CLYDE TERMINAL CONVERSION PROJECT

ENVIRONMENTAL IMPACT STATEMENT
Volume 3: Appendix F - I

PREPARED FOR THE SHELL COMPANY OF AUSTRALIA LTD
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CLYDE TERMINAL CONVERSION PROJECT

APPENDIX F

PRELIMINARY HAZARD ASSESSMENT

ASHURST AUSTRALIA

CLYDE TERMINAL CONVERSION PROJECT

THE SHELL COMPANY OF AUSTRALIA PTY LTD

CLYDE REFINERY SITE

PRELIMINARY HAZARD ANALYSIS

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PREPARED FOR	**:**	**Mark Brennan, Partner** **Ashurst Australia**
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Title: **THE SHELL COMPANY OF AUSTRALIA PTY LTD** **CLYDE REFINERY SITE** **PRELIMINARY HAZARD ANALYSIS**	**QA Verified:** F. Wong
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CONTENTS

ABBREVIATIONS

ALARP	As Low As Reasonably Practicable
BLEVE	Boiling Liquid Expanding Vapour Explosion
BoM	Bureau of Meteorology
DPI	NSW Department of Planning and Infrastructure
g/mol	grams per mole
HAZID	Hazard Identification
HEMP	Hazard & Effects Management Process
HIPAP	Hazardous Industry Planning Advisory Paper
kg/s	kilograms per second
km	kilometres
kPa	kilo-Pascals
kW/m^2	kilo-watts per square metre
LOPA	Layer of Protection Analysis
LPG	Liquefied Petroleum Gas
m	metres
m/s	metres per second
m^2	square metres
MA	Major Accident
mg/m^3	milligrams per cubic metre
MHF	Major Hazard Facility
min.	minutes
mm	millimetres
pmpy	per million per year
ppm	parts per million
QRA	Quantitative Risk Assessment
UK	United Kingdom
VCE	Vapour Cloud Explosion

HOLDS

There are no holds in this document.

1.0 SUMMARY

1.1 Introduction

Shell Refining (Australia) Pty Ltd (Shell) proposes to convert the Clyde Refinery to a finished products storage terminal with no production or processing facilities.

Motor gasoline, diesel and jet fuel will be delivered via pipeline from the Gore Bay terminal. Pigging facilities for the pipelines are located north of the pump house No. 2 building. Butane which is used for winter fuel blends will be brought in by road tanker, delivered at the existing gantry and stored in the existing spheres. Product export will be via road tanker using the existing Parramatta terminal, or via the Silverwater and Hunter pipelines or JUHI pipeline to Sydney Airport.

Sherpa Consulting Pty Ltd were retained to prepare a Preliminary Hazard Analysis (PHA) for the proposed Terminal operations at the Clyde Refinery site to determine if it is 'hazardous' and/or 'offensive' in the context of NSW State Environmental Planning Policy No. 33 (SEPP33 - Ref.1).

A separate PHA has been prepared for the associated Gore Bay Terminal site (Ref 2).

1.2 PHA Process

The Department of Planning and Infrastructure (DPI) multi-level risk assessment guideline (Ref.3) was consulted to identify the most appropriate level of risk assessment.

This PHA is based on a Level 3 Risk Assessment where the results are sufficiently quantified to allow an assessment of the offsite risk levels against acceptance criteria.

The risk assessment process and risk acceptance criteria set out in Hazardous Industry Planning Advisory Paper (HIPAP) No. 6 (Ref.4) and HIPAP 4 (Ref.5) were followed.

1.3 Findings

1. The hazards inherent to the proposed development include flammable liquids in bulk storage (ie butane, gasoline, jet fuel, ethanol and diesel) and in packages (kerosene in drums). Loss of containment scenarios include leaks from pipework and fittings, overfill of and leaks from atmospheric tanks and LPG storage spheres, overfill of road tankers and fitting leaks at the tanker loading gantries.
2. The consequence assessment found that off-site impact could occur due to:
    a. Tank roof fire: Tank 90
    b. Tank overfill cascade leading to flash fire/ vapour cloud explosion: all gasoline tanks
    c. Tank bund fires: Tank Farm B , Tank Farm B1 and Tank Farm K
    d. Pipe track pool fires
    e. Pipe track leaks (medium/ large leaks) leading to flash fire/ vapour cloud explosion

f. LPG fires at the storage spheres and at the tanker loading gantry

g. LPG leaks (large only) at the storage spheres and at the tanker loading gantry leading to flash fire/ vapour cloud explosion

h. LPG BLEVE at the storage spheres and at the tanker loading gantry

1.4 Conclusions

1. The hazards associated with the development have been identified and the risks were conservatively assessed and found to be below the NSW Land-Use Planning Risk Tolerability Criteria set by DPI.
2. Shell has in place systems for ensuring the risk is minimised during design and for managing the residual risk associated with the facility when in operation.
3. The proposed facility does not present a significant risk to surrounding land use.
4. In the context of SEPP33, the facility is therefore considered:
   a. 'potentially hazardous' (rather than 'hazardous'); and
   b. 'potentially offensive' (rather than 'offensive').

2.0 INTRODUCTION

2.1 Background

Shell Refining (Australia) Pty Ltd (Shell) proposes to convert the Clyde Refinery to a finished products storage terminal with no production or processing facilities. Motor gasoline, diesel and jet fuel will be delivered via pipeline from the Gore Bay terminal. Pigging facilities for the pipelines are located north of the pump house No. 2 building. Butane which is used for winter fuel blends will be brought in by road tanker, delivered at the existing gantry and stored in the existing spheres. Product export will be via road tanker using the existing Parramatta terminal, or via the Silverwater and Hunter pipelines or JUHI pipeline to Sydney Airport.

Sherpa Consulting Pty Ltd were retained to prepare a Preliminary Hazard Analysis (PHA) for the proposed Terminal operations at the Clyde Refinery site to determine if it is 'hazardous' and/or 'offensive' in the context of NSW State Environmental Planning Policy No. 33 (SEPP33 - Ref.1).

A separate PHA has been prepared for the associated Gore Bay Terminal site (Ref.2).

2.2 Study Objectives

The objectives of this study are to:

1. Conduct a suitable level of PHA for the proposed Clyde Terminal with reference to the NSW Department of Planning and Infrastructure (DPI) Multi-level Risk Assessment guide.
2. Validate the level of risk assessment chosen based on the findings of the PHA.
3. Assess the proposed development against the DPI Land-Use Safety Planning Risk Criteria.
4. Determine if the proposed facility is 'hazardous' and/or 'offensive' in the context of SEPP33.

2.3 Scope of Study

The scope of this study includes the Clyde Terminal; specifically:

- Atmospheric product (gasoline, jet fuel, diesel) storage tanks and bunds.
- Non-LPG product pumps (Pumphouse 2 Area) and pigging facilities.
- Atmospheric underground ethanol storage tank.
- Non-LPG road tanker product loading gantry.
- Gate 1 Warehouse dangerous goods (flammable/combustible).
- Vapour recovery unit.
- Aboveground LPG storage spheres and pumps.
- LPG road tanker unloading gantry.

2.4 Report Overview

This report follows the methodology described in Applying SEPP33, Multi-level Risk Assessment and the Hazardous Industry Planning Advisory Paper (HIPAP) No. 6 Guidelines for Hazard Analysis (Ref.4).

The PHA process assesses the potential impact of the facility on the surrounding area and can be summarised as follows:

- Description of proposed development.
- Hazard Identification.
- Selection of an appropriate level of assessment.
- Analysis of the consequences of a hazard, should it be realised.
- Analysis of the frequency of hazards occurring, noting that the depth of analysis will be dependent on the results of the consequence analysis.
- Calculation of overall risk results.
- Comparison of risk results against NSW Land-Use Planning risk tolerability criteria.
- Discussion of risk management approach.
- Discussion and conclusion on risk levels.

Each stage of the process is reported in this PHA report. Detailed calculations, where they have been undertaken, are contained within the relevant section of this report. The intention is to provide sufficient detail in the report to allow an objective assessment of the risk.

In addition to the above, the Director General's Requirements (DGRs), Ref. 6, contain the following item not usually included in a PHA:

- Address all relevant recommendations arising from the Buncefield incident.

2.5 Limitations

The following limitations apply to this study:

- This study evaluates the immediate (acute) effects to people and asset from the consequences of loss of containment scenarios only. Any potential human health effects are covered in the human health risk assessment shown separately in the Environmental Impact Statement (EIS). Any potential biophysical environmental effects from a loss of containment are addressed in the ecological assessment shown separately in the EIS
- Heat radiation from BLEVEs is typically of short duration and therefore unlikely to cause escalation; however, shrapnel/ projectiles may lead to incident escalation (by impact). Due to the uncertainty in modelling trajectories, the potential effects of projectiles following BLEVE were not estimated.

2.6 Assumptions

The following assumptions have been made in preparing this study:

- Pipe connection failures were modelled as 50 mm release sizes with leak frequency assumed to be one order of magnitude lower than instrument fittings.
- Pump seal leaks were modelled as 10 mm release sizes.
- Pump casing failures were modelled as full-bore release at the pipe flow rate.
- Jet fuel (kerosene) is not capable of generating flammable vapour clouds; it will be handled at ambient conditions.
- Bunds for dangerous goods storage tanks comply with AS1940 (by equivalent level of safety) and hence are assumed to have sufficient integrity to withstand a sudden loss of containment.

3.0 FACILITY DESCRIPTION

3.1 Site Location

Clyde Refinery is located where the Parramatta and Duck Rivers join, 16 km west of Sydney's CBD.

Figure 3.1 shows the Refinery plot and the surrounding topography (source: Department of Lands Geospatial Portal, Ref.7).

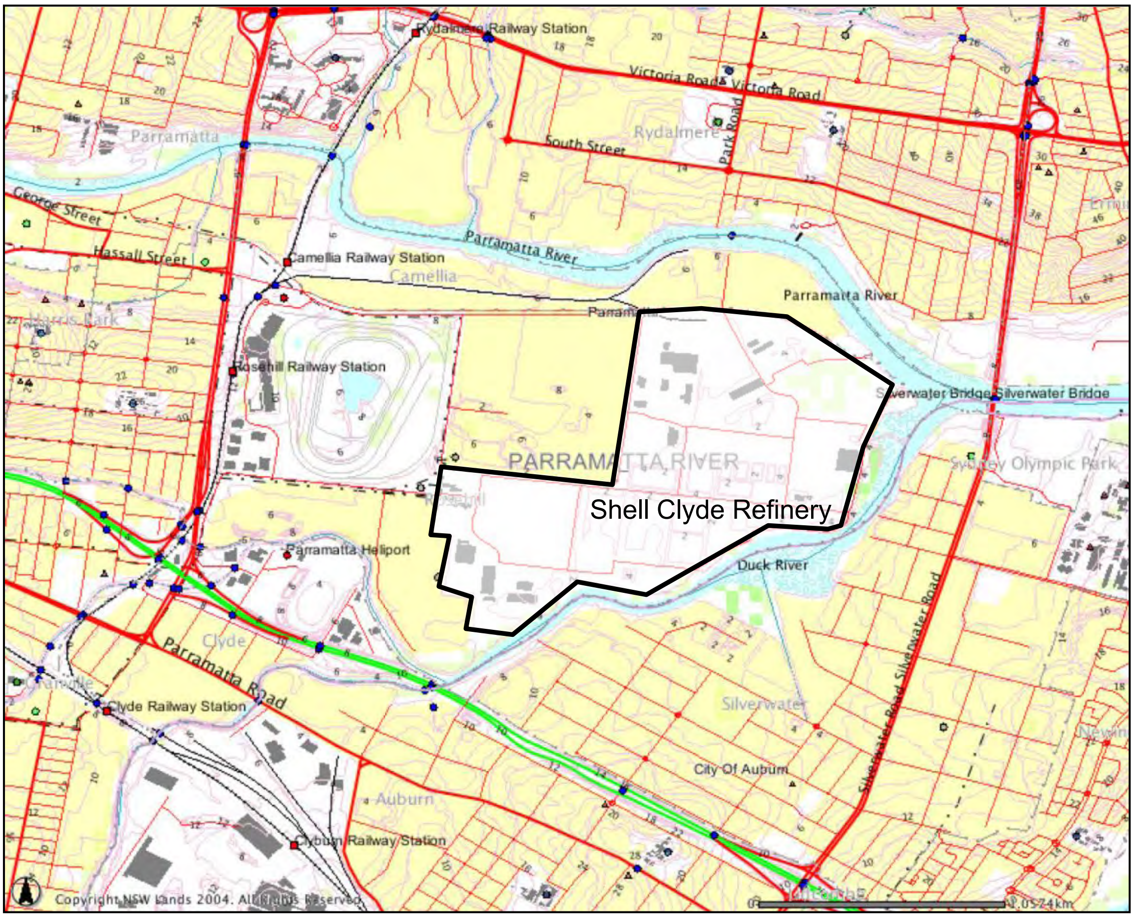

FIGURE 3.1: CLYDE REFINERY AND SURROUNDING TOPOGRAPHY

The part of the Refinery that will be decommissioned, as part of the conversion to Terminal operations, includes process areas and some tank farms, as shown in Figure 3.2. The terminal operations will comprise the remaining areas to the east and north of the Refinery plot.

Figure 3.3 further clarifies the Terminal site boundary and shows the Lyondell Basell site.

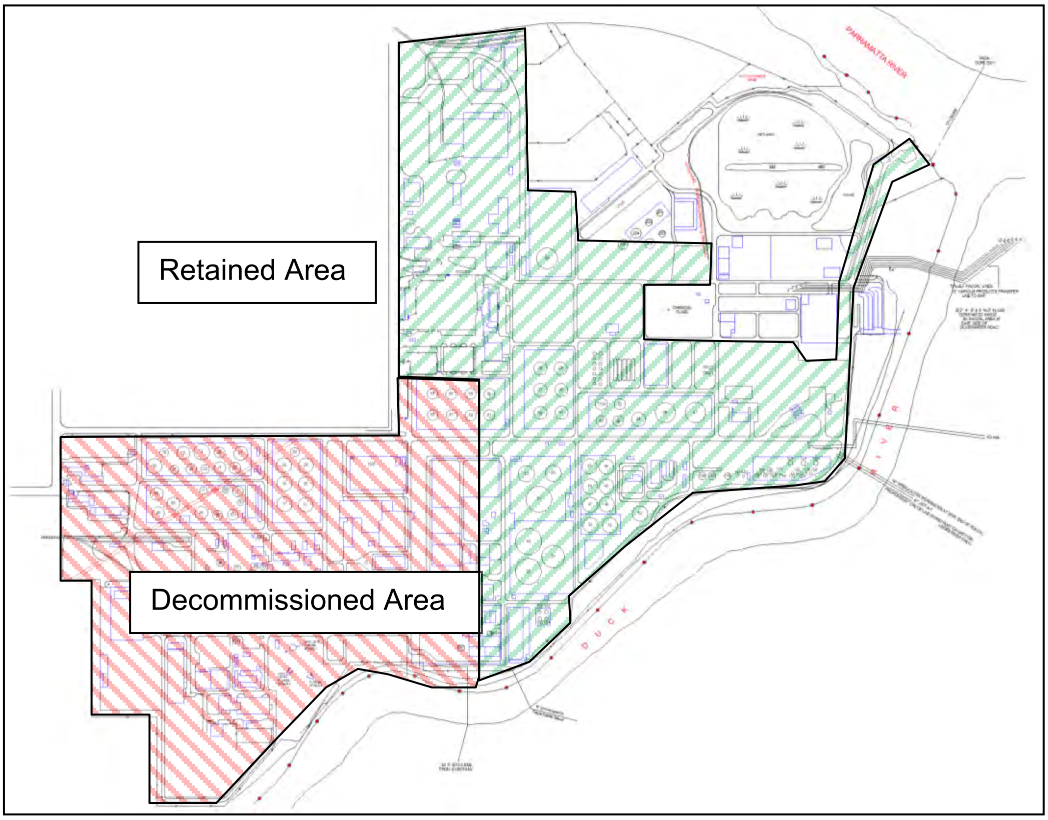

FIGURE 3.2: CLYDE REFINERY DECOMMISSIONED AREAS (APPROXIMATE)

FIGURE 3.3: CLYDE TERMINAL LAYOUT SHOWING LYONDELL BASELL

3.2 General Facility Description

The site will operate as a fuel import and distribution terminal with no production or processing. Motor gasoline, diesel and jet fuel will be delivered via pipeline from the Gore Bay terminal. Pigging facilities for the pipelines are located north of the pump house No. 2 building. Butane and ethanol, which are used for fuel blending will be brought in via tanker and stored in the existing facilities. Product export will be via tankers using the existing Parramatta terminal, or via the Silverwater and Hunter pipelines or JUHI pipeline to Sydney Airport.

Motor gasoline is pumped through filters prior to loading at the Parramatta terminal. The filters, adjacent to T90 will be kept for terminal operations. The pumps (P5030 A/B) are located with the filters.

Storage of fuels will be in 17 existing tanks, refurbished as necessary. Other tanks will be decommissioned with the possibility to use some for firewater storage. Some new pumps may be purchased where existing pumps cannot be used. Pumps will be in the pump house No. 2 area. Pump house No. 2 area includes a pumphouse building and external pumps associated with tanks and movements.

The location of the control room will be at the Old Movements Control Room building. The State Office building will remain.

The LPG road gantry will be used by Shell to unload butane to the existing LPG Spheres (for winter fuel blends).

The existing underground ethanol storage tank (at the Parramatta Terminal) will remain, as will the existing Gate 1 Warehouse, storing packaged Dangerous Goods.

Current bulk lubricants storage tanks in the northern portion of the site will be retained in their current service. These are controlled by Parramatta Terminal.

3.3 Materials Stored

Materials that are to be stored onsite will comprise bulk fuel products (gasoline, diesel, jet fuel, ethanol and butane) and smaller quantities of chemicals (eg for cleaning and lubricants) for site maintenance.

3.3.1 Atmospheric Tank Storage

Table 3.1 summarises the dimensions and the volumes for each of the tanks as given in the site's dangerous goods register. Some of the floating roof tanks may be converted to internal floating roof tanks by the addition of a new cone roof or geodesic dome. The location of tanks and basis for this study are shown in Figure 3.2.

