NEW RISK ASSESSMENT

In terms of:

THE MAJOR HAZARD INSTALLATION REGULATIONS

Prepared as per SANS 1461:2018 for:

..SFF Association

For the installations at:

Saldanha Bay

By:

MHR Consultants

Major Hazard Risk Consultants

Nominated Representative:	C C Thackwray
Technical Signatory:	T C Thackwray

21 February 2019

GOVERNMENT
APPROVED
INSPECTION
AUTHORITY
No MHI 0007

sanas
Inspection Body

MHI 0017

DETAILS AND CONTROL PAGE

TYPE OF ASSESSMENT							
New Installation		**Changes to Existing Installation**	**x**	**5 Year Renewal**	**x**	**Other**	

Name	Strategic Fuel Fund
Address	Extension of Camp Road Saldanha Bay
Contact Person	Nazier Cassiem Tel: 021 524 2715
Date of Assessment	22 January 2019
Date of Report	21 February 2019
Dates of Previous Assessments	<table><tr><th>Date</th><th>Reference</th><th>AIA</th></tr><tr><td>Nov 2013</td><td>SFF001</td><td>MHR</td></tr><tr><td></td><td></td><td></td></tr><tr><td></td><td></td><td></td></tr><tr><td></td><td></td><td></td></tr><tr><td></td><td></td><td></td></tr></table>
Technical Signatory	T C Thackwray 13 Slade Street Parklands North Tel: 083 746 8933
Reference Number	DW003
Revision	First Revision

This is to verify that although the MHI Risk Assessment has been completed according to the MHI Regulations and SANS 1461, it is intended as a Specialist Report for the Environmental Impact Assessment. The risks associated with the MHI were found to be acceptable.

This Risk Assessment is valid for the duration of 5 years from the above date, unless:

- Changes have been made to the plant that can alter the risks on the facility;
- The emergency plan was invoked or there was a near miss;
- The changing neighbourhood could result in offsite risks;
- There is reason to suspect that the current Assessment is no longer valid.

Signed

TC THACKWRAY
TECNICAL MANAGER

A QUANTITATIVE RISK ASSESSMENT OF THE TOXIC AND FLAMMABLE INSTALLATIONS ON THE PREMISES OF SFF IN SALDANHA AND THE PORT OF SALDANHA

EXECUTIVE SUMMARY

1. INTRODUCTION

The Strategic Fuel Fund Association (SFF) was established in 1964 as a Section 21 company. The first crude oil it ever owned was stored in its Milnerton tanks in 1967. In 1975, a levy on fuel payable into the Central Energy Fund was introduced and managed by CEF (Pty) Ltd and a portion of this was paid to SFF.

In 1976, the construction of the Saldanha tank farm began and in 1978 United Nations oil sanctions were imposed against South Africa.

In 1979, the world suffered the second major international oil crisis. That was the year when the procurement of crude oil by SFF on behalf of the oil companies began and when the first Saldanha oil tank was completed. The whole Saldanha project was completed in 1982.

SFF's oil storage installation at Saldanha is the largest facility of its kind in the world. It comprises six in-ground concrete storage tanks which have a combined capacity of 45 million barrels of oil. The tanks are linked by pipeline to an oil terminal at the Port of Saldanha where tankers either discharge or backload oil. A second 107 km long pipeline, belonging to the refinery in Milnerton links the oil storage facilities with the refinery. Crude oil is also stored for other companies and delivered to ships via the pipeline to the port.

LPG has been declared a strategic fuel and SFF is proposing to build a bulk LPG import storage terminal on their premises linked by a pipeline to a proposed LPG import terminal at the Port of Saldanha.

As crude oil and LPG have the potential to cause onsite and offsite incidents, Major Hazard Risk Consultants cc was commissioned to conduct a Risk Assessment in accordance with the Major Hazard Installation Regulations to determine the impact of the facilities on the surrounding area.

This investigation would serve as a basis for the notification of the facility in accordance with the Major Hazard Installation Regulations. The purpose of this report is to convey the essential details, including a short description of the hazards, the receiving environment, the design, the risks and consequences of an accident.

The main aim of the investigation was to quantify the risks to employees and neighbours regarding the facilities in Saldanha Bay.

