case study was to illustrate how the fire and explosion model may be used by an engineer wishing to perform an ignition analysis of a given...
plant. This section describes the compilation of data necessary to provide input to the model, and discusses how this can be made more convenient. Finally, this section describes how the leak source inventory data for the hydrocarbon plant were sorted to provide input for the fire and explosion model.

The most onerous task to be performed in order to be able to use the model on a plant is the compilation of the plant leak source inventory. To produce the hydrocarbon plant leak source inventory it was necessary to sort through the material take-off list for the amounts of pipe, gaskets and valves used in the plant. The equipment specification sheets were then examined, firstly to determine the number of pumps on the plant, and then to use the process operating conditions associated with each item to estimate the proportion of the plant handling flashing liquids, non-flashing liquids and gas. Having assembled the leak source inventory data it was then compiled into lists and ordered according to size. These lists were then encoded into the computer program. The process of manually sorting the leak inventory data was very time-consuming, and the resulting computer model input in the form of ordered lists of component leak sources made the program very cumbersome to alter if it were to be used for the ignition analysis of other, different plants. However, it was apparent that much of the data were already computerised in their initial state (e.g. as stock order lists). It was therefore envisaged that an engineer wishing to use the model with a computerised stock list would find it a relatively simple task to directly extract the relevant information and generate a data spreadsheet which could be used as input to the suitably adapted program. This could also be extended to the extraction of data from equipment specification sheets if these were available in the form of a database compiled as part of the plant documentation.

Turning to the compilation of information necessary for this case study, the principal difference between the input requirements for the leak source inventories of the standard plant and of the hydrocarbon plant was that whereas the standard plant was assumed to only consist of four pipe (and thus component) diameters, the inventory of the hydrocarbon plant consists of sixteen different sizes. The input to the program was therefore essentially derived from Tables 5.9, 5.10 and 5.19. From these tables it was possible to derive the following data directly:

- Total pipe lengths
- Total gasketted connections
- Total valves
- Total process pumps
It was not possible to estimate numbers of small-bore connections from the material take-off, since full knowledge of such items as instrument connections was either not available, or could not be clearly discerned from the documentation. A figure was therefore approximated from the standard plant, which was presumed to contain 700. This figure was scaled using the relative numbers of pumps on both plants (see Section 5.3) to a total of 420. Note that although the number of small-bore connections was not known in this case, this was principally due to unfamiliarity with the plant, rather than obvious lack of data. An engineer with a greater knowledge of the plant’s construction could be expected to provide a more precise estimate.

The component totals were then split according to whether the items handled non-flammable material e.g. reactor coolant (steam), or flammable material. The proportion of components handling flammable material was then further split to estimate the numbers of items handling gas, non-flashing and flashing liquids. These were approximated using the method described in Section 5.4 of taking the equipment item specification sheets and determining the type of fluid being handled by each pump, vessel, heat exchanger, etc. By totalling the numbers of items handling non-flammable material, the overall leak source inventories were reduced by 17% to arrive at the number of components handling flammable material only, as shown in Table 5.19. The relative quantities of each type of leak source handling flammable gas, non-flashing and flashing liquids were estimated to correspond to the ratio of 0.38:0.26:0.38. It was generally not possible to establish a more detailed split, for example the proportion of pipework of each diameter handling flashing fluid could not be discerned. However, it was possible to establish separate phase splits for pumps with respect to size, since the type of fluid being handled by each pump could be inferred from the individual pump data sheet. These were as follows:

<table>
<thead>
<tr>
<th>Pump connection diameter (mm)</th>
<th>50</th>
<th>80</th>
<th>100</th>
<th>150</th>
<th>200</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pumps handling flashing liquid</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>3</td>
</tr>
<tr>
<td>Pumps handling non-flashing liquid</td>
<td>1</td>
<td>4</td>
<td>1</td>
<td>0</td>
<td>1</td>
</tr>
</tbody>
</table>

The process conditions for the hydrocarbon plant were estimated in a similar fashion to the overall phase split described above, by referring to the equipment specification sheets. A process pressure of 15 bar abs. was used for the case study by taking an overall average of the operating pressures for each item, and also an average temperature of 300 K. The physical properties of the process fluids were as estimated for the standard plant i.e. approximately those of propane (propane is the main hydrocarbon feedstock to the plant).
The program was thus run for the hydrocarbon plant using hole size distributions, leak frequency estimates and ignition probabilities generated from the initial iterative study described in Section 12. Note that for the case study it was the final ignition and explosion probabilities, i.e. the probabilities obtained after iteration, that were applied. Similarly, the case study used the iterated values of component leak frequencies. Although leak frequencies for only four component sizes were derived in the analysis of the standard plant, these were assigned to the sixteen component sizes used here as follows. The derived frequency for 25 mm items was applied to take-off items between 15 mm and 40 mm. The derived frequency for 50 mm items was applied to 50 mm and 80 mm take-off items, whilst the derived frequency for 100 mm items was applied to take-off items between 100 mm and 250 mm. Finally, the derived leak frequency for 300 mm items was applied to all take-off items between 300 mm and 600 mm. The leak size distributions remained the same in each case i.e. $A:0.1A:0.01A$ where $A$ is the component cross-sectional area.