TABLE 3.1: STORAGE TANK CAPACITIES

Tank Farm	Tank No.	Diameter (m)	Height (m)	Tank Roof (Existing)[Note]	Product	Volume (m^3)
K	90	39	22	EFR	Unleaded petrol	26,000
E1	36	24.4	16.5	EFR	Unleaded petrol	7,700
	37	24.4	16.5	EFR	Unleaded petrol	7,700
	38	24.4	16.5	EFR	Unleaded petrol	7,700

Tank Farm	Tank No.	Diameter (m)	Height (m)	Tank Roof (Existing)[Note]	Product	Volume (m^3)
	39	24.4	16.5	EFR	Unleaded petrol	7,700
B	50	34	22	EFR	Unleaded petrol	20,000
	51	49	22	EFR	Diesel	41,000
	53	34	22	EFR	Unleaded petrol	21,000
B1	34	39	13	EFR	Jet fuel	15,000
	35	44	18	EFR	Jet fuel	28,000
	42	44	18	EFR	Jet fuel	28,000
E2	82	17	13	EFR	Slops	2,500
	84	24	22	CR	Unleaded petrol	10,000
	86	39	22	EFR	Unleaded petrol	26,000
	87	39	22	EFR	Unleaded petrol	26,000
B2	32	36	16	CR	Diesel	16,000
	33	36	16	CR	Diesel	16,000
-	91	6	6	EFR	Slops	179
	92	6	6	EFR	Slops	179
-	103	8	5	CR	Slops	250
	105	8	5	CR	Slops	250

Note: EFR = External floating roof
CR = Cone roof

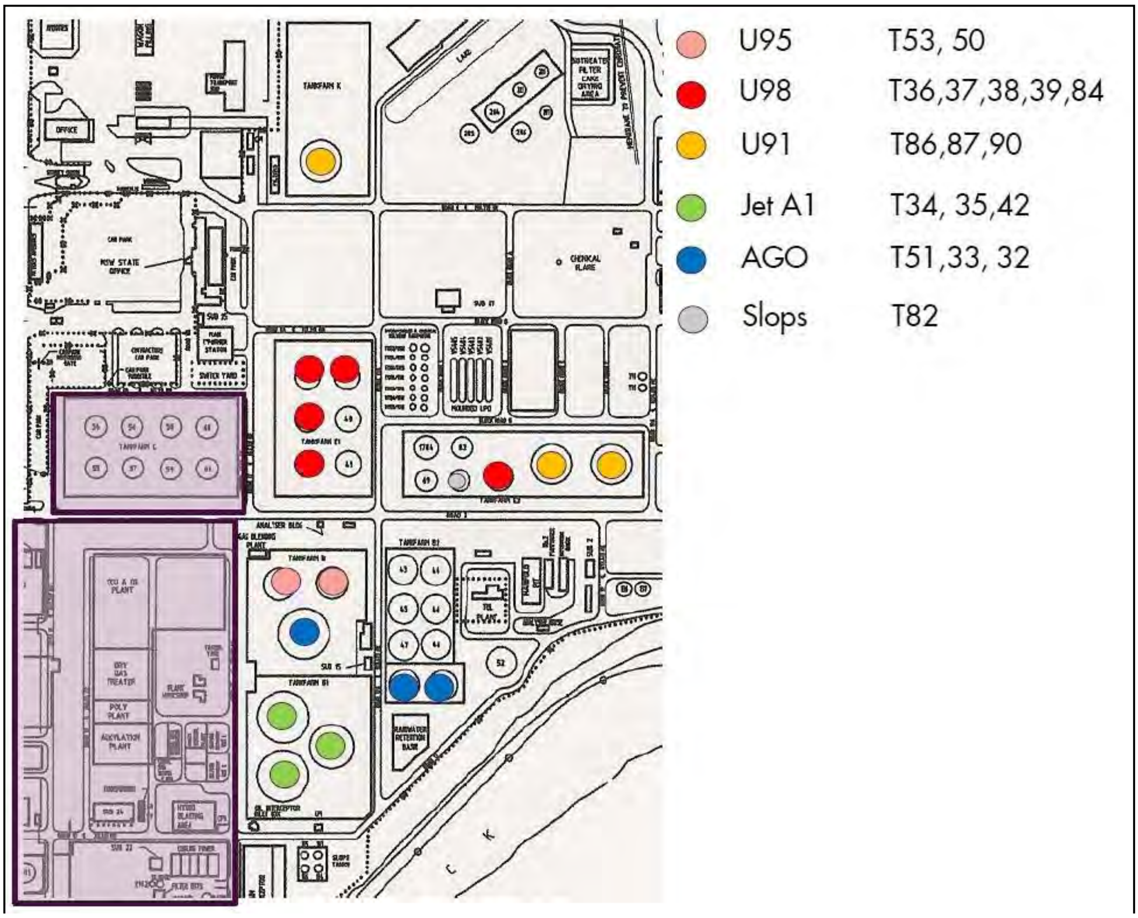

FIGURE 3.4: DESIGNATED ATMOSPHERIC STORAGE TANKS

3.3.2 Butane Sphere Storage

Storage volumes for the spheres are summarised in Table 3.2 based on the capacities in the site's dangerous goods register.

TABLE 3.2: BUTANE SPHERE STORAGE VESSEL CAPACITIES

Location	Vessel	Product	Volume (m^3)
Spheres	V-137	Butane	600
	V-140	Butane	950

3.3.3 Gate 1 Warehouse

The existing warehouse currently located at the Parramatta Terminal will be retained for storage of packed oils and greases, various small quantities of hazardous materials for general cleaning and mechanical items. Package sizes will be 200 L or less.

3.4 Pipeline from Gore Bay Terminal Site

Products will be transferred to the Clyde Terminal site from Gore Bay via an existing pipeline. The pipeline is approximately 18 km long, 300 mm in diameter and operates up to 6,500 kPa(g).

The pipeline is regulated by a NSW WorkCover licence (rather than the Pipelines Act) and is compliant with Australian Standard 2885 (an AS2885 risk assessment is in place).

3.5 Employment and Operating Hours

There will be approximately 30 to 50 people employed at the terminal site and operations will be 24 hours per day. For first response to an incident, there would generally be 3 people available during day shift and 2 during night shift.

3.6 Surrounding Land Uses

The existing Clyde Refinery is zoned, under the Parramatta LEP 2011, as Heavy Industrial.

West of the Clyde Refinery site is zoned *Racecourse*, south-west is zoned as *Private Open Space* and *Industrial* and further from the site is zone *Residential*.

Table 3.3 summarises the land uses adjacent to the existing Refinery site. Other industrial facilities across Durham Street are not included.

TABLE 3.3: LAND-USES ADJACENT TO PROPOSED TERMINAL

Company	Location	Facility and Features
LyondellBasell Australia Pty Ltd	East	Polypropylene plant. Access through refinery Gate No. 4.
SITA	North	Waste management. Fence along boundary
Patrick	North	Shipping container storage
AquaNet Sydney Pty Limited	Northwest at Durham Street and Grand Parade	Water recycling

4.0 INCORPORATING LESSONS FROM THE BUNCEFIELD INCIDENT

This section describes how the Clyde Terminal has addressed the recommendations of the Buncefield Major Incident Investigation Board, including:

- response to the lessons from the Buncefield incident, including actions taken, actions planned and proposed time frames for risk reduction related activities
- consideration of lessons learnt from other major incidents
- consideration of the issues raised in the UK Health and Safety Executive (HSE) publication: *Safety and the environmental standards for fuel storage sites* (Ref.8).

4.1 Response to Lessons from Buncefield Incident

Shell has been represented on working groups that contributed to the UK HSE publication: *Safety and Environmental standards for fuel storage sites* (Ref.8). Communication of the lessons within Shell was via Shell's Learning from Incidents (LFI) Action Awareness and Action Alerts process. In addition, Shell has incorporated the learning into mandatory Process Safety Basic Requirements (PSBRs), and assisted sites in the assessment of risks from Tank Overfill by developing a risk assessment model, known as a Model Bowtie (MBT).

More information on the risk management processes is provided in the Section 4.2; the following paragraphs describe the content of the Tank Overfill LFI (Ref.9) and PSBR (Ref.10) relating to avoiding tank overfill followed by vapour cloud explosion as well as the resultant actions taken or planned by the site.

The Tank Overfill LFI (Ref.9) contained the following recommended actions:

- Conduct a risk assessment for overfilling each tank containing finished gasoline or gasoline components.
- Verify that procedures are in place to require tank operators to validate tank gauging systems, and to proactively manage tank transfers.

The risk assessment that was initially restricted to gasoline or gasoline components was expanded to include evaluation of all storage tanks (including those storing flammable products that do not have the potential for vapour cloud explosion, i.e. jet fuel). The tank overfill MBT (Ref.11) was utilised and a Layer of Protection Analysis (LOPA) was developed and compared with the relevant risk criteria. LOPA was used to confirm the validity of barriers and to ensure sufficient independent barriers were in place. Based on this comparison, actions were proposed to either strengthen existing safeguards or add further safeguards. Shell reviewed the analyses for all affected sites globally, after which projects were developed to implement any required upgrades. A number of minor works were commissioned for Clyde in this respect and detailed with Workcover.

Procedures to validate storage tank gauging systems are in place and as part of the MBT analysis, these were reviewed and records checked. Manual check dips, to verify that the tank gauging system is accurately measuring the level, are carried out on a subset of tanks every month, with each tank checked approximately twice per year.

Independent high level alarms, where installed and accessible, are tested at approximately the same frequency. In the final terminal configuration, the tank gauging and alarm verification and testing for all tanks will be validated monthly.

PSBR 7 includes requirements to create a list of all storage tanks containing fluids that have the potential to overfill resulting in vapour cloud explosion, to assess the risk of each tank and document and implement the resulting remedial steps. This is closely aligned to the activities carried out for the Overfill LFI using the MBT.

The MBTs and LOPAs submitted as part of the site's ALARP Demonstration in response to the NSW Major Hazard Facilities Safety Case submission in 2011 were updated for the End-State Terminal operating mode and included in this PHA for tank overfill frequency analysis (see APPENDIX C). The major change in the analysis was that, as noted in the Development Application, the site no longer stores crude oil.

4.2 Risk Management Processes – Lessons from Major Incidents

Learning from Incidents (LFI) is a process used to alert Shell sites about the causes of significant incidents that have occurred both at Shell and more generally in industries in which Shell operates. A description of the process is shown graphically in Figure 4.1, where it can be seen that after an incident Shell:

- investigates the incident and determines whether it would provide useful information for learning
- shares information with manufacturing sites via LFI alerts and bulletins
- implements the learning by updating documentation and requiring sites to take action as a result of the LFI alerts.

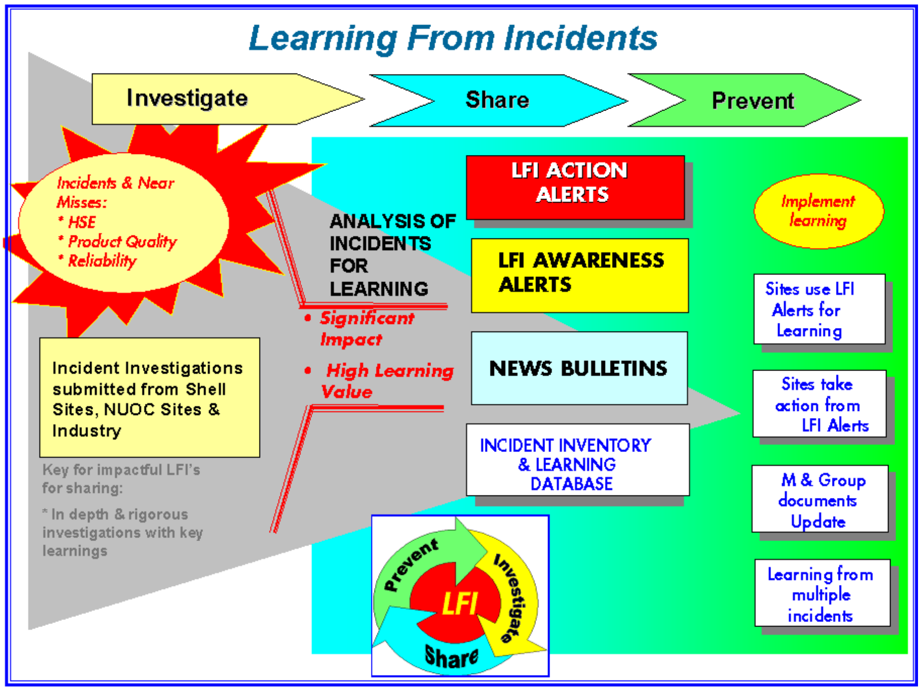

FIGURE 4.1: LEARNING FROM INCIDENTS PROCESS DESCRIPTION

4.3 Response to 'Safety and Environmental Standards for Fuel Storage Sites'

The recommendations from the Buncefield Major Incident Investigation Board have been collated into six key areas of concern, for which detailed guidance is provided in *Safety and the environmental standards for fuel storage sites report* (Ref.8). Much of the guidance is already incorporated in the MBT and the site's Management System; however, during the production of the MHF Safety Report, the information from the six key areas was used.

The first key area relates to safety integrity level assessment, and one method by which to assess the required Safety Integrity Level (SIL) is by Layer of Protection Analysis (LOPA). WorkCover has drawn attention to RR716 (Ref.12) that evaluates LOPA for tank overfill from various companies in the UK, and recommends that:

- Training and knowledge of the LOPA technique be improved
- Better procedures and guidance be produced, including standards of documentation and support information
- the quality of data used in LOPA be improved

As the LOPA technique is used for the tank overfill analyses carried out by Shell, it should be noted that:

- all facilitators have been trained on how to conduct LOPA, in the context of the Shell risk assessment methodology
- a MBT has been used for the tank overfill scenarios that includes comprehensive guidance on the standard of documentation and the analyses were carried out with the on-site assistance from those responsible for the MBT
- Documentation (Ref.13) has been provided to Clyde Terminal that gives guidelines on LOPA, and standard equipment reliability data sets.

5.0 LEVEL OF ASSESSMENT

5.1 Introduction

This study followed the guidelines given in Applying SEPP33 (Ref.1) and Multi Level Risk Assessment (Ref.3).

5.2 Level of Assessment

The DPI document *Multi Level Risk Assessment* was consulted to identify the level of assessment required in the PHA. The DPI document sets out three levels of risk assessment that may be appropriate for a PHA. These are:

Level	Type of Analysis	Comments
1	Qualitative	Where there are no major offsite consequences and societal risk is negligible.
2	Partially Quantitative	Where there are offsite consequences but with a low frequency of occurrence.
3	Quantitative	Where level 1 and 2 are exceeded.

Based on the findings of the HAZID (Section 6.0), a quantitative, Level 3, assessment was undertaken.

6.0 HAZARD IDENTIFICATION

6.1 Overview

The hazard identification (HAZID) involved the following steps:

- Identification of hazardous materials.
- Identification of loss of containment and fire/ explosion scenarios.
- Identification of safeguarding that will be provided.
- Development of specific scenarios to carry forward for assessment.

6.2 Hazardous Materials Stored and Handled

Properties of the materials stored and handled at the site are summarised in Table 6.1. Properties come from the IP Refining Code and Kuchta (Refs. 14 and 15) and are approximate for mixtures. Although large quantities of smoke can be produced from hydrocarbon fires, especially liquids, it rarely leads to dangerous conditions at ground level due to rise of the hot plume and dispersion.

All products are refined with no water content; therefore boilover is not credible for tank fires.

TABLE 6.1: HAZARDOUS PROPERTIES OF MATERIALS

Material	DG Class	UN Number	Hazchem Code	Flash Point (°C)	Auto-ignition Temperature (°C)
Butane/ Iso-butane	2.1	1011 1969 (iC4)	2YE	< -70	370
Gasoline	3 PGII	1203	3YE	< -35	280
Jet fuel	3 PGIII	1863	3Y	> 23	210
Diesel	C1	3082	-	> 60	210
Ethanol	3 PGII	1170	2YE	13	363

6.3 Potential Hazardous Incident Scenarios

The hazard identification word diagram for the site is included in APPENDIX A. The table contains the following information:

- major accident event (resulting in leak or fire)
- cause
- prevention measures
- consequences
- detection measures
- protection measures.

All scenarios listed in APPENDIX A were carried forward for further analysis, with the following exceptions:

- Combustible liquids: diesel has a high flash-point and is handled at ambient temperatures. The chance of ignition and involvement in a fire is remote unless due to an existing fire.
- Rim seal fires: these occur at elevation and the tank shell and wind girder provide shielding to anyone at grade. The study conservatively assumes that all rim seal fires will escalate to full-surface tank roof fires (which, although unlikely, have been considered in this study).

A summary of the scenarios carried forward is provided in Table 6.2.

TABLE 6.2: SCENARIOS CARRIED FORWARD FOR ANALYSIS

Equipment	Scenario	Comments
Atmospheric Storage Tanks and Bunds	Tank roof fire	Ignition of seals (external floating roof tanks) or vents/vapour space (internal floating roof tanks) by lightning.
	Full bund fire	Due to tank overfill, strake/structural catastrophic failure, pipe/flange leak, valve leak, drain leak, floor leak, corrosion.
	Vapour Cloud Explosion/ Flash Fire	A potential outcome of gasoline tank overfill.
Butane Spheres, Gantry and Pumps	Pool fire	Pool fire size based on a distribution of leak rates.
	BLEVE	-
	Vapour Cloud Explosion/ Flash Fire	Unignited pool evaporation.
Pump House No. 2 and Pump Pits	Bund fire	Fire covering full bunded area of pump house.
Pipe Tracks	Pool fire	Fire covering pipe track routes.
Gate 1 Warehouse Package Store	Pool fire	Pool fire size based on bunded area.
Ethanol Tanker Unloading Bay	Bund fire	Pool fire size based on bunded area.
Road Tanker Loading Gantry	Bund fire	Pool fire size based on bunded area.
Vapour Recovery Unit	Pool fire	Pool fire size based on bunded area.

7.0 CONSEQUENCE ASSESSMENT

7.1 Introduction

Consequence analysis involves the analysis and quantification of the potential for a hazardous scenario to cause injury, fatality, damage or loss. The consequence of an incident is assessed independently of the likelihood.