The focus of this report is on the risks associated with the off-loading of crude oil and LPG at the Port of Saldanha, the transportation of crude oil and LPG via pipeline to and from the port to the bulk storage facility as well as the bulk oil and LPG storage installations.

Risk is the severity of the consequence of a hazardous event and the probability of the event occurring.

This Risk Assessment was conducted in accordance with the Major Hazard Installation Regulations and SANS 1461:2018 and could be used as notification of the facility. The Risk Assessment includes the following:

- Identifying likely hazards associated with the processes of the installations including the causes, consequences and their effects;
- Quantifying the likely hazards in terms of their magnitude;
- Quantifying the consequences for each hazard (thermal radiation, domino effect, toxic cloud formation, etc.);
- Determining the lethality of the effects of the consequences;
- Determining the frequency of all the hazardous events;
- Calculating the individual risk values considering all accidents, meteorological conditions and lethality;
- Using the population density around the facility to determine the societal risk posed by the facility;
- Reporting on the risks in terms of internationally acceptable criteria;
- Providing an assessment of the adequacy of emergency response programmes, fire prevention and fire-fighting measures;
- Proposing measures to reduce or eliminate the risks.

The Risk Assessment may not meet the requirements of environmental legislation as it is not intended as an Environmental Risk Assessment.

2. CONCLUSIONS

This Assessment modelled the risks of the existing crude oil installations and proposed LPG installations in the port, the pipeline corridor and the SFF bulk terminal facility. The total individual risks are as follows:

Jetty at the Port of Saldanha

- The total individual risk involving the current crude oil jetty loading/offloading operations in the port is acceptable, with the 1.0e-6 (one-in-a-million) contour extending 12m beyond the Saldanha side of the jetty edge. The 3.0e-7 (one-in-thirty-million) contour extends 21m beyond the Saldanha side of the jetty edge and does not reach any sensitive areas or MHIs.

- The total individual risk involving the proposed LPG jetty loading/offloading operations in the port is acceptable, with the 1.0e-6 (one-in-a-million) contour extending 34m beyond the Langebaan side of the jetty edge. The 3.0e-7 (one-in-thirty-million) contour extends 81m beyond the Langebaan side of the jetty edge and does not reach any sensitive areas or MHIs.

- The total individual risk involving the current crude oil and the proposed LPG jetty loading/offloading operations in the port is acceptable, with the 1.0e-6 (one-in-a-million) contour extending 35m beyond the jetty edge. The 3.0e-7 (one-in-thirty-million) contour extends 85m beyond the jetty edge and does not reach any sensitive areas or MHIs.

Pipeline to the SFF Import Terminal Facility

- The total individual risks involving the current crude oil pipelines are acceptable, with the 1.0e-6 (one-in-a-million) contour extending 25m beyond pipelines. The 3.0e-7 (one-in-thirty-million) contour extends 45m beyond the pipelines and does not reach any sensitive areas or MHIs.

- The total individual risks involving the proposed LPG pipeline are acceptable, with the 1.0e-6 (one-in-a-million) contour extending 25m beyond the. The 3.0e-7 (one-in-thirty-million) contour extends 60m beyond the pipeline and does not reach any sensitive areas or MHIs.