1.4 Case study results and discussion

The final output from the ignition analysis of the hydrocarbon plant is presented in Tables I.1 and I.2. These tables correspond to the summary of the ignition analysis of the standard plant given in the tables on pages G22 and G26. If these four sets of tables are compared it is apparent that they are consistent with the change in leak source inventory. In Table I.1 the overall event frequencies are lower for the hydrocarbon plant than for the standard plant due to the slightly lower overall inventories involved, whilst in Table I.2 the change in phase split results in a higher proportion of events involving gas emissions. The most notable difference is the considerable contribution made by pump leaks to the ignition and explosion frequencies for this case study compared to the initial study of the standard plant. This is an effect of the higher proportion of pumps handling flashing flow. Similarly, because the pumps in the case study are larger than for the standard plant, the proportion of minor leaks (i.e. the leaks below 1 kg.s$^{-1}$) is low compared to the number of intermediate leaks. This version of the model, though rather basic to draw firm conclusions with, suggests that it may be appropriate to recommend that effort be directed to reducing the leak frequency of these pumps (given the hole size distribution), possibly by employing barrier seals.

On a more general level, it is noticeable that over 64% of leak incidents (and in this case, ignition incidents) result from pipework, in other words pipes and gasketted connections.
This is similar to the results of the initial study, where approximately 60% of leaks were from pipework. These leak sources differ from pumps and valves for the reason that the leaks from the latter items usually arise as a result of wear incurred during operation, whereas pipes and flanges tend to leak as a result of corrosion and incidents featuring third party interference, for example through inadequate maintenance or direct damage. It can therefore be argued from Table I.1 and I.2 that the leak and ignition frequencies can be significantly reduced by concentrating on measures to improve piping integrity, rather than trying to improve valve and pump reliability.

To summarise, this Appendix has examined how the interim fire and explosion model presented in Section 12 may be applied to actual leak source inventory data, using information presented in Section 5. It was concluded that such an application presented no real problem, although in the general case the application could be considerably simplified if the inventory data were available in the form of a spreadsheet.

The results generated by the ignition analysis of the hydrocarbon plant showed that there is an increased risk of ignitions due to pump leaks involving flashing liquid, compared to the analysis of the standard plant in Section 12. The results from this case study also served to emphasise a point highlighted by the initial study, which was the relatively high risk presented by pipework leaks compared to leaks from seals on valves and pumps.
Table I.2 Case study of process plant: event frequencies by fluid phase

**Leak frequencies (leaks/(plant.year))**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>0.43</td>
<td>0.137</td>
<td>0.00215</td>
<td>0.57</td>
<td>28.6</td>
</tr>
<tr>
<td>Gas</td>
<td>0.68</td>
<td>0.014</td>
<td>0.00011</td>
<td>0.70</td>
<td>35.1</td>
</tr>
<tr>
<td>Two-phase</td>
<td>0.63</td>
<td>0.090</td>
<td>0.00297</td>
<td>0.72</td>
<td>36.4</td>
</tr>
<tr>
<td>Total</td>
<td>1.74</td>
<td>0.24</td>
<td>0.0052</td>
<td>1.99</td>
<td></td>
</tr>
<tr>
<td>Percent</td>
<td>87.6</td>
<td>12.1</td>
<td>0.26'</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Fire frequencies (fires/(plant.year) x 10⁴)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid</td>
<td>43.0</td>
<td>25.5</td>
<td>1.64</td>
<td>70.1</td>
<td>27.7</td>
</tr>
<tr>
<td>Gas</td>
<td>68.5</td>
<td>4.4</td>
<td>0.31</td>
<td>73.2</td>
<td>28.9</td>
</tr>
<tr>
<td>Two-phase</td>
<td>67.4</td>
<td>34.5</td>
<td>7.73</td>
<td>109.6</td>
<td>43.4</td>
</tr>
<tr>
<td>Total</td>
<td>178.8</td>
<td>64.4</td>
<td>9.68</td>
<td>252.8</td>
<td></td>
</tr>
<tr>
<td>Percent</td>
<td>70.7</td>
<td>25.4</td>
<td>3.83</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Explosion frequencies (explosions/(plant.year) x 10⁴)**