The purpose of the consequence analysis is to determine if the identified hazardous incidents have an offsite impact that exceeds the impairment criteria described in HIPAP 4 (Ref.5).

The distinction between the physical consequences of a release and the effects on people or property are discussed in the subsequent sections.

7.2 Physical Consequence Models

The proprietary modelling package Shell FRED (Fire Radiation Explosion, Dispersion) Version 6.0 was used to quantify the consequences of the identified scenarios using the view factor method.

The following physical models were required for this study:

- Release rate
- Tank overfill cascade leading to vapour cloud formation (see Section 7.2.1)
- Gas dispersion
- Jet fire
- Pool fire
- Vapour cloud explosion/ flash fire
- Boiling Liquid Expanding Vapour Explosion (BLEVE)

The following modelling assumptions were adopted:

- Wind speed: two cases assessed 5 m/s and 2 m/s
- Atmospheric stability: D (neutral) and F (very stable)
- Relative humidity: 70%
- Ambient temperature: 20ºC

The following consequences were analysed for their effects on people and equipment:

- fire heat radiation
- vapour cloud dispersion (including flash fire)
- explosion overpressure

7.2.1 Incorporating the Findings of the Buncefield Incident Investigation

Flammable vapour cloud formation due to tank overfill and subsequent cascade was considered for tanks storing the flammable products described in Appendix 1, Part 2 (Table 6) of the UK HSE's Final Report on the Buncefield Incident (Safety and environmental standards for fuel storage sites) as having the potential to form flammable vapour clouds (Ref.16).

Advice on modelling the overfill cascade and the resulting source term for dispersion modelling is provided in Appendix 1, Part 1 of the UK HSE Report (Ref.16), specifically in the Symposium Series No. 154 Research Paper: Liquid dispersal and vapour production during overfilling incidents. Shell Global Solutions, a co-author of this paper and a party to both the Phase 1 Joint Industry Group and the Phase 2 Technical Group, undertook these specialist analyses on behalf of Shell and provided the gas dispersion results for various wind speed and atmospheric stabilities, which were used in Quantitative Risk Assessment (QRA) models.

The QRA model for the terminal uses the dispersion results to simulate gas spread through the plant and evaluate the potential for ignition in both open, uncongested areas (generating flash fire) and within congested process plant (generating explosion overpressure).

Consistent with the findings of the Buncefield incident, the QRA model includes fatal effects within the extent of the flammable vapour cloud.

7.3 Effects Models

The following consequences were analysed for their effects on people and equipment:

- fire/ fireball heat radiation
- flammable vapour cloud flash fire
- flammable vapour cloud explosion (see note below Table 7.1)

The selection of impairment criteria and modelling techniques is detailed in APPENDIX D and summarised in the subsequent sections.

7.3.1 Effects on People

The impairment criteria for people are summarised in Table 7.1. These values relate to acute effects. Impairment is considered to occur if the levels are equal to or higher than those given in the table.

TABLE 7.1 IMPAIRMENT CRITERIA

Impediment	Effect Criteria
Thermal Radiation	≤ 4.7 kW/m^2 (injury) 4.7 – 14 kW/m^2 (50% chance of fatality) ≥ 14 kW/m^2 (100% chance of fatality)
Flash Fire	100% chance of fatality within flammable vapour cloud defined by LFL concentration
Vapour Cloud Explosion	7 kPa (injury) Fatal explosion overpressure was taken to be within the dimensions of the flash fire (consistent with Buncefield Incident). The study assumes that people within buildings will be fatally injured by a flash fire (conservative), but this is to account for building damage due to explosion overpressure. (see Note 1)

NOTES: 1. For the purpose of calculating the total (location specific) risk contours in Shepherd, the effects on people (in terms of fatalities) from vapour cloud explosion overpressure are accounted for by the fireball consequence size (i.e. personnel within the flash fire are assumed to be fatalities). See APPENDIX D (Section D.1.2) for further discussion.

7.3.2 Effects on Equipment and Structures

Equipment and structures subject to direct flame impingement from fires can weaken with time, from a combination of thermal radiation and convective heating. Eventually failure occurs, resulting in possible escalation of the incident, escape route impairment, and significant plant damage.

It is difficult to assign a specific value for structural failures, since failure is determined by structural characteristics (eg material type, pipe thickness and beam dimensions), handling conditions (whether the equipment is subject to internal pressure) and flame characteristics (surface emissive power, flame dimensions).

Heat-up calculations were undertaken, during the Clyde Refinery Formal Safety Assessment (2000), to estimate failure times of specific critical structures under jet fire loading. The findings of that study (summarised in Table D.3 in APPENDIX D) were verified using proprietary software Vessfire and were carried forward to this study.

The effect of explosion overpressures in the refinery will depend on the location of the explosion and the likely targets; the impairment criteria are summarised in Table D.4 in APPENDIX D.

7.4 Findings

The consequence analysis results are summarised in the subsequent section and in Table 7.2 though to Table 7.9.

The analyses include tank overfill cascade dispersion modelling (i.e. the Buncefield scenario) for gasoline storage tanks only, noting that diesel and jet fuel are of too-low volatility to generate vapour clouds at atmospheric conditions.

TABLE 7.2 TANK ROOF FIRE CONSEQUENCES

Tank No.	Contents	Tank Diameter (m)	Surface Emissive Power (kW/m^2)	Fire/ Radiation Distances (m) *				Escalation Potential	Distance to Boundary (m) +	Off-Site Impact?
				Flame	4.7 kW/m^2	14 kW/m^2	23 kW/m^2			
32	Diesel	36	-	Not a tank-on-fire and escalation unlikely.				-	-	-
33	Diesel	36	-	Not a tank-on-fire and escalation unlikely.				-	-	-
34	Jet fuel	39	22.4	46	64	48	NG	Tanks 35 and 42 depending on wind direction	78	no
35	Jet fuel	44	21.5	50	69	51	NG	Tanks 34 and 42 depending on wind direction	94	no
36	Gasoline	24	30.5	36	50	38	36	Tanks 37 and 39 depending on wind direction	110	no
37	Gasoline	24	30.5	36	50	38	36	Tanks 36 and 38 depending on wind direction	110	no
38	Gasoline	24	30.5	36	50	38	36	Tank 37 depending on wind direction	110	no
39	Gasoline	24	30.5	36	50	38	36	Tank 36 depending on wind direction	150	no
42	Jet fuel	44	21.5	50	69	51	NG	Tanks 34 and 35 depending on wind direction	90	no
50	Gasoline	34	24.0	42	59	44	42	Tanks 51 and 53 depending on wind direction	140	no
51	Diesel	49	-	Not a tank-on-fire and escalation unlikely.				-		-
53	Gasoline	34	24.0	42	59	44	42	Tanks 50 and 51 depending on wind direction	85	no
82	Slops	34	41.8	30	42	32	29	Tank 84 depending on wind direction	190	no
84	Gasoline	17	30.5	36	50	38	36	Tanks 82 and 86 depending on wind direction	155	no
86	Gasoline	24	22.4	48	64	49	NG	Tank 84 depending on wind direction	151	no
87	Gasoline	39	22.4	48	64	49	NG	Tank 86 depending on wind direction	140	no
90	Gasoline	39	22.4	48	64	49	NG	-	50	**Yes**
91	Slops (gasoline)	39	85.3	15	24	18	16	Tank 92 depending on wind direction	255	no
92	Slops (gasoline)	6	85.3	15	24	18	16	Tank 91 depending on wind direction	255	no
103	Slops (gasoline)	6	76.0	18	27	21	19	Tank 105 depending on wind direction	50	no
105	Slops (gasoline)	8	76.0	18	27	21	19	Tank 103 depending on wind direction	50	no

Note: *, + all distances measured from tank centre. + distance to site boundary given to closest relevant land-use, NG = Not Generated

TABLE 7.3 TANK OVERFILL CASCADE VAPOUR DISPERSION

Tank No.	Contents	Distance (m) to LFL*		Distance (m) to Boundary+	Off-Site Impact?
		D5 Case	F2 Case		
36	Gasoline	223	469	56	**Yes**
37	Gasoline	223	469	56	**Yes**
38	Gasoline	223	469	56	**Yes**
39	Gasoline	223	469	88	**Yes**

Tank No.	Contents	Distance (m) to LFL*		Distance (m) to Boundary+	Off-Site Impact?
		D5 Case	F2 Case		
50	Gasoline	228	571	96	**Yes**
53	Gasoline	228	571	47	**Yes**
84	Gasoline	231	487	155	**Yes**
86	Gasoline	233	589	151	**Yes**
87	Gasoline	233	589	140	**Yes**
90	Gasoline	233	589	50	**Yes**

Note: *, + all distances measured from tank centre. + distance to site boundary given to closest relevant land-use, NG = Not Generated

TABLE 7.4 BUND AND PUMP PIT FIRES

Area, Tank Bund or Pump Pit	Product	Equivalent Pool Diameter (m)	Surface Emissive Power (kW/m^2)	Distances (m) Note				Escalation Potential	Distance (m) to Closest Boundary	Off-Site Impact?
				Flame	4.7 kW/m^2	14 kW/m^2	23 kW/m^2			
Gate 1 Warehouse Package Store	Kerosene	49	20.9	55	74	55	NG	Low. No hydrocarbon inventories within 55 m.	75	No
Ethanol Tanker Unloading Bay	Ethanol	12	99.2	33	55	41	37	Low. No hydrocarbon inventories within 37 m.	> 100	No
Road Tanker Loading Gantry	Gasoline	37	22.9	46	62	48	NG	Low. No hydrocarbon inventories within 46 m. Tanker assumed to fail and add to pool fire.	> 100	no
Vapour Recovery Unit	Gasoline-type	12	56.1	23	35	27	24	Low. No hydrocarbon inventories within 24 m.	> 100	no
Tank Farm B	Gasoline	127	20	102	149	105	NG	Escalation to tanks in Tank Farm B1 possible depending on wind direction.	110	**Yes**
Tank Farm B1	Jet fuel	137	20	104	158	109	NG	Escalation to tanks in Tank Farm B possible depending on wind direction.	110	**Yes**
B/B1 pump pit WEST	Gasoline	17	42.8	28	42	32	29	Low. No hydrocarbon inventories within 29 m.	110	no
Tank Farm B2	Diesel	-	-	-	-	-	-	-	-	-
B/B1 pump pit EAST	Diesel	-	-	-	-	-	-	-	-	-
Tank Farm E1	Gasoline	89	20	80	114	82	NG	Escalation to tanks in Tank Farm E2 possible depending on wind direction.	130	no

Area, Tank Bund or Pump Pit	Product	Equivalent Pool Diameter (m)	Surface Emissive Power (kW/m^2)	Distances (m) Note				Escalation Potential	Distance (m) to Closest Boundary	Off-Site Impact?
				Flame	4.7 kW/m^2	14 kW/m^2	23 kW/m^2			
Tank Farm E2	Gasoline	74	20	71	99	73	NG	Escalation to tanks in Tank Farm E1 possible depending on wind direction.	170	no
Pump House 2	Gasoline	16	43.5	28	41	31	29	Low. No hydrocarbon inventories within 29 m.	60	no
Tank Farm K	Gasoline	87	20	79	112	81	NG	Low. No hydrocarbon inventories within 79 m.	40	**Yes**
ULP delivery pump Pit	Gasoline	18	39.3	30	44	33	31	Low. No hydrocarbon inventories within 31 m.	> 100	no
Gasoline slops tankfarm	Gasoline	28	27.3	39	54	41	39	Low. No hydrocarbon inventories within 39 m.	280	no

Note: all distances measured from bund/pool centre. NG = Not Generated

TABLE 7.5 PIPE TRACK POOL FIRES

Pipe	Press. (bara)	Hole (mm)	Release Rate (kg/s)	Equivalent Pool Diameter (m)	Surface Emissive Power (kW/m^2)	Distances (m) Note				Escalation Potential	Distance (m) to Boundary	Off-Site Impact?
						Flame	4.7 kW/m^2	14 kW/m^2	23 kW/m^2			
Gasoline from Gore Bay	20.3	2.5	0.2	5	90.7	11	17	13	11	Low. No hydrocarbon inventories within 11 m.	50	no
		20	10	47	21.1	43	58	44	NG	Low. No hydrocarbon inventories within 43 m.		**Yes**
		50	63	124	20.1	78	111	80	NG	Low. No hydrocarbon inventories within 78 m.		**Yes**
		100	198 (full-bore)	190	20.1	115	170	119	NG	Low. No hydrocarbon inventories within 115 m.		**Yes**
Gasoline from Pump House 2 to Silverwater Export Line	1.5	2.5	0.03	2	123	3	9	6	5	Low. No hydrocarbon inventories within 5 m.	50	no
		20	1.6	14	50.5	18	49	34	17	Low. No hydrocarbon inventories within 17 m.		no
		50	10	36	23.3	38	102	71	23	Low. No hydrocarbon inventories within 23 m.		**Yes**
		100	40	75	20.1	66	178	123	NG	Low. No hydrocarbon inventories within 66 m.		**Yes**

Pipe	Press. (bara)	Hole (mm)	Release Rate (kg/s)	Equivalent Pool Diameter (m)	Surface Emissive Power (kW/m^2)	Distances (m) Note				Escalation Potential	Distance (m) to Boundary	Off-Site Impact?
						Flame	4.7 kW/m^2	14 kW/m^2	23 kW/m^2			
Gasoline from Pump House 2 to Tank 90	1.5	2.5	0.03	2	128	3	9	6	5	Low. No hydrocarbon inventories within 5 m.	50	no
		20	1.6	14	66.8	18	49	34	21	Low. No hydrocarbon inventories within 21 m.		no
		50	10	36	30.1	38	103	70	36	Low. No hydrocarbon inventories within 36 m.		**Yes**
		100	40	75	20.7	66	179	122	NG	Low. No hydrocarbon inventories within 66 m.		**Yes**

Note: all distances measured from pool centre

TABLE 7.6 PIPE TRACK VAPOUR DISPERSION

Pipe	Pressure (bara)	Weight Flow (kg/s)	Hole (mm)	Release Rate (kg/s)	Distance (m) to LFL Note		Distance (m) to Boundary	Off-Site Impact?
					D5 Case	F2 Case		
Gasoline from Gore Bay	20.3	198	2.5	0.2	5	11	50	no
			20	10	46	120		**Yes**
			50	63	136	343		**Yes**
			100	198	240	531		**Yes**
Gasoline from Pump House 2 to Silverwater Export Line	1.5	91	2.5	0.03	2	4	50	no
			20	1.6	20	42		no
			50	10	57	120		**Yes**
			100	40	119	280		**Yes**
Gasoline from Pump House 2 to Tank 90	1.5	161	2.5	0.03	2	4	50	no
			20	1.6	20	42		no
			50	10	57	120		**Yes**
			100	40	119	280		**Yes**

Note: all distances measured from leak source

TABLE 7.7 LPG SPHERES, GANTRY & PUMPS JETFIRES

Vessel/ Area	Volume (m^3)	Press. (bara)	Scenario	Hole (mm)	Release Rate (kg/s)	Surface Emissive Power (kW/m^2)	Distances (m)[Note]				Escalation Potential	Distance (m) to Boundary	Off-Site Impact?
							Jet Flame	4.7 kW/m^2	14 kW/m^2	23 kW/m^2			
V-137	663	4.5	Fitting	1.4	0.02	173.5	3	4	3	3	Low. No hydrocarbon inventories within 3 m.	30	**Yes**
			Flange	2.5	0.06	178.4	4	7	6	5	Low. No hydrocarbon inventories within 5 m.		
			Pipe	100	25.5	174.2	46	107	80	68	Adjacent vessel (covered in BLEVE)		
			Overfill	rate	25.5	174.2	46	107	80	68	Adjacent vessel (covered in BLEVE)		
V-140	948	4.0	Fitting	1.4	0.02	173.5	3	4	3	3	Low. No hydrocarbon inventories within 3 m.	30	**Yes**
			Flange	2.5	0.06	178.4	4	7	6	5	Low. No hydrocarbon inventories within 5 m.		
			Pipe	100	25.5	174.2	46	107	80	68	Adjacent vessel (covered in BLEVE)		
			Overfill	rate	25.5	174.2	46	107	80	68	Adjacent vessel (covered in BLEVE)		
LPG Tanker Unloading Gantry	50	23.9	Coupling failure	1.4	0.05	173.7	4	6	5	4	Low. No hydrocarbon inventories within 4 m.	95	**Yes**
			Flange	2.5	0.15	177.8	6	9	8	7	Low. No hydrocarbon inventories within 7 m.		
			Hose failure	10	2.4	184.3	18	31	25	22	Road tanker (covered in BLEVE)		
			Pipework Rupture	20	9.8	185.7	31	57	45	40	Road tanker (covered in BLEVE)		
P5012	-	7	Seal leak	3	0.13	177.5	7	13	10	9	Low. No hydrocarbon inventories within 9 m.	20	no
P5015	-	7	Seal leak	3	0.13	177.5	7	13	10	9	Low. No hydrocarbon inventories within 9 m.	20	no
P5018	-	7	Seal leak	3	0.13	177.5	7	13	10	9	Low. No hydrocarbon inventories within 9 m.	20	no

Note: all distances measured from leak point

TABLE 7.8 LPG SPHERES, GANTRY & PUMPS VAPOUR DISPERSION

Vessel/ Area	Volume (m^3)	Pressure (bara)	Scenario	Hole (mm)	Release Rate (kg/s)	Distance (m) to LFL*		Distance (m) to Boundary+	Off-Site Impact?
						D5 Case	F2 Case		
V-137	663	4.5	Fitting	1.4	0.02	2	2	30	no
			Flange	2.5	0.06	3	7		no
			Pipe	100	25.5	110	325		**Yes**
			Overfill	Rate	25.5	110	325		**Yes**
V-140	948	4.0	Fitting	1.4	0.02	2	2	30	no
			Flange	2.5	0.06	3	7		no
			Pipe	100	25.5	110	325		**Yes**
			Overfill	Rate	25.5	110	325		**Yes**
LPG Tanker Unloading Gantry	50	23.9	Coupling failure	1.4	0.05	2	3	95	no
			Flange	2.5	0.15	4	8		no
			Hose failure	10	2.4	28	46		no
			Pipework Rupture	20	9.8	70	95		**Yes**
P5012	-	7	Seal leak	3	0.13	4	8	20	no
P5015	-	7	Seal leak	3	0.13	4	8	20	no
P5018	-	7	Seal leak	3	0.13	4	8	20	no

Note: * all distances measured from leak point; + distances measured from sphere/gantry/pump

TABLE 7.9 LPG BLEVE

Vessel/ Area	Volume (m^3)	Pressure (bara)	Liquid Mass (tonne) at Rupture	Fireball Duration (s)	Surface Emissive Power (kW/m^2)	BLEVE Fireball/ Radiation Distances (m)*			Distance (m) to Boundary+	Off-Site Impact?
						Fireball Diameter	Injury	Fatality		
V-137	663	4.5	300	25.9	384	265	1103	394	30	**Yes**
V-140	948	4.0	430	27.4	409	299	1314	479	30	**Yes**
LPG Tanker	50	23.9	22.6	15.3	275	112	314	83	95	**Yes**

Note: * distances measured from centre of fireball; + distances measured from sphere/gantry

7.5 Combustion Products

Toxic products of combustion, e.g. carbon oxides and soot, have the potential to affect (by respiratory irritation) those attending a fire emergency and possibly people off-site.