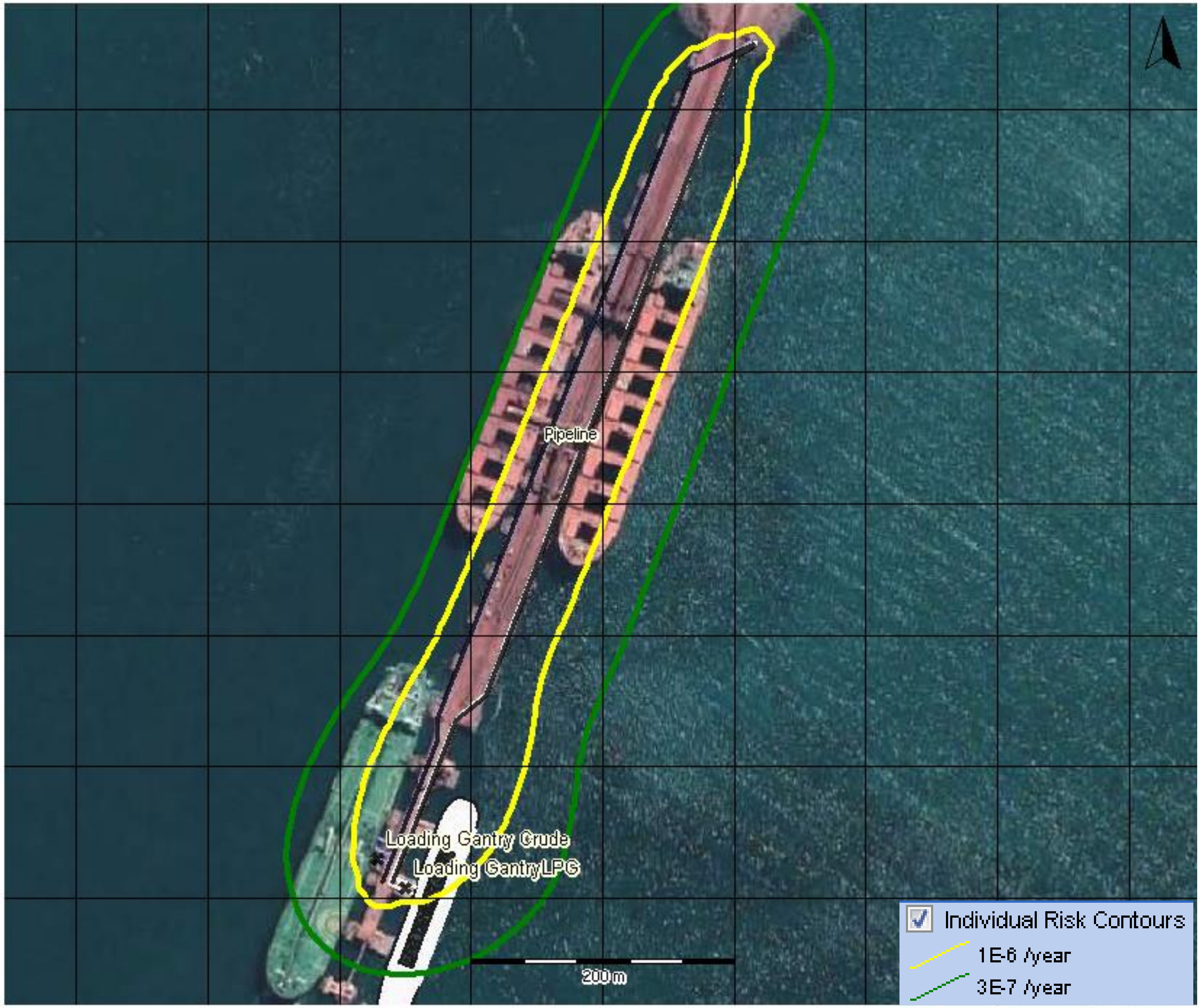
- The total individual risks involving the proposed crude oil and LPG pipelines are acceptable, with the 1.0e-6 (one-in-a-million) contour extending 46m beyond the. The 3.0e-7 (one-in-thirty-million) contour extends 98m beyond the pipelines and does not reach any sensitive areas or MHIs.