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt; 1</th>
<th>1 - 50</th>
<th>&gt; 50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>0.60</td>
<td>0.34</td>
<td>0.073</td>
<td>1.01</td>
<td>15.1</td>
</tr>
<tr>
<td>Two-phase</td>
<td>1.30</td>
<td>2.58</td>
<td>1.776</td>
<td>5.66</td>
<td>84.9</td>
</tr>
<tr>
<td>Total</td>
<td>1.90</td>
<td>2.92</td>
<td>1.85</td>
<td>6.67</td>
<td></td>
</tr>
<tr>
<td>Percent</td>
<td>28.5</td>
<td>43.8</td>
<td>27.7</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Table I.1 Case study of process plant: event frequencies by equipment type

### Leak frequencies (leaks/(plant.year))

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt;1</th>
<th>1-50</th>
<th>&gt;50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.36</td>
<td>0.110</td>
<td>0.0032</td>
<td>0.48</td>
<td>24.0</td>
</tr>
<tr>
<td>Valves</td>
<td>0.18</td>
<td>0.018</td>
<td>0.0003</td>
<td>0.20</td>
<td>10.0</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.75</td>
<td>0.046</td>
<td>0.0</td>
<td>0.80</td>
<td>40.1</td>
</tr>
<tr>
<td>Pumps</td>
<td>0.01</td>
<td>0.040</td>
<td>0.0017</td>
<td>0.05</td>
<td>2.6</td>
</tr>
<tr>
<td>Small connections</td>
<td>0.44</td>
<td>0.027</td>
<td>0.0</td>
<td>0.46</td>
<td>23.2</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>1.74</td>
<td>0.24</td>
<td>0.005</td>
<td>1.99</td>
<td></td>
</tr>
<tr>
<td><strong>Percent</strong></td>
<td>87.6</td>
<td>12.1</td>
<td>0.26</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

### Fire frequencies (fires/(plant.year) x 10⁴)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt;1</th>
<th>1-50</th>
<th>&gt;50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>40.2</td>
<td>30.7</td>
<td>5.88</td>
<td>76.8</td>
<td>30.4</td>
</tr>
<tr>
<td>Valves</td>
<td>18.5</td>
<td>5.2</td>
<td>0.60</td>
<td>24.3</td>
<td>9.6</td>
</tr>
<tr>
<td>Flanges</td>
<td>75.3</td>
<td>10.2</td>
<td>0.0</td>
<td>85.5</td>
<td>33.8</td>
</tr>
<tr>
<td>Pumps</td>
<td>1.18</td>
<td>12.2</td>
<td>3.20</td>
<td>16.6</td>
<td>6.6</td>
</tr>
<tr>
<td>Small connections</td>
<td>43.5</td>
<td>6.0</td>
<td>0.0</td>
<td>49.5</td>
<td>19.6</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>178.8</td>
<td>64.3</td>
<td>9.68</td>
<td>252.8</td>
<td></td>
</tr>
<tr>
<td><strong>Percent</strong></td>
<td>70.7</td>
<td>25.4</td>
<td>3.8</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

### Explosion frequencies (explosions/(plant.year) x 10⁴)

<table>
<thead>
<tr>
<th>Flow (kg.s⁻¹)</th>
<th>&lt;1</th>
<th>1-50</th>
<th>&gt;50</th>
<th>Total</th>
<th>Percent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pipes</td>
<td>0.68</td>
<td>1.57</td>
<td>1.14</td>
<td>3.38</td>
<td>50.7</td>
</tr>
<tr>
<td>Valves</td>
<td>0.17</td>
<td>0.25</td>
<td>0.12</td>
<td>0.54</td>
<td>8.1</td>
</tr>
<tr>
<td>Flanges</td>
<td>0.68</td>
<td>0.31</td>
<td>0.0</td>
<td>0.98</td>
<td>14.7</td>
</tr>
<tr>
<td>Pumps</td>
<td>0.03</td>
<td>0.62</td>
<td>0.60</td>
<td>1.24</td>
<td>18.6</td>
</tr>
<tr>
<td>Small connections</td>
<td>0.35</td>
<td>0.17</td>
<td>0.0</td>
<td>0.52</td>
<td>7.9</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>1.90</td>
<td>2.92</td>
<td>1.85</td>
<td>6.67</td>
<td></td>
</tr>
<tr>
<td><strong>Percent</strong></td>
<td>28.5</td>
<td>43.8</td>
<td>27.7</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>