The products of combustion rising from a fire typically have a temperature in the range 800-1200ºC and a density a quarter that of air (Ref.17). Therefore, impact from toxic products of combustion will be significant only local to the fire, since the plume of combustion products would be buoyant and the combustion products will tend to rise and disperse with the prevailing weather (unless a temperature inversion exists).

7.6 Conclusions

The study found negligible potential for off-site escalation at adjacent industrial facilities (including at the boundary with the LyondellBasell Plant).

The following scenarios were carried forward for likelihood and risk analysis based on their potential for off-site impact:

- Tank roof fire: Tank 90
- Tank overfill cascade leading to flash fire/ vapour cloud explosion: all gasoline tanks
- Tank bund fires: Tank Farm B , Tank Farm B1 and Tank Farm K
- Pipe track pool fires
- Pipe track leaks (medium/ large leaks) leading to flash fire/ vapour cloud explosion
- LPG fires at the storage spheres and at the tanker unloading gantry
- LPG leaks (large only) at the storage spheres and at the tanker loading gantry leading to flash fire/ vapour cloud explosion
- LPG BLEVE at the storage spheres and at the tanker loading gantry

8.0 LIKELIHOOD ASSESSMENT

8.1 Introduction

The likelihood analysis is used in conjunction with the consequence analysis to determine the risk of an event. The likelihood analysis is a method for predicting the occurrence of future events based on past data. In terms of the QRA, events may occur due to either an equipment failure, for example a leak from a flange, or following a process control failure, e.g. a tank overfill.

A leak frequency data set was developed for equipment failure scenarios and was combined with a parts count.

The frequency of tank overfill was modelled using the residual risk values evaluated in the site's Model Bow Tie Layer of Protection Analysis (Ref.18), which was updated to reflect the proposed Terminal operating conditions.

The subsequent sections summarise the frequency analysis, which is detailed in APPENDIX C and summarised in Table 8.3.

8.2 Ignition Model

8.2.1 Immediate Ignition

The immediate ignition probabilities that will be used for the study were derived from data provided in Cox, Lees and Ang (Ref.19). A summary is provided in Table 8.3 for a range of leak rates.

TABLE 8.1 SUMMARY OF IMMEDIATE IGNITION PROBABILITIES

Release Rate (kg/s)	Immediate Ignition Probability	
	Vapour / Mixed Releases	Liquid Releases
<1 kg/s	0.0096	0.0096
1 - 50 kg/s	0.0616	0.0264
>50 kg/s	0.21	0.056

8.2.2 Delayed Ignition

Releases which are not immediately ignited (and hence result in vapour cloud dispersion) may be ignited by ignition sources at the facility, including pilot lights, flare stacks, non-intrinsically-safe electrical equipment, vehicles and mechanical sparks.

In Shepherd, all ignition sources, both on-site and off-site, which can result in delayed ignition of a leak, are added as point, line or area objects with appropriate ignition probabilities or, the user can specify an ignition probability per unit area of plant. The following delayed ignition probability densities are recommended by the UK HSE:

Ignition Source Densities (per hectare)	Industrial Areas	Urban Areas	Rural Areas
Day	0.25	0.20	9.9×10^{-3}
Night	0.17	0.13	6.5×10^{-3}

Noting that the area around the site is a uniform distribution of both urban and industrial development, an average value of 0.2 x 10^{-4} per m^2 was carried forward for the off-site ignition density.

In consultation with the Shepherd (QRA model) software developer, a value of 0.35 x 10^{-4} per m^2 was recommended for on-site ignition density, for consistency with the Cox, Lees and Ang immediate ignition probabilities described above.

8.3 Equipment Failure & Leak Frequency Data

The Shepherd software package used to carry out the QRA differentiates between LPG and non-LPG equipment. LPG equipment is defined as equipment that handles butane and/ or propane mixtures that are liquefied by pressure.

The frequency data are detailed in APPENDIX C and summarised in Table 8.3.

8.4 Parts Count

The equipment parts count is summarised in Table 8.2. The parts count was based on a standard design for equipment conforming to Shell Design & Engineering Practices (DEPs) and confirmed by cross-check with actual parts (in the field or off drawings, where possible). The parts count is the average count across all equipment items of a given type.

TABLE 8.2 PARTS COUNT

Equipment Type	Average Leak Source Count per Equipment Item		
	Connections (> 25 mm)	Flange / Valve Equivalents	Fittings (< 25 mm)
Pumps	1	18	9
Pipe	3	20	3
LPG Bullet / Sphere	-	50	10

8.5 BLEVE Frequency

The frequency of BLEVE is determined by the QRA model based on a summation of the frequency of fire scenarios that have the potential to impinge on the storage vessels/ road tankers for the period of time determined to result in weakening of the shell (see Table D.3 in APPENDIX D).

8.6 Summary of Frequencies Used in QRA

The frequency data are detailed in APPENDIX C and summarised in Table 8.3.

TABLE 8.3 SUMMARY OF FAILURE & EVENT FREQUENCIES USED IN QRA

Equipment Item	Equivalent Leak Size Diameter (mm)	Frequency (per item-year)
LPG Equipment Leaks		
Flanges and equivalent valves	2.5	5.6×10^{-6}
Instrument fittings and connections	1.4	5.6×10^{-6}
Pipe (including pipelines) x (L/D)	100	4.9×10^{-7}
Hose / hard-arm (hose failure)	10	6.65×10^{-6} per operation
Hose / hard-arm (coupling failure) (Excess flow valve limited)	1.4	5.2×10^{-6} per operation
Catastrophic vessel failure	-	2.4×10^{-8}
Overfill of Aboveground LPG Storage Vessels	-	7.6×10^{-5}
Non-LPG Equipment Leaks		
Flanges and equivalent valves	2.5	2.2×10^{-4}
Instrument fitting (< 1 inch diameter)	20	1×10^{-4}
Connection (> 1 inch diameter)	50	1×10^{-5}
Pipe rupture	≥ 300	7×10^{-8} per m
	< 300	2×10^{-7} per m
Pump Seal	10	3×10^{-3} (single seal)
Pump Casing Failure	Full Bore	3×10^{-5}
Atmospheric Storage Tanks		
Full-surface tank roof fire	-	1.2×10^{-4} per tank
Tank Farm E1 tank overfill (summed across all tanks in bund)	-	8.5×10^{-5}
Tank Farm E2 tank overfill (summed across all tanks in bund)	-	8.9×10^{-5}
Tank Farm K tank overfill (summed across all tanks in bund)	-	2.6×10^{-5}
Tank Farm B tank overfill (summed across all tanks in bund)	-	4.2×10^{-5}
Tank Farm B1 tank overfill (summed across all tanks in bund)	-	6.6×10^{-5}

9.0 RISK ASSESSMENT

9.1 Introduction

The risk analysis brings together the physical consequence model, effects models, leak frequency and parts count. The modelling also includes site specific issues such as equipment layout and prevailing weather conditions.

The PHA process requires an assessment of the offsite risk against set criteria. This section presents the comparison of the risk, developed from the consequence and frequency analyses, against the HIPAP 4 Offsite Risk Tolerability Criteria (Ref.5).

9.2 Modelling Approach

The QRA methodology adopted in this study is consistent with the guidance provided in HIPAP 6 (Ref.4).

The risk profile for the site was produced by the proprietary package Shell Shepherd, which uses the concept of Process Blocks to define hazardous processes/ streams (usually isolatable sections of plant) and the associated loss of containment scenarios.

9.3 Assessment Criteria

The risk guidelines provided in the DPI publication Risk Criteria for Land Use Safety Planning (Ref.5) are outlined below for fatality, injury, accident propagation and damage to the biophysical environment.

9.3.1 Fatality, Injury & Accident Propagation Risk Criteria

TABLE 9.1: NSW INDIVIDUAL FATALITY RISK CRITERIA

Limit (per year)	Land-Use
0.5×10^{-6}	Hospitals, child-care facilities and old age housing developments
1×10^{-6}	Residential developments and places of continuous occupancy such as hotels and tourist resorts
5×10^{-6}	Commercial developments, including offices, retail centres, warehouses with showrooms, restaurants and entertainment centres
10×10^{-6}	Sporting complexes and active open space areas
50×10^{-6}	Industrial – must not be exceeded any boundary adjacent to another industrial facility

TABLE 9.2 INDIVIDUAL INJURY RISK CRITERIA

Limit (per year)	Land-Use
50×10^{-6}	Residential areas – 4.7 kW/m^2 heat flux radiation
50×10^{-6}	Residential areas – 7 kPa explosion overpressure
10×10^{-6}	Residential areas – injurious toxic concentrations *NOTE: risk contour not evaluated – no toxics handled at Terminal*
50×10^{-6}	Residential areas – toxic concentrations causing irritation *NOTE: risk contour not evaluated – no toxics handled at Terminal*

TABLE 9.3 ACCIDENT PROPAGATION/ ESCALATION RISK CRITERIA

Limit (per year)	Land-Use
$50x10^{-6}$	Potentially hazardous installations – 23 kW/m^2 heat flux radiation (flame impingement)
$50x10^{-6}$	Potentially hazardous installations – 14 kPa explosion overpressure

9.3.2 Criteria for Risk to the Biophysical Environment

The risk tolerability criteria suggested by DPI for sensitive environmental areas relate to the potential effects of an accidental emission on the long-term viability of the ecosystem or any species within it. The criteria are expressed as follows:

- Industrial developments should not be sited in proximity to sensitive natural environmental areas where the effects of the more likely accident emissions may threaten the long-term viability of the ecosystem or any species within it.
- Industrial developments should not be sited in proximity to sensitive natural environmental areas where the likelihood of impacts that may threaten the long-term viability of the ecosystem, or any species within it, is not substantially lower than the background level of threat to the ecosystem.

An accidental emission from the proposed site is unlikely to affect the long-term viability of any ecosystem since the area surrounding the proposed Terminal is industrial in nature and does not include any sensitive environmental areas.

9.3.3 Societal Risk

The Department of Planning suggests that judgements on societal risk be made on the basis of a qualitative approach rather than on specifically set numerical criteria. Societal risk estimation is warranted only where significant and potentially vulnerable populations exist beyond the boundary of the proposed development. This is not the case for the proposed terminal, since the surrounding area (within the worst-case consequence distance generated by the Terminal hazards) includes only industrial, and no residential, land-uses. Societal risk was therefore not evaluated in this study.

9.4 Risk Assessment

9.4.1 Refinery Mode of Operations

Individual fatality risk contours for the facility when operating as a refinery were produced for the Clyde Refinery Safety Report, Ref. 20, and these are reproduced in Figure 9.1 as a baseline from which all proposed changes can be reviewed. It can be seen that the red $5x10^{-5}$/ year contour extends marginally offsite and all other contours cover large portions of the surrounding area. This is typical of an operating refinery storing substances that on release have large consequence distances such as Hydrofluoric Acid.

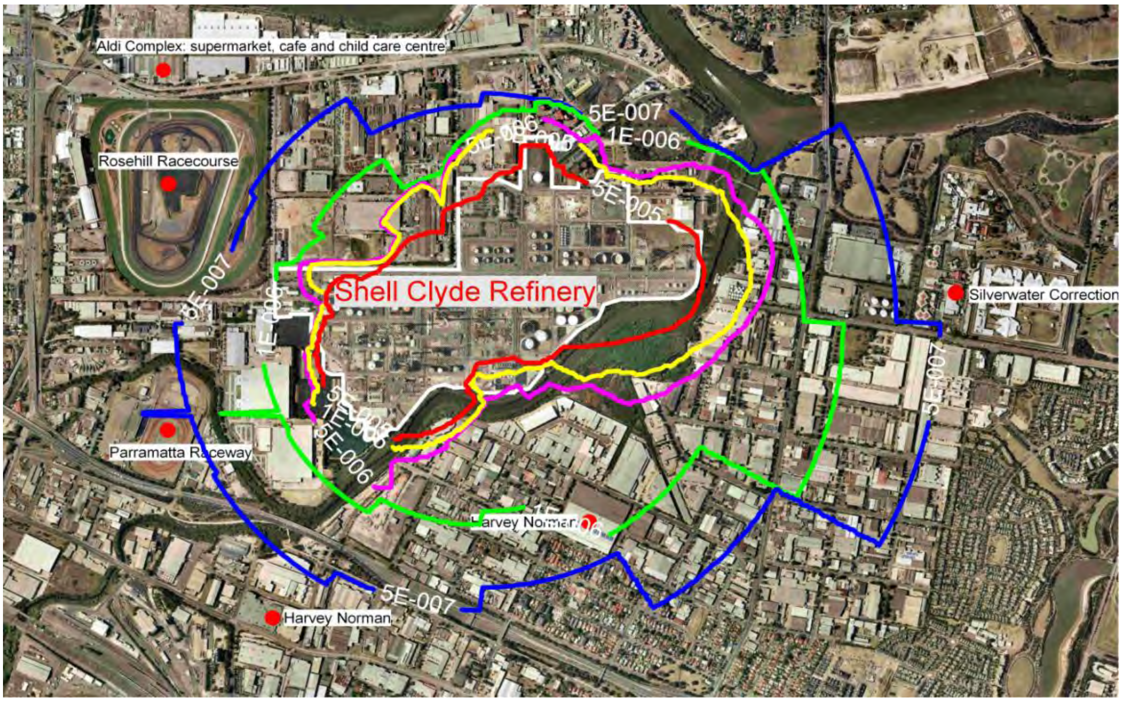

Figure 9.1 Individual Fatality Risk Contours (Safety Report)

9.4.2 Proposed Terminal Mode of Operations

Individual fatality risk contours for the Terminal are shown in Figure 9.2 and a summary of the risk assessment findings against the HIPAP 4 fatality risk criteria is provided in Table 9.4. When compared against the Refinery contours, the picture is very different with much smaller contour extents, typical of a fuel storage and distribution site.

Heat radiation injury risk contours are shown in Figure 9.3. The explosion overpressure injury risk was found to be less than $50x10^{-6}$ per year and thus the contour is not shown. A summary of the findings against the injury risk criteria is provided in Table 9.5.

Accident propagation risk contours are shown in Figure 9.4 for heat radiation. The explosion overpressure escalation risk was found to be less than $50x10^{-6}$ per year and thus the contour is not shown. A summary of the risk assessment findings against the guidance only criteria is provided in Table 9.6.

TABLE 9.4 INDIVIDUAL FATALITY RISK ASSESSMENT

Land-Use	Outcome
Hospitals, child-care facilities and old age housing developments	The risk at the nearest existing hospital, child-care facility and old age housing development is less than 0.5×10^{-6} /year **(blue)**
Residential developments and places of continuous occupancy such as hotels and tourist resorts	The risk at the nearest existing residential area is less than 1×10^{-6}/year **(green)**
Commercial developments	The risk at the nearest existing commercial development is less than 5×10^{-6} /year **(purple)**
Sporting complexes and active open space areas	The risk at the nearest existing sporting complex is less than 10×10^{-6} /year **(yellow)**
Industrial – must not be exceeded any boundary adjacent to another industrial facility	The risk at the boundary with adjacent industrial land use is less than 50×10^{-6} /year **(red)**

Figure 9.2 Individual Fatality Risk Contours

TABLE 9.5 INDIVIDUAL INJURY RISK ASSESSMENT

Land-Use	Contour Level (colour)
Residential areas – 4.7 kW/m^2 heat flux radiation	Heat flux radiation levels of 4.7 kW/m^2 do not impact residential development at frequencies of more than 50 chances in a million per year **(red)**
Residential areas – 7 kPa explosion overpressure	Explosion overpressure levels of 7 kPa or greater would not be expected to occur at frequencies of more than 50 chances in a million per year (note: no contour plot shown)

Figure 9.3 Heat Radiation Injury Risk Contours

TABLE 9.6 ACCIDENT PROPAGATION/ ESACALATION RISK ASSESSMENT

Land-Use	Contour Level (colour)
Potentially hazardous installations – 23 kW/m^2 heat flux radiation (flame impingement)	Flame or incident heat flux radiation levels of 23 kW/m^2 do not exceed a risk of 50 per million per year at neighbouring potentially hazardous installations. **(red)**
Potentially hazardous installations – 14 kPa explosion overpressure	Incident overpressures of 14 kPa or greater would not be expected to occur at frequencies of more than 50 chances in a million per year (note: no contour plot shown)

Figure 9.4 Heat Radiation Accident Propagation (Escalation) Risk Contours

9.5 Risk to the Biophysical Environment

An assessment of the potential for long-term effects, due to an accidental emission of hydrocarbon from the site, on the viability of ecosystems in the area surrounding the Terminal is undertaken in the Environmental Impact Statement (EIS – Ref.21).

The EIS found that:

- the effects of the more likely accident emissions do not threaten the long-term viability of the local ecosystems and the species within it; and
- the likelihood of the impacts that potentially threaten the long-term viability of the ecosystem, or any species within it, is lower than the background level of threat to the ecosystem.

9.6 Conclusions

The main contributors to off-site fatality risk were found to be BLEVE of the LPG spheres and flash fire (vapour cloud explosion) following overfill of gasoline storage tanks.

The risk assessment indicates that the site complies with all relevant HIPAP criteria; and is, therefore, considered 'potentially hazardous', rather than 'hazardous', and ‘potentially offensive’, rather than ‘offensive’, in the context of SEPP33.

10.0 RISK CONTROLS

10.1 Introduction

Demonstrating that the risks meet the acceptance criteria set is only one element of demonstrating that risks are being suitably managed.

In order to demonstrate that the risk is being controlled, the Multi-Level Risk Assessment requires a discussion of the technical controls, risk reduction measures and management measures in place.

Demonstration of control measure adequacy is detailed in the sites compliance with the Shell Hazards & Effects Management Process (HEMP), which is covered in detail in the site's NSW MHF Safety Report. This section summarises the risk controls under two broad headings:

- Safety in Design
- Safety in Operation

The controls were identified from the site's hazard register, bow tie hazard analyses and layers of protection analyses.

10.2 Safety in Design

10.2.1 Codes and Standards

All tanks are/ will be constructed to recognised Australian and International Standards. These will be supported by the Shell Design and Engineering Practices (DEPs). Project documentation includes a complete list of standards.

An example of applicable codes and standards are AS1940 for bund walls and AS1170.2 and AS1170.4 for wind and earthquake design loadings.

10.2.2 Risk Management in Design

The design will be subject to the Shell risk management process. Risk management activities that directly relate to the NSW Seven Stage Planning Process are shown in Table 10.1.

TABLE 10.1 RISK MANAGEMENT ACTIVITIES

Activity	Status
Hazard Identification	Complete and reported in this document
Preliminary Hazard Analysis	Complete and reported in this document
Hazard & Operability Study	Planned activity
Fire Safety Study	Currently underway
Final Hazard Analysis	Planned activity
Emergency Plan Review	Currently underway
Construction Safety Study	Planned activity
Safety Management System Update	Planned activity

10.2.3 Safety Systems

A summary of the proposed safety systems is given in Table 10.2. Further details are available in project documentation.

TABLE 10.2 PROPOSED SAFETY SYSTEMS

System	Comment
Process Control	The process control system (tank level gauging) integrated within the Distributed Control System will be further upgraded for the new terminal operations.
Process Shutdown Systems	Existing pump interlocks will be retained and new tank high level trips will be provided as required to demonstrate ALARP risk.
Bund Walls and Drains	The existing bunds and drains will be retained.
Fire Water	The existing firewater main, monitors and hydrants will be modified for the new terminal operations. Further details are provided in the Fire Safety Study.
Fire protection system	The existing tank fire system will be revised and remotely operable for the new terminal operations.
Hazardous Area Classification	Ignition sources will be controlled by the application of suitable hazardous area classification standards

10.3 Safety in Operation

The Clyde Refinery and Gore Bay Management System covers the following activities (relevant to Terminal operations):

- training of operators on new plant;
- operating procedures; and
- spares and maintenance of new equipment.

As the design progresses the relevant sections of the management system will be triggered.

In addition, the management system will be updated as the site moves from refining to product import, storage and distribution.

10.4 Proposed Automation & Safeguarding Upgrades

The following safeguards and automation upgrades are proposed:

- Yokogawa Prosafe SGS will be installed to replace the functionality of the existing relay logic.
- Permissives (interlocks) will be improved to prevent the incorrect valves being opened.
- Motorised valves will be installed inside tank bunds to allow quicker acting valves and remote operation.
- The reliability of telemetry between Clyde/Gore Bay will be improved.
- The Independent High Level Alarm and tank gauging systems will be improved.
- Pump trip systems will be improved.

- The site fire system and dump valve logic will be improved.
- Fully integrated and remotely operable foam system will be installed.
- Non-safeguarding controls will also be upgraded.

APPENDIX A. HAZARD IDENTIFICATION WORD DIAGRAM

TABLE A.1: HAZARD IDENTIFICATION WORD DIAGRAM

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
Atmospheric Storage Tanks and Bunds				
Tank and associated pipework integrity failure	Internal scoring by floating roof	Tanks inspected internally on a routine basis by Inspection. Routine operator surveillance.	Leak into bund and large pool/ bund fire, if ignited. Note: diesel requires strong ignition source (e.g. adjacent tank/ bund fire) to result in fire. Diesel tanks considered to be tanks-on-fire only if there is potential for escalation from gasoline or jet fuel tank/ bund fire.	**Detection** Fire detection by operator or CCTV. **Protection** Foam and cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	Corrosion	Routine operator surveillance. 10 yearly inspection as per AS1940. Tank inspected internally as per API653 on a routine basis by Inspection. Corrosion allowance added to tank design. Tank outer walls painted. Online ultrasonic thickness testing undertaken routinely (as part of standard maintenance procedure), epoxy coating of the tank floor and half way up the first strake on the tank wall.		
	Underfloor corrosion	Tank inspected internally on a routine basis by Inspection. Magnaflux testing undertaken. API653 and AS1940 - Inspections and operator surveillance.		
	Terrorist attack	Clyde Refinery Security plan.		
	Flooding	Clyde Refinery Emergency Response Plan.		
	Vacuum pulled on tank during lowering by Pressure/Vacuum valve failure	PV valves installed and overhauled on an ad-hoc basis by Inspection. Floating roof tanks are fitted with leg vents (vacuum breakers) which are serviced during tank turnarounds. Operator surveillance.		
	Natural disaster (lightning strike)	Tank earth straps fitted. Floating roofs earthed through shell. Straps inspected on a routine basis by IE - 2 yearly. Operator surveillance.		
	Crane impact	No crane lifts post commissioning.		
	Excessive filling rate	Flowrates come through to DCS, though not alarmed. OMS tracks but does not alarm on high tank rates. Recommissioning procedures.		
	Incorrect material lined up to tank	Operational vigilance. By design, difficult to do.		
	Hot work on tank	Permit to Work system. Hot work undertaken under constant gas monitoring. Tanks taken out of service and mechanically ventilated prior to hot work. Individual procedure issued for each tank outage. Gas monitoring of adjacent tanks.		
	Vapour space ignition - geo dome option	Ignition sources excluded by PTW. Anti-flash gauze fitted to PV valves (flame arrestor). Earthing of the roof and operator surveillance.		

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
Tank overfill	Level gauging failure	Level gauges calibrated on a routine basis by hand dip. Independent high high level alarms fitted on tank, routinely tested. Independent high level alarms on floating roof tank tested operationally. OMS system tracks tank movements and high level alarms. Panel operator monitor tank levels 4 hourly.		
	Passing valves	OMS tracks tank levels, some tanks have independent high level alarms fitted. OMS generates alarms when tank level moves when it shouldn't. Panel operator monitors tank levels 4 hourly.		
	Isolation failure, inadvertent operation/ gravitation	Operator training and vigilance. OMS tracks tank levels, independent high level alarms fitted. Panel operator monitor tank levels 4 hourly. Non-return valve in line.		
	Incorrect material lined up to tank (low flash added to high flash tank)	Operational vigilance. By design, difficult to do.		
Floating tank roof sink, exposing liquid.	Roof drain failure (floating roof option)	Routine operator surveillance to ensure water is not on roof and no hydrocarbon is passing into tank farm. Tanks are internally inspected on a routine basis by Inspection.	Full-surface tank roof fire, if ignited. Note: diesel requires strong ignition source (e.g. adjacent tank/ bund fire) to result in fire. Diesel tanks considered to be tanks-on-fire only if there is potential for escalation from gasoline or jet fuel tank/ bund fire.	**Detection** Leak/ fire detection by operator or CCTV. Linear fire detection above rim seals for some tanks. **Protection** Foam header for tank roof (rim seal pourers) foam application by fire brigade tender. Foam and cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	Jammed roof	Routine operator surveillance to ensure roof is not jammed (only applicable to roofs that can be seen - cannot readily see internal pans on internal floating roof tanks) Fill rate kept inside limits. Routine maintenance to gauge pole rollers.		
	Corrosion (holed roof)	10 yearly AS1940 inspection and testing of roof surface and pontoons. Operator surveillance.		
	Gas build-up under roof.	PV valves installed and overhauled on an ad-hoc basis by Inspection - fixed valves. Floating roof tanks are fitted with leg vents (vacuum breakers) which are serviced during tank turnarounds.		
	Leaking pontoon	10 yearly AS1940 inspection and testing of roof surface and pontoons. Operator surveillance.		

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
Tank fire	Roof drain failure	Routine operator surveillance to ensure water is not on roof and no hydrocarbon is passing into tank farm. Tanks are internally inspected on a routine basis by Inspection.	Rim seal fire or full-surface tank roof fire Note: diesel requires strong ignition source (e.g. adjacent tank/ bund fire) to result in fire. Diesel tanks considered to be tanks-on-fire only if there is potential for escalation from gasoline or jet fuel tank/ bund fire.	**Detection** Leak/ fire detection by operator or CCTV. Linear fire detection above rim seals for some tanks **Protection** Foam header for tank roof (rim seal pourers) foam application by fire brigade tender. Foam and cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	SITA stack emissions may contain hot soot	Clyde Refinery security plan.		
	Terrorist attack	Clyde Refinery security plan.		
	Lightning strike	Tank earth straps fitted. Floating roofs earthed through shell. Straps inspected on a routine basis by IE. Routine operator surveillance of earth straps.		
	Hot work on tank	Permit to Work system. Hot work undertaken under constant gas monitoring. Tanks taken out of service and mechanically ventilated prior to hot work. Individual procedure issued for each tank outage. Gas monitoring of adjacent tanks.		
	Static electricity	Tank earth straps fitted. Floating roofs earthed through shell. Straps inspected on a routine basis by IE. Routine operator surveillance of earth straps. Filling rate on return-to-service is limited by procedure.		
Hydrocarbon release to bund	Leaking roof drains and product release via drain	Operator surveillance. Drains isolated if found to be leaking and floating roofs monitored. If tank is static OMS will alarm if level moves significantly.	Leak into bund and large pool/ bund fire, if ignited. Note: diesel requires strong ignition source (e.g. adjacent tank/ bund fire) to result in fire. Diesel tanks considered to be tanks-on-fire only if there is potential for escalation from gasoline or jet fuel tank/ bund fire.	**Protection** Leak/ fire detection by operator or CCTV. **Protection** Foam and cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	Leaking or overfilled tank	Tank inspected internally on a routine basis by Inspection. Corrosion allowance added to tank design. Tank outer walls painted. Online ultrasonic thickness testing undertaken routinely, epoxy coating of the tank floor and half way up the first strake on the tank wall. Routine operator surveillance to detect minor leakage.		
	Piping and gasket failures	Operator surveillance. Routine inspection. Only steel piping used.		
	Pump seal leaks	Operator surveillance. Routine inspection: temperature and vibration monitoring/ survey		
	Isolation and drain valve failure	Only steel piping used. Operator surveillance. Routine inspection.		
	Sample points and water drain left inadvertently open	Policy to install double valves. Operator training and surveillance. Water draining continually monitored.		
	Pipe failure through bund wall	Operator surveillance. Ad-hoc inspection.		

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
Butane Storage Spheres				
Sphere and associated pipework integrity failure	Corrosion	Routine operator surveillance. 10 yearly inspection as per AS1596. Tank inspected internally as per API653 on a routine basis by Inspection. Corrosion allowance added to tank design. Sphere outer walls painted. Online ultrasonic thickness testing undertaken routinely (as part of standard maintenance procedure).	Pool fire if ignited, potential BLEVE. Flash fire if delayed ignition.	**Detection** Gas detection by fixed gas detectors. Melt tubes. CCTV. **Protection** V-140: deluge over full surface V-137: deluge up to approx.. 8 m height Remote operated monitor on south side of V-137 Cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	Terrorist attack	Clyde Refinery Security plan.		
	Flooding	Clyde Refinery Emergency Response Plan.		
	Natural disaster (lightning strike)	Tank earth straps fitted. Straps inspected on a routine basis by IE - 2 yearly. Operator surveillance.		
	Crane impact	No crane lifts post commissioning.		
	Excessive filling rate	Flowrates come through to DCS, though not alarmed. OMS tracks but does not alarm on high tank rates. Recommissioning procedures.		
	Hot work	Permit to Work system. Hot work undertaken under constant gas monitoring. Spheres taken out of service and mechanically ventilated prior to hot work. Individual procedure issued for each tank outage. Gas monitoring of adjacent tanks.		
Overfill/ Overpressure	Level gauging failure	Level gauges calibrated on a routine basis by hand dip. Independent high level alarms, routinely tested. Safety instrumented (SIL-1) high level trips, routinely tested. Independent high pressure alarms. Safety instrumented (SIL-1) high pressure trips, routinely tested. OMS system tracks movements and high level alarms. Panel operator monitor tank levels 4 hourly.		
	Passing valves	OMS tracks tank levels, some tanks have independent high level alarms fitted. OMS generates alarms when tank level moves when it shouldn't. Panel operator monitors tank levels 4 hourly.		
	Isolation failure, inadvertent operation/ gravitation	Operator training and vigilance. OMS tracks tank levels, independent high level alarms fitted. Panel operator monitor tank levels 4 hourly. Non-return valve in line.		

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
LPG Tanker Unloading Gantry				
Tanker Overfill/ Overpressure	Level gauging failure	Scully overfill protection system. Vapour Recovery System & Sealed Tanker compartments.	Leak into gantry bund and large pool/ bund fire, if ignited. Potential BLEVE of laden road tanker.	**Detection** Closed drain. Melt tube fire detection **Protection** Automatic foam deluge. Foam and cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	Driver error: fails to stop filling or incorrect (full) compartment selected	Driver monitoring loading via tanker indication of compartment volume. Loading procedures and Omega Loading System does not allow load of more volume than is available. Vapour Recovery System & Sealed Tanker compartments. Scully overfill protection system.		
	Pressure relief vents on tanker fail to open. Flame Arrestor on vapour return line blocked.	Compartment PV Vents lift and relieve pressure. Designed to Industry Standards		
Gantry equipment/ pipework integrity failure	Tanker or gantry piping, hose and gasket failures Tanker compartment leak	Only steel piping and braided hoses used. Statutory inspection of hoses. Tanker integrity managed by contract. Design of the compartments and fittings, built to Standards. Tanker Inspections to confirm Safe Load Pass. Driver in attendance visually observing compartments and equipment.		
Tanker drive-away with loading hose/ arm attached	Driver error	Brake interlock – Drive-away bar lock		
		Breakaway on loading hose/ arm minimises leak size		
Pump House No. 2				
Loss of containment	Piping and gasket failures	Operator surveillance. Routine inspection. Only steel piping used.	Leak into bund and large pool/ bund fire, if ignited. Note: diesel requires strong ignition source.	**Detection** IR flammable gas detectors in pump house building. Melt tube fire detection in pump house building and various pumps. **Protection** Automatic firewater deluge Foam and cooling water application via hydrants and monitors (fixed/ portable). Emergency Response Plan.
	Pump seal leaks	Operator surveillance. Routine inspection: temperature and vibration monitoring/ survey		
	Isolation and drain valve failure	Only steel piping used. Operator surveillance. Routine inspection.		
	Sample points left inadvertently open	Policy to install double valves. Operator training and surveillance.		
Road Tanker Loading Gantry and Underground Ethanol Storage Tank				
Tanker Overfill/ Overpressure	Level gauging failure	Scully overfill protection system. Vapour Recovery System & Sealed Tanker compartments.	Leak into gantry bund and large pool/ bund fire, if ignited. Note: diesel requires strong ignition source.	**Detection** Closed drain. Melt tube fire detection **Protection** Automatic firewater deluge. Foam and cooling water application via hydrants and
	Driver error: fails to stop filling or incorrect (full) compartment selected	Driver monitoring loading via tanker indication of compartment volume. Loading procedures and Omega Loading System does not allow load of more volume than is available. Vapour Recovery System & Sealed Tanker compartments. Scully overfill protection system.		

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
	Pressure relief vents on tanker fail to open. Flame Arrestor on vapour return line blocked.	Compartment PV Vents lift and relieve pressure. Designed to Industry Standards		
Gantry equipment/ pipework integrity failure	Tanker or gantry piping, hose and gasket failures Tanker compartment leak	Only steel piping and braided hoses used. Statutory inspection of hoses. Tanker integrity managed by contract. Design of the compartments and fittings, built to Standards. Tanker Inspections to confirm Safe Load Pass. Driver in attendance visually observing compartments and equipment.		
Tanker drive-away with loading hose/ arm attached	Driver error	Brake interlock – Drive-away bar lock Breakaway on loading hose/ arm minimises leak size		
Overfill of underground ethanol storage tank	Driver error	Unload undertaken by gravitation and tank vent point higher than ethanol tanker (such that overfill is unlikely)	Leak and ethanol pool fire, if ignited	**Detection** Operator detection. **Protection** Firewater application via hydrants and monitors. Emergency Response Plan.
Gate 1 Dangerous Goods Warehouse				
Spill outside of warehouse during unloading	Damaged drums from unsecured load and load movement during transport Collision of forklift and truck (excess speed, poor visibility) Dropped load due to uneven ground, human error in forklift operation Restriction on vehicle turning circle	Certified DG freighters Licensed drivers and vehicles Licensed forklift drivers (incl. driver training and procedures) Only flame proof forklifts used Regular inspection/ maintenance of forklifts Site speed restrictions (15 km/hr) Low traffic area - potential for collision is low	Leak and pool fire, if ignited	**Detection** Driver or warehouse operator detection **Protection** Extinguishers. Emergency Response Plan.
Spill inside of warehouse during unloading	Dropped load Leaking drum Package instability Packages ruptured by forklift tynes Drum corrosion due to ingress of water	Licensed forklift drivers (incl. driver training and procedures) Only flameproof forklifts will be used Regular inspection/ maintenance of forklifts Racking design to industry standard (fit for purpose + guarding against forklift impact) Lighting is flameproof, ignition control per Hazardous Zoning Fire protection equipment provided in accordance with AS1940 Store is bunded Store is roofed and cladded to minimise rain water ingress	Leak and pool fire, if ignited	**Detection** Driver or warehouse operator detection **Protection** Extinguishers Emergency Response Plan.
External fire	Hot-work during maintenance activities Faulty equipment (forklift and lighting) Smoking	Permit to work system (hot work) used on site Change of management procedure Dedicated smoking area Intrinsically safe equipment is certified Separation distance of store to protected works per AS1940 requirements		

Accident Event (Leak or Fire)	Causes	Prevention Measures	Consequences	Detection and Protection Measures
Vapour Recovery Unit				
Loss of containment from VRU	Overfill of underground condensate tank	Inspection by Jordan (vendor) Technician	Pool fire, if ignited	**Detection** High level alarm
	Equipment failure: flange, pipe, fitting, pump leaks	Regular inspection/ maintenance (including coating and conditioning monitoring) Inspection by Jordan (vendor) Technician Surge control	Pool fire, if ignited	**Detection** Visual/ operator rounds **Protection** Manual ESD system Ignition control Extinguishers Emergency Response Plan.
	Leaking of underground tanks	Regular inspection/ maintenance (including coating and conditioning monitoring)	Release to the environment	
Overheating of VRU adsorption bed	Incompatible solvent vapour in the VRU due to switch loading	Switch loading procedures Maintenance and inspection	Runaway isotherm and auto-ignition and fire within VRU	**Protection** Inert gas flooding (via ESD)

APPENDIX B. METEOROLOGICAL DATA

In the Shepherd QRA modelling software, all modelling is done for weather stability classes / categories D and F. Weather category D with a wind speed of 5 m/s (D5) is representative of the "typical" day weather category and F2 includes the calm, night weather category (for observation of gas dispersion over longer distances). Weather categories A and B are typical observed in hotter / tropical climates. Though not incorporated into Shepherd, they are deemed to be sufficiently well described by the D category (D will generally yield a more conservative assessment) and have thus not been explicitly included in the specification.

Shepherd calculations are based on a 12-sector wind rose; each sector representing one hour in the clock. Values of the wind speed and Pasquill class are specified by the user in each of the sectors of the wind rose for weather categories D and F, with a maximum of 3 wind speeds. In Shepherd the wind direction is defined as the being in a given 30° sector (blowing towards rather than from a sector). Note that weather data normally present the fraction the wind is blowing from a certain direction.

Data of wind speed and probability of it being in/from a certain direction is specified by the user and is based on meteorological data (wind rose), from a nearby weather station, obtained from the Bureau of Meteorology (BoM). The BoM does not have direct observations from Clyde. The nearest observations were taken from Olympic Park (about 4 km to the east of the refinery), where an automatic weather station (BoM site number 066195) provides observations of temperature, average wind speed and direction and standard deviation of wind direction.

The closest cloud observations were taken from an automatic ceilometer at Bankstown Airport (BoM site number 066137), about 10 km south of Clyde, which gives a cloud base measurement every 12 seconds. For times when the ceilometer was not available, manual cloud observations from Sydney Airport (site number 066037) were also obtained. All these data are stored in ADAM, the Bureau of Meteorology's national climate data archive, and go back to at least December 1998. However, a 30-degree error in wind direction from Olympic Park occurred some time between an inspection in May 1999 and one in June 2000. For this reason, the four-year period from January 2001 to December 2004 was chosen by the BoM for the data analysis.

As summary of the wind speed and weather stability data, in the format required by Shepherd (i.e., 12-sector wind direction toward sector and D and F stability classes), is provided below:

	Probability of Wind Direction Toward (degrees true)											
Stability and Wind Speed	**0**	**30**	**60**	**90**	**120**	**150**	**180**	**210**	**240**	**270**	**300**	**330**
D 5	0.028	0.032	0.057	0.062	0.037	0.033	0.031	0.031	0.050	0.048	0.060	0.040
D 9	0.006	0.001	0.014	0.018	0.006	0.006	0.001	0.006	0.010	0.007	0.031	0.024
F 5	0.004	0.005	0.018	0.021	0.008	0.008	0.004	0.008	0.004	0.002	0.006	0.004
F 1.5	0.011	0.018	0.045	0.062	0.026	0.014	0.019	0.021	0.020	0.013	0.013	0.008

APPENDIX C. FREQUENCY DATA

Overview

The generic equipment failure frequencies adopted for this study are based on published leak frequency data and reflect the data used by the site to compile their QRA (as shown in the Clyde Refinery NSW MHF Safety Report). The data has been consolidated from previous Shell safety studies and the more recently published UK HSE offshore hydrocarbon release program.

The development of the leak frequency takes into account the following factors:

- Credible leak scenarios.
- The format of input required by the models.
- Historical leak data analysed to match the leak scenarios and input requirements.

Each factor is discussed as a separate element in this appendix.

Shepherd differentiates between LPG and non-LPG equipment.

The leak frequency results and representative leak sizes are summarised in Table C.1. These have been adopted from the site's existing Safety Report and QRA.

Reliability Data Sources

The following data sources were reviewed in compiling the reliability data for this study:

J.L. Hawksley, Some Social, Technical and Economical Aspects of the Risks of Large Technical Plants, Paper to Chemrawn III, June 1984

HSE/SRD PD052/WP1, BLEVE Probability of a 100 te LPG Storage Vessel, Blything, K. W., Reeves, A. B., Warrington, UK, December 1987, HMSO.

Technica C3045, Three Site Risk Study for Shell Malaysia, February 1992

ASSESS 6, A computer programme to determine the risks from LPG distribution sites, Release version 6.1, SLEU, 1998

COVO1982, Risk Assessment of 6 potentially hazardous industrial objects in the Rijmond area - a pilot study

SINTEF: OREDA-97 – Offshore Reliability Data Handbook, 3rd Edition, Hovik, Norway, 1997

Institute of Electrical and Electronics Engineers: Guide to the Collection and Presentation of Electrical, Electronic, Sensing Component, and Mechanical Equipment Reliability Data for Nuclear-Power Generating Stations, USA.

The Oil Industry International Exploration and Production Forum (E&P Forum): Quantitative Risk Assessment Datasheet Directory

Center for Chemical Process Safety: Process Equipment Reliability Data, American Institute of Chemical Engineers, New York, NY, U.S.A, 1989.

Lees, F.P.: Loss Prevention in the Process Industries, Volume 1, Section 9.2, page 9/6, 3rd Edition, Butterworth-Heinemann, Oxford, U.K., 2005.

Cox, A.W., Ang, M.L., and Lees, F.P.: Classification of Hazardous Locations, IChemE, Rugby, UK, 1990.

Pape, R. P. and Nussey, C.: A Basic Approach for the Analysis of Risk from Major Toxic Hazards - The Assessment and Control of Major Hazards. IChemE Symposium Series No. 93., 1985, IChemE, Rugby, UK, pp. 367-388 (ISBN 0-85295-189-2),

Large Atmospheric Storage Tank Fire Project (1997): LASTFIRE Project Analysis of Incident Frequency Survey, Doc. No.s: OP.97.47084, SIEP.97-5577.

Health and Safety Executive (2001): Offshore Hydrocarbon Release Statistics, HID Statistics Report HSR 2001 002, January 2002.

Health and Safety Executive (2012): Failure rate and event data for use within risk assessments 28-6-2012, www.hse.gov.uk/landuseplanning/failure-rates.pdf

EQUIPMENT FAILURE MODES

Two types of failure equipment failure mode were considered:

- Loss of containment due to leaks from material defects in equipment (pumps, pipework, tanks):
  - flange leak;
  - valve leak;
  - instrument fitting rupture;
  - connection leak;
  - pump seal leak;
  - loading hose/arm leaks;
  - leaks from pressure vessels; and
  - leaks from piping.
- Loss of containment due to process deviations or loss of control:
  - overfill of storage tanks; and
  - overfill of LPG storage spheres.

For each component, credible failure mechanisms and associated hole sizes have been identified:

LEAK RULE-SET FOR LPG EQUIPMENT

Failure Type	Hole Size	Justification	Reference
Pipework (LPG liquid and vapour lines)			
Flange	2.5 mm	Spiral wound gaskets are used for LPG liquid and vapour services. A spiral wound gasket failure results in leaks along the spiral path.	Assumption
Instrument Fittings and Connections	1.4 mm	Instrument fittings on LPG pipework have a 20 mm outer-bore diameter and the internal bore diameter has a 1.4 mm controlling orifice, in accordance with AS 1596. Therefore, instrument fitting failure will result in 1.4 mm leaks.	AS 1596

Failure Type	Hole Size	Justification	Reference
Valve (excluding flanges)	2.5 mm	A valve gland failure for pipes sized 50 mm or larger are typically represented by 10 mm leak orifices. SHEPHERD models flange and valve leaks as one component; therefore, a 2.5 mm hole (see flange leaks) was carried forward.	Cox, Lees, Ang (1991)
Pipe Material Failure	20 mm	Material failure or poor installation may result in a major pipe leak 20 mm in size.	Cox, Lees, Ang (1991)
Pipe Rupture	Full Bore (100 mm)	Excessive stress, corrosion/erosion and impact are potential causes of a full-bore rupture of a pipe. The release size is dependent on the diameter of the pipe, but is limited to 100 mm. NOTE: For pipe failure, SHEPHERD requires only the maximum hole size to be specified and then uses an internal algorithm to determine the frequency of other hole sizes.	Kletz (1990)
Flexible Hose (LPG road gantry)			
Hose	10 mm	A split or tear due to stress and fatigue in the flexible filling hose would cause a 10 mm hole.	Blything and Reeves (1988)
Hose	Full-bore (Exces Flow Valve limited – 1.4 mm)	Breakage of the crimp connection on the flexible filling hose would result in a full-bore release. However, failure of the hose coupling (due to the incorrect connection of the transfer hose to the tanker by the tanker driver) was postulated to be a more credible cause of a full-bore release. The release size is dependent on the transfer (pumping) rate. A release from the tanker would be limited due to activation of the Excess Flow Valve (EFV) on the tanker.	Blything and Reeves (1988) AS 1596
LPG Pump			
Seal	2.5 mm	Mechanical seals limit the leak size due to close tolerances and small bleed points. The leak is typically approximated by 2.5 mm in the worst case.	Assumption
Casing	Full Bore	Catastrophic failure of a casing may be due to external causes (e.g. external impacts, unchecked vibration) resulting in a leak size equivalent to a full bore rupture of the pipework attached to the pump. This scenario is not relevant to Clyde Refinery and has not been carried forward: compressors are not used for recompression of vapour return from LPG road tanker loading and LPG pump seal leaks would not result in a release to atmosphere, since the pumps are submerged within the LPG storage vessels.	Assumption SRAP
LPG Storage Vessel (and LPG Road Tanker)			
Vessel Instrument Fitting	1.4 mm	The instrument fittings on the vessel have a 20 mm outer bore diameter and the internal bore diameter is a 1.4 mm controlling orifice, in accordance with AS 1596. Therefore, a vessel instrument fitting failure will result in a 1.4 mm leak.	AS 1596
Valve (excluding flanges)	2.5 mm	A valve gland failure for valves sized 50 mm or larger are typically represented by 10 mm leak orifices. SHEPHERD models flange and valve leaks as one component; therefore, a 2.5 mm hole (see flange leaks) was carried forward.	Assumption

Failure Type	Hole Size	Justification	Reference
Vessel Failure	-	A 10 mm-equivalent leak could occur from a vessel due to a weld failure (poor quality assurance) however, this is extremely conservative as vessels are built to Class 1H standard with all welds 100% radiographed, 100% NDT (None Destruction Testing) tested and are part of a strict inspection regime. Fatigue, overload, external corrosion or impact may cause an instantaneous release of vessel inventory. (Incident has never occurred in the industry)	Assumption

LEAK RULE-SET FOR NON-LPG EQUIPMENT

Failure Type	Hole Size	Justification	Reference
Pipework			
Flange	2.5 mm	Spiral wound gaskets are used for all flanges in hydrocarbon service at the Refinery. A spiral wound gasket failure results in leaks along the spiral path.	Assumption
Valve (excluding flanges)	2.5 mm	A valve gland failure for pipes sized 50 mm or larger are typically represented by 10 mm leak orifices. SHEPHERD models flange and valve leaks as one component; therefore, a 2.5 mm hole (see flange leaks) was carried forward.	Cox, Lees, Ang (1991)
Instrument Fitting Material Failure	20 mm	Failure of an instrument fitting (typically 20 mm inner-bore diameter) could result in a 20 mm hole size. Material failure or poor installation may result in a major pipe leak 20 mm in size.	Cox, Lees, Ang (1991)
Connection	50 mm	Failure of a connection (typically 50 mm inner-bore diameter) could result in a 50 mm hole size.	
Pipe Rupture	Full Bore (100 mm)	Excessive stress, corrosion/erosion and impact are potential causes of a full-bore rupture of a pipe. The release size is dependent on the diameter of the pipe.	Kletz (1990)
Pump/ Compressor			
Seal	10 mm	Mechanical seals limit the leak size due to close tolerances and small bleed points. The leak is approximated by 10 mm in the worst case.	Assumption
Casing	Full Bore	Catastrophic failure of a casing may be due to external causes (e.g. external impacts, unchecked vibration) resulting in a leak size equivalent to a full bore rupture of the pipework attached to the pump.	Assumption

A review of the source of reliability data is provided below for LPG and Non-LPG equipment.

REVIEW OF FREQUENCY DATA FOR LPG EQUIPMENT

The Shell Shepherd model is provided with default failure frequencies, event tree probabilities and hole sizes as recommended by AEGPL for risk assessments of LPG marketing installations.

The subsequent LPG equipment failure data and accompanying discussion were reproduced from the Shell Shepherd Technical Manual (Ref.22).

The leak frequencies derived from historical data represent the industry average values. Within the industry average there are large differences between the “best” and the “worst” sites. This study adopted the average values.

The table below contains Shepherd default generic LPG failure frequencies. These are supported by the descriptions given in the following sections.

Scenario	Frequency	Unit
pipeline x (L/D)	4.9×10^{-7}	per meter per year
flanges and equivalent valves	5.6×10^{-6}	per flange per year
fittings	5.6×10^{-6}	per fitting per year
hose / hard-arm	6.65×10^{-6}	per operation
transfer rupture	1.0×10^{-5}	per operation

LPG Loading Hose Coupling Failure

A coupling failure, postulated to be due to human error, was considered to be a failure to successfully perform a familiar, highly practised, routine task by an experienced person totally aware of the implications of failure for which HEART (Lees, 2005) gives a human unreliability of 4×10^{-4} per operation.

Pipework extending from the LPG vessels must, in accordance with AS 1596, be provided with Excess Flow Valves (EFVs), which activate in the event that the flow across the valve exceeds 150% of the normal operating flow rate. Hence, should a release occur in the event of a coupling failure, the EFV will close and stop the outflow from the vessel once the flow exceeds the flow rate closure setting of the EFV, i.e. 150%.

In the event that the EFV fails to close, the outflow would not be isolated and the release would continue, albeit restricted by a 1.4 mm controlling orifice, in accordance with AS 1596.

There are no reliable equipment failure data for EFVs. Blything and Reeves adapted an EFV failure probability of 0.013 per demand based on the failure of non-return valves. This value was conservatively carried forward in the analysis, noting that, in view of LPG industry performance, its use in the assessment of EFV failure is considered pessimistic.

Therefore, the probability of a release, given human error in coupling failure, carried forward is 5.2×10^{-6} per operation (i.e. $4 \times 10^{-4} \times 0.013$).

Note: SHEPHERD will multiply the release probability by the frequency of transfer hose couplings performed by an operator, i.e. the number of transfers, per year to determine the leak frequency from this failure mode.

LPG Vessel Catastrophic Failure

The Shepherd Technical manual proposes a 'Catastrophic' LPG vessel failure frequency of 0, based on the fact that there has never been a cold catastrophic LPG vessel failure.

This is supported by world data (Ref.23), which indicate that in 38 million LP Gas vessel-years of operation (to 1993), there has not been a single cold catastrophic failure of an LP Gas vessel.

The Mean Time Between Failures (MTBF) for a component that has no reported failures may be estimated by conservatively assuming that a failure is imminent. Assuming that the failure rate follows a χ^2 (Chi-square) distribution, the Lower Confidence Limit (LCL), i.e. minimum (conservative), value for the MTBF may be estimated from the following equation (Ref.24):

$$\text{LCL of MTBF} = \frac{2T}{\chi^2_\alpha(\nu = 2n)}$$

Where: T is the no. of demands with no failures

ν is the degrees of freedom (i.e. the shape of the distribution)

n is the number of failures

α is the confidence level of interest.

For a single failure, i.e. 2 degrees of freedom, and a conservative confidence level of 60% (as per the methodology adopted in NPRD - Ref.25), the Chi-square percentile is,

$$\chi^2_{0.6}(\nu = 2) = 1.83$$

And:

$$\text{LCL of MTBF} = \frac{2(38x10^6\, years)}{1.83} = 4.15x10^7\, years$$

That is, the Mean Time Between Catastrophic Failures of an LP Gas storage vessel is expected to be greater than $4.15x10^7$ years. Therefore, the frequency of LP Gas vessel catastrophic failures would be expected to be less than $2.4x10^{-8}$ per year (i.e., 1 / ($4.15x10^7$ years)). This value was carried forward.

LPG Pipework

The recommended values for Pipeline Failure Frequencies, in accordance with J. L. Hawksley are given in the table below. The same values are recommended for above and below ground pipelines.

RECOMMENDED VALUES FOR PIPELINE FAILURE FREQUENCIES

Pipe diameter (mm)	Failure frequency rate (x 10^{-6} per metre per year)		
	Full-bore rupture	20% diameter hole	5% diameter hole
50	0.4	0.9	2.1
75	0.3	0.65	1.5
100	0.2	0.5	1.2
150	0.15	0.4	0.9
200	0.12	0.3	0.7
350	0.08	0.15	0.4
400/450/500	0.05	0.11	0.3
600/750/900	0.03	0.07	0.1

The recommended values are at the lower end of generally accepted values for pipework failures.

LPG Storage Overfill

The frequency of overfilling the LPG spheres (7.6 x 10^{-5} pa per sphere) was taken from the site's HEMP Bow Tie Layers of Protection Analyses (LOPA); bow tie no. 10M12H-09.13.

REVIEW OF FREQUENCY DATA FOR NON-LPG EQUIPMENT

Non-LPG Flange / Valve Equivalents Leaks

FLANGE LEAK FREQUENCY DATA

Source	Leak Frequency (per year)	Comment
E&P Forum HC Leak and Ignition Database	8.8×10^{-5}	Based on offshore data. 96% of leaks are of size 0.1D. 4% of leaks are of size D. E&P Forum questions whether a hole size of D is achievable, where D = inside diameter of connecting pipe.
Cox et al.	3×10^{-5} (section leak) 3×10^{-4} (minor leak)	Size of section leak = A, minor leak = 0.1A, where A is the cross-sectional area of the hole defined by the part of the circumference between adjacent bolts and the thickness of the gasket.
Smith in Lees (p. A14/8)	1.8×10^{-4} (lower limit) 8.8×10^{-3} (upper limit)	Failure mode of gasket and equivalent hole size of leak is not given.
Pape and Nussey	3×10^{-6} (0.6mm gasket) 5×10^{-6} (3mm gasket)	Failure is loss of one section between two adjacent bolts. Data was derived for use in an assessment of a chlorine installation.
UK HSE 2001	5×10^{-5}	Based on all failures (approximately 80% of failures are less than 10mm and 90% are less than 25mm)

The E&P forum, HID and Cox at al show good correlation with a leak frequency of between 3×10^{-5} and 8.8×10^{-5} for releases equivalent to the cross section area between bolts on a flange. The data from Pape and Nussey relates to a Chlorine installation where a lower leak frequency would be expected.

From the above data, a failure frequency of 5×10^{-5} per year was chosen for flange leaks. This leak frequency does not take account of the lower expected failure frequencies from spiral wound gaskets. Typically these types of flanges do not suffer from the same 'blow out' failure mode.

VALVE LEAK FREQUENCY DATA

Source	Ref.	Leak Frequency (per year)	Comment
Cox et al.	26	1×10^{-4} (minor leak) 1×10^{-5} (major leak) 1×10^{-6} (rupture leak)	Size of minor leak = 0.01A Size of major leak = 0.1A, Size of rupture leak = A Where A is the cross-sectional area of the pipe. No distinction between valve type.
OREDA-92 (Tax. No. 1.2)	27	External leakage: - critical = 5.2×10^{-3} - degraded = 0.016	Control and safety equipment in offshore application. Crude oil, gas, two-phase fluid or three-phase fluid, including water.
IEEE	28	Manual - 1.8×10^{-4} Motor operated - 8.8×10^{-4} Air operated - 8.8×10^{-4} Check Valve - 4.4×10^{-4}	Typical data source is nuclear industry.
UK HSE 2001		1.7×10^{-4} for all hole sizes	Approximately 80% of releases are 10mm or less.

Applying the hole size distribution of Cox et al, a 2.5 mm leak would be expected to occur with a frequency of $1x10^{-4}$ p.a. This is consistent with the HID results for valve leaks. The leak frequency adopted in this study for leaks from valves in non-LPG service, equivalent to a 2.5 mm hole, was that reported by the HID study (1.7×10^{-4} p.a.).

Due to analogies in their failure mechanisms, leaks from flanges and valves are treated in combination by the Shell Shepherd software. As such, the generic frequency of each was assessed and a frequency for Flange / Valve Equivalents was developed.

The combined frequency of 2.2×10^{-4} per year (5×10^{-5} {flange} + 1.7×10^{-4} {valve}) was carried forward to the QRA. It can be seen that the frequency is dominated by the valve leak frequency.

Non-LPG Instrument Fitting Rupture

In order to cause a significant leak, the failure mode of an instrument fitting needs to be a rupture. Smaller failures would result in minor leaks and spurious instrument readings. There is a scarcity of failure rate data for instrument fittings. Leak frequency data for instrument fittings (also termed small bore fittings) are summarised below:

INSTRUMENT FITTING LEAK FREQUENCIES

Source	Leak Frequency (per year)	Comment
Cox et al.	1×10^{-4} (rupture leak) 1×10^{-3} (major leak)	Size of major leak = 0.1A, rupture leak = A where A is the cross-sectional area of the pipe (given as 10mm).
E&P Forum HC Leak and Ignition Database	3.8×10^{-4} rupture (d/D = 1) 0.3×10^{-4} (d/D = 0.2) 0.5×10^{-4} (d/D = 0.1)	Hole size (d/D) Proportion of leaks in range 0.1 13% 0.2 7% 1.0 80% Where d = equivalent diameter of hole, D = diameter of small bore fitting. Failure data from North Sea oil company.
UK HSE 2001	5.6×10^{-4}	

The failure rate data in the references for rupture leaks are 1×10^{-4} p.a. from Cox et al. and 3.8×10^{-4} p.a. from E&P Forum. The data of Cox et al. is considered more suitable for this analysis since it is based on process industries rather than the offshore oil and gas industry. Therefore, the frequency of instrument fitting leaks used in this study is 1×10^{-4} p.a., equivalent to a 20 mm hole.

Non-LPG Connection Failures

Connections, other than instrument fittings, greater than 1” in diameter, have been treated similarly to fittings; however, it is assumed that the failure frequency of 50mm-equivalent connection failures would be an order of magnitude lower than that for instrument fittings (20 mm equivalent hole size).

Therefore, the frequency of connection leaks used in this study is 1×10^{-5} p.a., equivalent to a 50 mm hole. This value is in agreement with the data obtained from the HID Statistics Report (HSR 2001 002), which reported a connection failure frequency of 1.2×10^{-5} p.a.

Non-LPG Pump Seal Leak

PUMP SEAL LEAK FREQUENCY DATA

Source	Leak Frequency (per calendar year)	Comment
Cox et al.	3×10^{-5} (rupture) 3×10^{-4} (major leak) 3×10^{-3} (minor leak)	Size of minor leak = 0.01A, major leak = 0.1A, rupture leak = A Where A is the cross-sectional area of the pipe.
E&P Forum HC Leak and Ignition Database	1.4×10^{-2} (0 – 10 mm) 2.4×10^{-3} (10 – 50 mm) 6.8×10^{-4} (50 – full bore)	Pumps are double seal centrifugal.
OREDA (Taxonomy No. 1.3 in OREDA-97 and No. 1.3 in OREDA-92)	Significant external leakage: 7.7×10^{-3} External leakage: 0.032	All pumps. Generally centrifugal.
Blything and Reeves	5.2×10^{-4}	Leak is 'total seal failure (effective orifice size = 17 mm without throttle brush, 10 mm with throttle brush)' Data from LPGITA
UK HSE 2001	4.5×10^{-3}	Single Seal

With the exception of the E&P forum minor leak frequency the data sets show a range of values between 10^{-3} and 10^{-4}.

The frequency of seal leaks was taken to be 3×10^{-3} p.a., which corresponds to a minor leak as given by Cox et al and is in line with the HID results.

Non-LPG Pump Casing Failure

Catastrophic failure of a pump casing may be due to external causes, e.g. external impacts or unchecked vibration, and have the potential to result in a leak orifice size equivalent to a full bore rupture of the pipework attached to the pump. Cox et al. (Ref.26) report the frequency of a rupture leak, corresponding to a catastrophic pump casing failure, to be 3×10^{-5} per annum

Non-LPG Process Piping Ruptures

The typical range of pipework failures are addressed in Shepherd as part of the flange, fitting and connection leak scenarios. Therefore, only full-bore ruptures of pipework are modelled in Shepherd.

Leak frequency and equivalent hole size data for process piping is often expressed as a function of the pipe diameter. The tables below summarise full-bore leak frequency data from a number of sources taken from the refinery FSA (2000), the HID report and UK HSE Land-Use Planning Data. Where the hole diameters were expressed as a function of pipe diameter, these were converted to an actual hole size for comparison.

FULL-BORE LEAK FREQUENCIES FOR PIPING UP TO 275 MM IN DIAMETER

Source	Hole Size (mm)	Full Bore Leak Frequency (per metre-year)
Cox et al.	100 to full bore	1.5×10^{-7}
E&P Forum	100 to full bore	1.8×10^{-6}

Source	Hole Size (mm)	Full Bore Leak Frequency (per metre-year)
Gulf data in Lees	Rupture	2.5×10^{-8}
UK HSE 2001	Over 100	7.6×10^{-6}
UK HSE 2012	150 - 299	2×10^{-7}

FULL-BORE LEAK FREQUENCIES FOR PIPING OVER 275 MM IN DIAMETER

Source	Hole Size (mm)	Full Bore Leak Frequency (per metre-year)
Cox et al.	300	5×10^{-8}
E&P Forum	230 to full bore	1.4×10^{-6}
UK HSE 2001	None reported	None reported
UK HSE 2012	300 – 499	7×10^{-8}

The full-bore (rupture) pipe leak frequencies selected for the study are summarised below. They were selected principally on the data from the Health and Safety Executive (2012) *Failure rate and event data for use within risk assessments*.

FULL-BORE (RUPTURE) PROCESS PIPING LEAK FREQUENCIES USED IN QRA

Pipework	Diameter Range	Full-Bore Leak Frequency (per metre-year)
Medium piping	D < 300 mm	2×10^{-7}
Large piping	D ≥ 300 mm	7×10^{-8}

TANK ROOF FIRE FREQUENCY

The LASTFire (Large Atmospheric Storage Tank Fire) Study (Ref.29) was initiated by a consortium of 16 oil companies (including Shell) to review the risks associated with large diameter open-top floating roof storage tanks, since it was recognised that the fire hazards associated with large tanks were insufficiently understood to be able to develop fully justified site specific fire response and risk reduction policies.

The LASTFIRE Project objectives were to: determine the current levels of risks associated with fires in large open top floating roof storage tanks; provide techniques to enable individual operators to determine their level of fire-related risk, identify effective risk reduction measures, and to establish recommended design and operation practice and to make this knowledge available throughout the industry.

The LASTFire study reports that there have been 4 full surface tank roof fires in 33,909 tank-years yielding a tank roof fire frequency of 1.2×10^{-4} pa. This value was carried forward as the baseline tank fire frequency for gasoline and jet fuel tanks.

The study identified the potential for escalation between tanks, but the resulting tank fires would not have potential for off-site impact.

TANK BUND LEAK FREQUENCY

The tank bund leak frequency is comprised of tank leak and tank overfill frequencies, as described below.

Tank Leaks

The LASTFire study collated incident (loss of containment) data supplied by the participating oil companies and reviewed the collective incident history of 2420 tanks. The study found that in 33,909 tank-years of operation, there had been 96 'releases outside the tank shell'; however, only 2 of those releases were categorised as major spills which resulted in large bund fires (at a frequency of 5.9×10^{-5} per tank-year). The major spills were caused by:

- tank mixer falling off, resulting in a major loss of containment and ignition; and
- nitrogen injection into the tank, which displaced the tank contents and overflowed the tank (which was recorded as a 'tank overfill' scenario).

The scenarios above are not relevant to the Clyde Terminal, since the tanks are provided with neither a mixer nor nitrogen injection.

A review of the findings of the LASTFire study was undertaken to determine if a representative leak frequency could be established. The LASTFire study identified the following causes of a 'release outside the tank shell' (and the associated frequency), but did not provide guidance on the size of the releases.

STORAGE TANK RELEASES OUTSIDE THE TANK SHELL

Failure Mode	No. of Occurrences Leak	Average Leak Frequency (per tank-year)
Pipework/ flange/ valve leak	16	5×10^{-4}
Corrosion of roof leg pad	1	0.3×10^{-4}
Steam coil failure	3	0.9×10^{-4}
Mixer failure/ leak	9	3×10^{-4}
Overfill	14	4×10^{-4}
Corrosion of tank bottom	15	4×10^{-4}
Bottom annular plate corrosion	4	1×10^{-4}
Roof drain valve failure	13	4×10^{-4}
(Not recorded)	21	6×10^{-4}
TOTAL	**96**	**2.8×10^{-3}**

Catastrophic tank (strake or structural) failures, which are generally considered to be not credible, are not reported in the LASTFire study and have not been considered further.

The likelihood of a major release to the bund due to leaking pipework, instrument fittings, valves and flanges, is considered low. Such equipment failures would typically result in small pools, within the bund, which would be detected during routine operator surveillance and remedial action would be taken prior to the potential realisation of a catastrophic release. This cause was not carried forward.

Corrosion of the roof leg pad may occur, however such a failure would result in a release onto the roof of the tank and may lead to roof failure and sinking. Therefore, roof leg failure is considered to be a cause of a tank roof fire (not bund fire) and was not carried forward.

The jet fuel tanks will not be provided with a steam coil or a mixer.

The majority of the overfill incidents reported in the LASTFire study (14 off) were not categorised as major spills that resulted in large bund fires. As noted above, major spill due to tank overflow was observed for a single incident in which the tank contents were displaced during nitrogen injection – this scenario is not considered relevant.

Corrosion of the roof drain valve is identified in the LASTFire data as a potential cause of release outside the tank; however, the study does not indicate the nature of the released material (i.e. does not specify if it the releases were water or hydrocarbon) and does not indicate the volumes released. Notwithstanding this, the roof drain valves of floating roof tanks at the Terminal are left open at all times, such that rain water is allowed to flow freely into the bund drain. The LASTFire data relating to ‘valve failure’ is, therefore, not considered appropriate.

There is potential, however, for the roof drain line hinge (located inside the tank) to fail and allow product to enter the drain line. The Inspection/ Maintenance team provided anecdotal evidence of such failures occurring: in all cases, product was identified to be ‘dribbling’ out of the open drain valve (i.e. no major releases have been observed). In all cases, the minor leaks have been identified by operators during their daily surveillance or monthly inspection of the tank compounds. The procedure in place for these events involves isolation of the drain valve until the hinge can be repaired. The tank is also listed on an official register to ensure that its drain valve is opened (under a controlled procedure) during or following a period of rain. Given the historical evidence of the low potential for large releases (that may lead to the bund being filled) and the safeguards proposed (daily surveillance, monthly inspection and tank level gauging to detect movement of the product level in the tank), failure of the drain line hinge was not considered to be a credible cause of the bund being filled.

The LASTFire generic frequency data for releases caused by corrosion of the tank bottom or bottom annular plate were considered appropriate, since Shell has in place controls/ safeguards that are considered industry best-practice, viz.:

- 10-yearly internally inspections;
- Tanks will be water drained following every transfer;
- Bunds will be water drained after every rain event;
- Monthly operator tank and bund inspections; and
- Daily operator tank and bund surveillance.

Tank integrity inspections and non-destructive (magnetic particle) tests are effective in identifying corrosion and predicting incipient failures.

It should be noted that the majority of the tanks reviewed in the LASTFire study contained a large proportion of tanks that were used for un-refined (and potentially corrosive) crude oil. Therefore, application of the LASTFire data to the Clyde Terminal is conservative, given that:

- jet fuel and gasoline are negligibly corrosive compared to crude oil; and
- finished product that contains no water will be stored.

Corrosion of the tank bottom or bottom annular plate could potentially cause leaks underneath the tank and may lead to contamination of the ground and tank subsidence.

Corrosion is managed through the formal internal/ external inspections, monthly operator inspections and daily operator surveillance of tanks/ bunds. It is noted that inspection frequencies are based on the risk of leaks occurring from corrosion; that is, it is recognised that corrosion failures are incipient and are usually detectable within the inspection period and before a major failure (and large release) occurs. Larger corrosion leaks occurring beneath the tank would also be detected by the tank level gauging system.

If corrosion of the tank bottom or bottom annular plate occurs some time after the last internal or external inspection, filling of the bund may occur if the corrosion failure remains undetected for an extended period of time.

Failure to detect the corrosion during the formal monthly operator inspections was considered to be a human error. A human error probability of 0.003 (per Shell HEMP methodology) was adopted.

The LASTFire study found the frequency of releases due to corrosion of the tank bottom or bottom annular plate to be 5×10^{-4} per tank-year. Therefore, the likelihood that corrosion of the tank bottom or bottom annular plate would lead to a full-surface bund leak is 1.5×10^{-6} per year (i.e. $5 \times 10^{-4} \times 3 \times 10^{-3}$). This value is considered conservative in that the calculation assumes that, if the operator does not detect the leak in the first monthly inspection after the failure occurs, then a large release would occur. In reality, the rate of corrosion propagation would be relatively slow and the operator would have more than one opportunity to identify a leak (i.e. during subsequent monthly inspections or during daily surveillance, with perhaps more prominent symptoms) before a large release could occur.

Tank Overfill

The frequencies of tank overfills were taken from the site's HEMP Tank Overfill Model Bow Tie Layers of Protection Analyses (LOPA), which considers two threats:

- Failure of the automatic tank gauging (ATG) system leading to either:
  - incorrect ullage calculation prior to tank filling and hence the risk of too much product directed to the tank; or
  - incorrect level measurement during tank filling and hence the risk of a false indication of the fill level in the tank and incorrect level alarms from the tank gauging system.
- Valve line-up error during tank change-overs to tank filling leading to:
  - Changing the product flow path to, and thus filling, the incorrect tank (which may be full); or
  - Failing to change the product flow path and thus continuing to fill the previously filled tank.

A summary of the tank overfill LOPA calculation is given in TABLE C.1 and a description of the analysis methodology is given thereafter.

TABLE C.1 SUMMARY OF TANK OVERFILL LOPA

Tank Number	Service	Type of High Level Alarm	Type of High-High Level Alarm	Frequency of Tank Line-ups (pa)	PFD Line-up	IEF (A) Level Gauge Failure (pa)	IEF (B) Valve Line-up Error (pa)	PFD High Level Alarm & Operator Response C2(B)	PFD Hi-Hi Level Alarm & Operator Response C3	PFD High Level Trip C4	PFD Rough Dips C5(A)	PFD Independent Valve Line-up Check C6	Overfill Frequency (pa)
32	Diesel	3-Radar	Radar	23	0.001	0.02	0.023	0.1	0.1	1	0.1	0.1	2.2E-04
33	Diesel	3-Radar	Radar	23	0.001	0.02	0.02	0.1	0.1	1	0.1	0.1	2.2E-04
34	Jet fuel	3-Radar	Radar	19	0.001	0.02	0.02	0.1	0.1	0.1	0.1	0.1	2.2E-04
35	Jet fuel	3-Radar	Radar	19	0.001	0.02	0.02	0.1	0.1	0.1	0.1	0.1	2.2E-04
36	Petrol	3-Radar	Radar	13	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
37	Petrol	3-Radar	Radar	13	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
38	Petrol	3-Radar	Radar	13	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
39	Petrol	3-Radar	Radar	13	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
42	Jet fuel	3-Radar	Radar	19	0.001	0.02	0.02	0.1	0.1	0.1	0.1	0.1	2.2E-04
50	Petrol	3-Radar	Radar	9	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
51	Diesel	3-Radar	Radar	23	0.001	0.02	0.02	0.1	0.1	1	0.1	0.1	2.2E-04
53	Petrol	3-Radar	Radar	9	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
82/ 91/ 92/ 103/ 105	Slops	3-Radar	Switch	1	0.001	0.02	0.00	0.1	0.1	0.1	0.1	0.1	2.0E-04
84	Petrol	3-Radar	Radar	13	0.001	0.02	0.01	0.1	0.1	0.1	0.1	0.1	2.1E-04
86	Petrol	3-Radar	Radar	37	0.001	0.02	0.04	0.1	0.1	0.1	0.1	0.1	2.4E-04
87	Petrol	3-Radar	Radar	37	0.001	0.02	0.04	0.1	0.1	0.1	0.1	0.1	2.4E-04
90	Petrol	3-Radar	Radar	65	0.001	0.02	0.06	0.1	0.1	0.1	0.1	0.1	2.6E-05

With reference to the column headings in TABLE C.1, the tank overfill Model Bow Tie Layers of Protection Analysis (LOPA) was quantified in the context of the following:

- CM1 was not considered valid and no credit was given in either Initiating Event A or B.
- CM2 is not independent of Initiating Event A, ATG Failure, since the High Level Alarm is generated by the ATG system (which by definition has failed in Event A). No credit given for CM2 in Initiating Event A.
- CM3 is independent of the ATG so credit for both Initiating Events A and B.
- CM4 will be in place prior commencing operations in this mode and credit is given.
- CM5 requires that the site has in place a procedure for follow-up of deviations found with rough dips or indications that High Level Alarm (i.e. from the ATG) is not working. This is achieved via a single yearly tank dip to calibrate the ATG and daily stock reconciliation. This constitutes an alternate means of measuring tank level, which satisfies the intent of the barrier and thus valid.
- CM6 is in place and credit is given.

The LOPA calculation to evaluate overfill frequency can be summarised as follows:

Overfill frequency = [(IEF_AxCM3xCM4xCM5) + (IEF_BxCM2xCM3xCM4xCM6)]

Where:

IEF_A	=	Failure rate of ATG (1/30 pa for servo-gauge type or 1/50 pa for radar gauge type)
IEF_B	=	Frequency of tank line-ups x PFD Line-Up (i.e. the human error probability per critical line-up = 0.001 if there is written line-up procedure with a checklist)
CM2	=	Probability of failure on demand (PFD) of High Level Alarm (from ATG) & Operator Response = 0.1
CM3	=	Probability of failure on demand (PFD) of Independent High-High Level Alarm (from switch) & Operator Response = 0.1
CM4	=	PFD of High-High Level Trip = 1
CM5	=	PFD of procedure to follow up deviations from rough dips = 0.1
CM6	=	PFD of independent valve line-up check = 0.1

Note: the immediate ignition probabilities (for bund fire) are per Table 8.1. The Shepherd model applies the delayed (area-density) ignition probabilities based on the dispersion simulations at various wind speeds and directions to determine the frequency of vapour cloud fires (or explosions, if ignited from within congested areas).

TABLE C.2 SUMMARY OF RELIABILITY DATA

Equipment Item	Representative Release Orifice			Leak Frequency
	Diameter (mm)	**Justification**	**Reference**	**Frequency (per item-year)**
LPG Equipment				
Flanges and equivalent valves	2.5	Spiral wound gaskets are used for LPG liquid and vapour services. A spiral wound gasket failure results in leaks along the spiral path, assumed to be equivalent to a 2.5 mm hole. Valve gland failure, for pipes sized 50 mm or larger, are typically represented by 10 mm leak orifices. SHEPHERD models flange and valve leaks as one component; therefore, a 2.5 mm hole was carried forward.	Assumption Cox, Lees, Ang (1991) Assumption	5.6×10^{-6}
Instrument Fittings and Connections	1.4	Instrument fittings on LPG pipework have a 20 mm outer-bore diameter and the internal bore diameter has a 1.4 mm controlling orifice, in accordance with AS 1596. Therefore, a pipe work instrument fitting failure will result in a 1.4 mm leak.	AS 1596	5.6×10^{-6}
Pipe (including pipelines) x (L/D)	20 Full Bore (100 mm)	Material failure or poor installation may result in a major pipe leak 20 mm in size. Excessive stress, corrosion/erosion and impact are potential causes of a full-bore rupture of a pipe. The release size is dependent on the diameter of the pipe, but is limited to 100 mm **NOTE:** For pipe failure, SHEPHERD requires only the maximum hole size to be specified and then uses an internal algorithm to determine the frequency of other hole sizes.	Cox, Lees, Ang (1991) Kletz (1990)	4.9×10^{-7}
Hose / hard-arm (hose failure)	10	A split or tear due to stress and fatigue in the flexible filling hose would cause a 10 mm hole.	Blything and Reeves (1988)	(6.65×10^{-6} per operation[Note 1])
Hose / hard-arm (coupling failure)	Full-bore (Excess Flow Valve limited – 1.4 mm)	Breakage of the crimp connection on the flexible filling hose would result in a full-bore release. However, failure of the hose coupling (due to the incorrect connection of the transfer hose to the tanker by the tanker driver) was postulated to be a more credible cause of a full-bore release. The release size is dependent on the transfer (pumping) rate. A release from the tanker would be limited due to the 1.4 mm controlling orifice (AS 1596), in the event of failure of the Excess Flow Valve on the tanker.	Blything and Reeves (1988) AS 1596	(5.2×10^{-6} per operation[Note 1])

Equipment Item	Representative Release Orifice			Leak Frequency
	Diameter (mm)	Justification	Reference	Frequency (per item-year)
Catastrophic vessel failure	-	-	-	2.4×10^{-8}
Non-LPG Equipment				
Flanges and equivalent valves	2.5	Spiral wound gaskets are used for all flanges in hydrocarbon service at the Refinery. A spiral wound gasket failure results in leaks along the spiral path. Valve gland failure, for pipes sized 50 mm or larger, are typically represented by 10 mm leak orifices. SHEPHERD models flange and valve leaks as one component; therefore, a 2.5 mm hole was carried forward.	Cox, Lees, Ang (1991)	2.2×10^{-4}
Instrument fitting (< 1" diameter)	20	Failure of an instrument fitting (typically 20 mm inner-bore diameter) could result in a 20 mm hole size. Material failure or poor installation may result in a major pipe leak 20 mm in size.	Cox, Lees, Ang (1991)	1×10^{-4}
Connection (> 1" diameter)	50	Failure of a connection (typically 50 mm inner-bore diameter) could result in a 50 mm hole size.	Assumption	1×10^{-5}
Pump Seal	10	Mechanical seals limit the leak size due to close tolerances and small bleed points. The leak is approximated by 10 mm in the worst case.	Assumption	3×10^{-3} (single seal)
Pump Casing Failure	Full Bore	Catastrophic failure of a casing may be due to external causes (e.g. external impacts, unchecked vibration) resulting in a leak size equivalent to a full bore rupture of the pipework attached to the pump.	Assumption	3×10^{-5}
Pipe Rupture	< 300	Excessive stress, corrosion/erosion and impact are potential causes of a full-bore rupture of a pipe. The release size is dependent on the diameter of the pipe.	UK HSE (2012)	2×10^{-7}
	≥ 300		UK HSE (2012)	7×10^{-8}
Atmospheric Storage Tanks				
Full-surface tank fire	-	The LASTFire study reports that there have been 4 full surface tank roof fires in 33,909 tank-years.	LASTFire	1.2×10^{-4} (fire)
Tank Overfill	-	-	-	(see Table C.1)

APPENDIX D. CONSEQUENCE EFFECT & IMPAIRMENT CRITERIA

The impairment criteria adopted for the study are described in Section D.1 and the QRA model input data (including the consequence modelling findings) is provided in Section D.2.

D.1 Impairment Criteria

Impairment criteria were used to determine the effects that the physical consequences may have on defined receivers, viz.:

- Offsite Populations
- Structures and Equipment

The following types of effects were assessed:

- Thermal Radiation
- Explosion Overpressure

The type of effect for each receiver is discussed in this Appendix.

The selection of criteria is based on the inputs that are required for the Shell Shepherd QRA modelling software adopted for this study and a review of the existing refinery studies.

D.1.1 Thermal Effects on People

The effect of thermal radiation on people is a function of the incident heat flux and time of exposure. TNO Green Book (and is quoted in Lees, Ref.30) suggests the following probit (Y) equation for personnel protected by normal clothing:

$$Y = -37.23 + 2.56\ln(tI^{4/3})$$

For average exposure durations (times to escape), the probit equation gives the following:

TABLE D.1 INCIDENT HEAT FLUX FOR VARIOUS FATALITY LEVELS

Probability of Fatality	Incident Heat Flux (kW/m^2)		
	120 seconds exposure	60 seconds exposure	30 seconds exposure
1%	3.3	5.5	9.3
10%	4.5	7.5	12.7
50%	6.5	11.0	18.4
90%	9.5	15.9	26.8
99%	13.2	22.1	37.2

In terms of fatality calculations the SHEPHERD model represents a jet fire as a cone with three fatality zones. The user defines the heat radiation level and the fatality probability in each zone.

Based on the results above the following three zones were selected:

Heat Radiation Fatality Zone	Fatality Probability
>14 kW/m^2	100% fatality
Between 14 and 4.7kW/m^2	50% Fatality
< 4.7 kW/m^2	Injury

D.1.2 Explosion Overpressure Effects on People

For the purpose of calculating the total (location specific) risk contours in Shepherd, the effects on people (in terms of fatalities) from vapour cloud explosion overpressure are accounted for by the fireball consequence size (i.e. personnel within the flash fire are assumed to be fatalities). The Shepherd Technical Manual (Ref.31) cites the following rationale in justifying this modelling technique:

- "There are no validated models available [for the effects of overpressure from hydrocarbon explosions on people]. There are various Probit relations available in the open literature, which are largely based on a variety [of] explosion incidents. These explosions are not limited to the Vapour Cloud Explosions that the Risk Tools models"
- "It is difficult, if not impossible, to accurately model the effects of projectiles on people in a simplified modelling environment as presented by the Risk Tool"
- "Overpressure in the Risk Tool is a result of a vapour cloud travelling into a congested area and finding a source of ignition (i.e.: VCE). It can be argued that the cloud fire will yield higher number of fatalities than the explosion; not counting overpressure helps in preventing double counting"

Further clarification was sought from Shell Global Solutions (the Shepherd software developer) on the rationale above. Shell Global Solutions provided, in particular, the following additional comments (Ref.32):

> *With regard to overpressure causing fatalities, there are Probits available in the TNO Green Book which correlate impulse and overpressure [to potential for fatality], this substantially demonstrates that unless you are extremely close to the source of the explosion you will not become a fatality. The cause of fatality is quoted as lung collapse, head damage and whole body displacement and there are individual Probits for each effect.*
>
> *[Normally,] this level of overpressure / impulse is extremely unlikely to occur off-plot due to the decay of the blast wave, [usually] only on-site workers will be affected. This is also confirmed by real incidents, take for example the [recent] BP Texas incident, [wherein] people in the open who were outside of the cloud fire were knocked over by the blast, but substantially walked away unharmed. The majority of those who were caught in the cloud fire became fatalities. It is actually quite clear cut, almost a step function at the edge of the cloud fire. The other fatalities were people in buildings which collapsed on them; people who were outside and substantially closer to the incident survived the overpressure.*

D.1.4 Thermal Effects on Equipment and Structures

Shell FRED contains a heat-up model. The Clyde Refinery Formal Safety Assessment (Ref.33), conducted in 2000, identified 9 classes of fire (representing jet, spray and pool fires for different hydrocarbon releases) and 30 target vessels. The target vessels were rationalised based on the material type and wall thickness, and heat up calculations were executed. The conclusion of the study was a representative rule set for critical time to vessel failure due to fire. This rule set was verified using Vessfire and carried forward for this study:

TABLE D.3 RULE SET FOR CRITICAL TIME TO VESSEL FAILURE

Fire Scenario	Wall Thickness (mm)	Time to Failure (minutes)
Jet / Spray	< 20	2
	> 20	6
Pool fire (leak rate > 20kg/s)	< 20	10
	> 20	30
Pool fire (leak rate < 20kg/s)	< 20	5
	> 20	10
Non-impinging fire	< 20	30
	> 20	60

D.1.4 Explosion Overpressure Effects on Equipment and Structures

The Shepherd model allows the user to input the critical overpressure (i.e. the point of failure) for equipment located in a specific area. The effects of explosion overpressure are reported in HIPAP 6. The relationship between the explosion overpressure and the terminal locations is shown in the table.

TABLE D.4 RULE SET FOR CRITICAL OVERPRESSURE

Explosion Overpressure	Effect	Refinery Location
14kPa	Damage to house	All Areas
21kPa	Rupture of storage tanks	Movements

APPENDIX E. RECORD OF INCORPORATION OF NSW DPI COMMENTS

A draft version of this PHA Report (Clyde Terminal Conversion Project, Clyde Refinery Site, Document No J20648-001, rev. A, dated August 2012) was provided to the NSW Department of Planning and Infrastructure (DPI) for initial review and comment prior to formal submission.

Comments were received from Lilia Donkova (Lilia.Donkova@planning.nsw.gov.au), on behalf of the NSW DPI, by email on Friday 14/09/2012 10:27 AM.

The following table summarizes the updates made to address DPI's comments.

Item	NSW DPI Comments	Close-Out Discussion
1	Sufficient information on the equipment and the activities to be undertaken on site. For example:	Report includes additional information on activities to be undertaken on site (see below):
1.1	Details on the pipeline from the Wharf 1 at Gore Bay, including operating pressure, diameter, etc.;	Additional information is provided in Section 3.4.
1.2	Details on the surrounding land uses, including distance to the nearest residential and sensitive land use, where applicable;	Additional information is provided in Section 3.6.
2	During the presentation given by Shell to the Department it was stated that both developments would upgrade the safeguards and would increase the automation on site. It is recommended details on the upgrade to be provided.	Additional information is provided in Section 10.4.
3	Information on compliance with relevant standards and in particular AS 1940 and AS 2885. If compliance cannot be demonstrated, then information on the alternative (existing and proposed) safety measures in place to ensure the same or higher level of safety to be provided.	This has not been addressed herein. Queries regarding the degree of compliance with Australian Standards should be directed to the project engineering managers.
4	All DGRs must be addressed.	A PHA has been completed, including a discussion on how lessons from the Buncefield Incident have been incorporated (Section 4.0).
5	The analysis undertaken for selection of the appropriate frequencies. (The frequencies of the equipment should be based on review of the available data and should demonstrate that they are appropriate for the type and the age of the facility.)	More information regarding the review and selection of frequency data is now provided in APPENDIX C.
6	The assumption of the PHA, for example:	The assumptions in Section 2.6 have been expanded.
6.1	The integrity of a bund in an event of sudden loss of containment;	As above.
6.2	Drawings representing the arrangement of tanks within the bunds;	Figure 3.4 shows detail of tanks within bunds.

Item	NSW DPI Comments	Close-Out Discussion
6.3	Justification on the assumption to adopt thermal radiation of 14 kW/m^2 as end point for calculation of the distance to fatality. It is noted (Table D1 of Gore Bay report) that the probability of fatality calculated is 99% at 13.2 kW/m^2 for 120 sec of exposure. The PHA assumes 100% probability of fatality at heat radiation higher than 14kW/m^2. No information on assumed time of exposure in SHEPERD model is provided.	Based on the probit-evaluated heat fluxes (see APPENDIX D Section D.1.1) for 99% and 99.9% chance of fatality for 120s exposure (i.e. 13.2 kW//m^2 and 16.2 kW//m^2, respectively), a value of 14 kW/m^2 was conservatively assumed to cause 100% chance of fatality.
6.4	Time for isolation (in case of the pipe leak for example)	The leak detection and isolation time is discussed in APPENDIX D (Section D.1.1).
7	Methodology/theory for modelling – for example view factor or point source for was used for the pool fires	The view factor method, using Shell FRED software, was adopted as described in Section 7.2.
8	Details on the consequence modelling to allow understanding of the models used, such as:	Further details have been provided as requested (see below):
8.1	SEP of the flame for pool fires (if a view factor model is used)	SEPs are now provided in Table 7.2, Table 7.4, Table 7.5 and Table 7.7.
8.2	The assumed diameter for bund fires	Equivalent pool diameters are now provided, if evaluated in the study (see Table 7.2, Table 7.4, Table 7.5 and Table 7.7).
9	Based on the previous modelling undertaken by the Department and on the information provided for similar developments consequence distances to fatality for pool fires appears to be optimistic (in the range of 1-3 m). It is recommended the parameters and the assumptions of the models to be revisited.	It is unclear as to which analyses that the DPI comments apply; however, all consequence analyses have been revised.
10	The assumption that full bund fire will occur only as a result of overfill appears to be optimistic. Catastrophic failure of a tank, even though more unlikely event, should be considered.	Leaks into the bunds, other than overfill, have now been considered in detail (see APPENDIX C).
11	It is noted that the total effects on people in terms of fatalities from VCE are accounted for by the fireball consequence size. However this methodology does not account for injuries as a result of explosion overpressure. Account for injuries due to explosion should be made.	The PHA considers injury due to explosion overpressure.
12	Discussion on the potential for irritation due to combustible products of fires.	Discussion regarding combustion products is provided in Section 7.5.
13	Sec 6.4 states that the off-site impacts are based on a conservative assumption of the exposure time and the vulnerability of the people. Please provide details on the assumptions.	Section 6.4 has been deleted.
14	A map showing the location of Basel facility in respect to the refinery should be provided.	This is now provided in Figure 3.3.

APPENDIX F. REFERENCES

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