SFF Import Terminal Facility

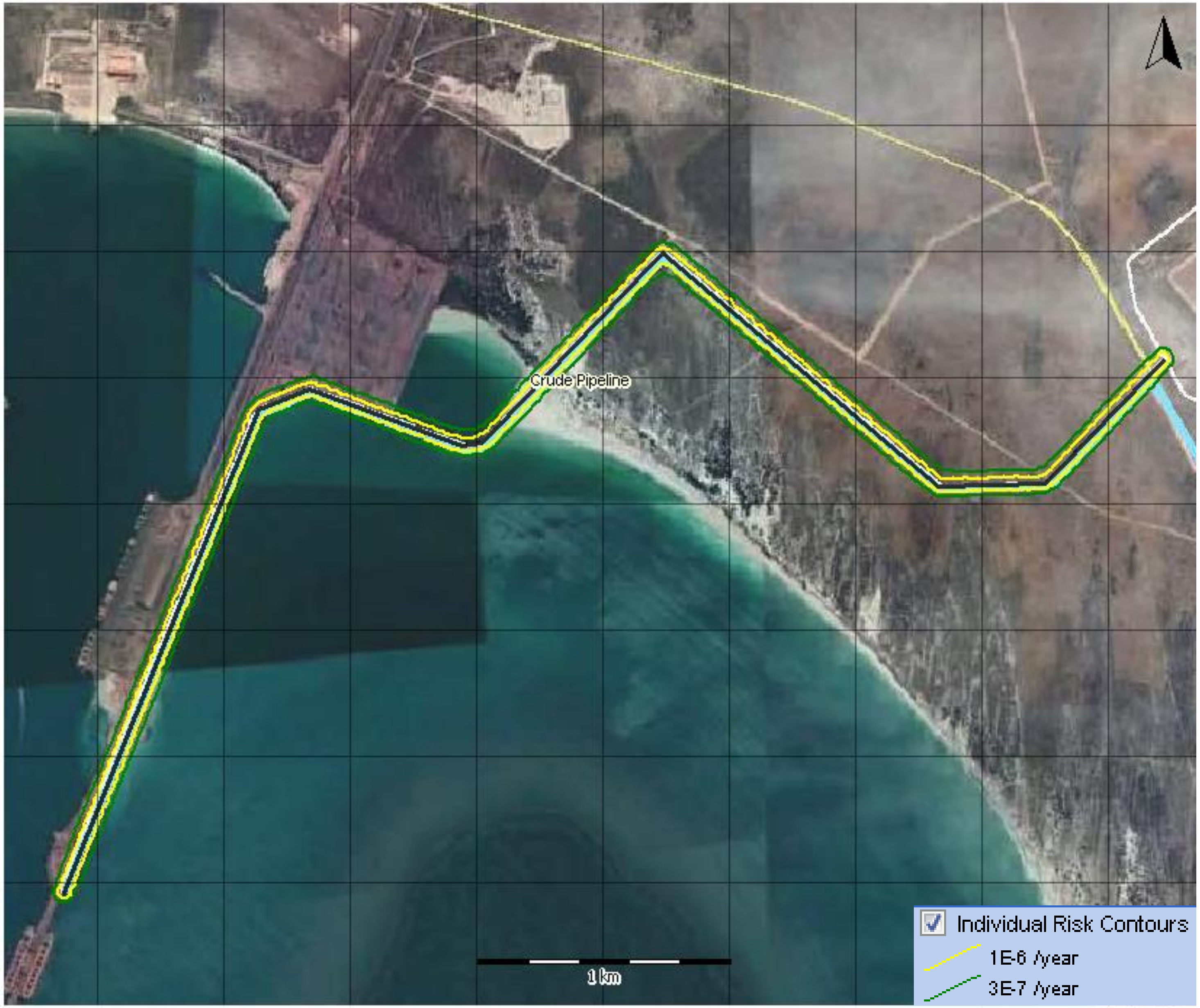
- The total individual risks involving the current crude oil installations are acceptable, with the 1.0e-5 (one-in-a-hundred thousand) reaching the edge of the bulk tanks but not reaching the property boundaries. The 1.0e-6 (one-in-a-million) contour extends beyond the property boundaries and just reaches the Moggs boundary to the east and across the main Saldanha/Langebaan road to the west. The 3.0e-7 (one-in-thirty-million) contour extends 570m beyond the boundaries, just reaching the Moggs tanks and does not reach any sensitive areas.
- The total individual risks involving the proposed LPG installations are acceptable, with the 1.0e-6 (one-in-a-million) contour extends 136m beyond the property boundaries but does not reach the Moggs boundary to the east and does not reach the main Saldanha/Langebaan road to the west. The 3.0e-7 (one-in-thirty-million) contour extends 706m beyond the boundaries, just reaching the Moggs tanks and does not reach any sensitive areas.
- The total individual risks involving the crude oil and LPG installations are acceptable, with the 1.0e-5 (one-in-a-hundred thousand) reaching the edge of the bulk tanks but not reaching the property boundaries. The 1.0e-6 (one-in-a-million) contour extends beyond the property boundaries and just reaches the Moggs boundary to the east and across the main Saldanha/Langebaan road to the west. The 3.0e-7 (one-in-thirty-million) contour extends 706m beyond the boundaries, just reaching the Moggs tanks and does not reach any sensitive areas.

Individual Risk Crude Oil Jetty

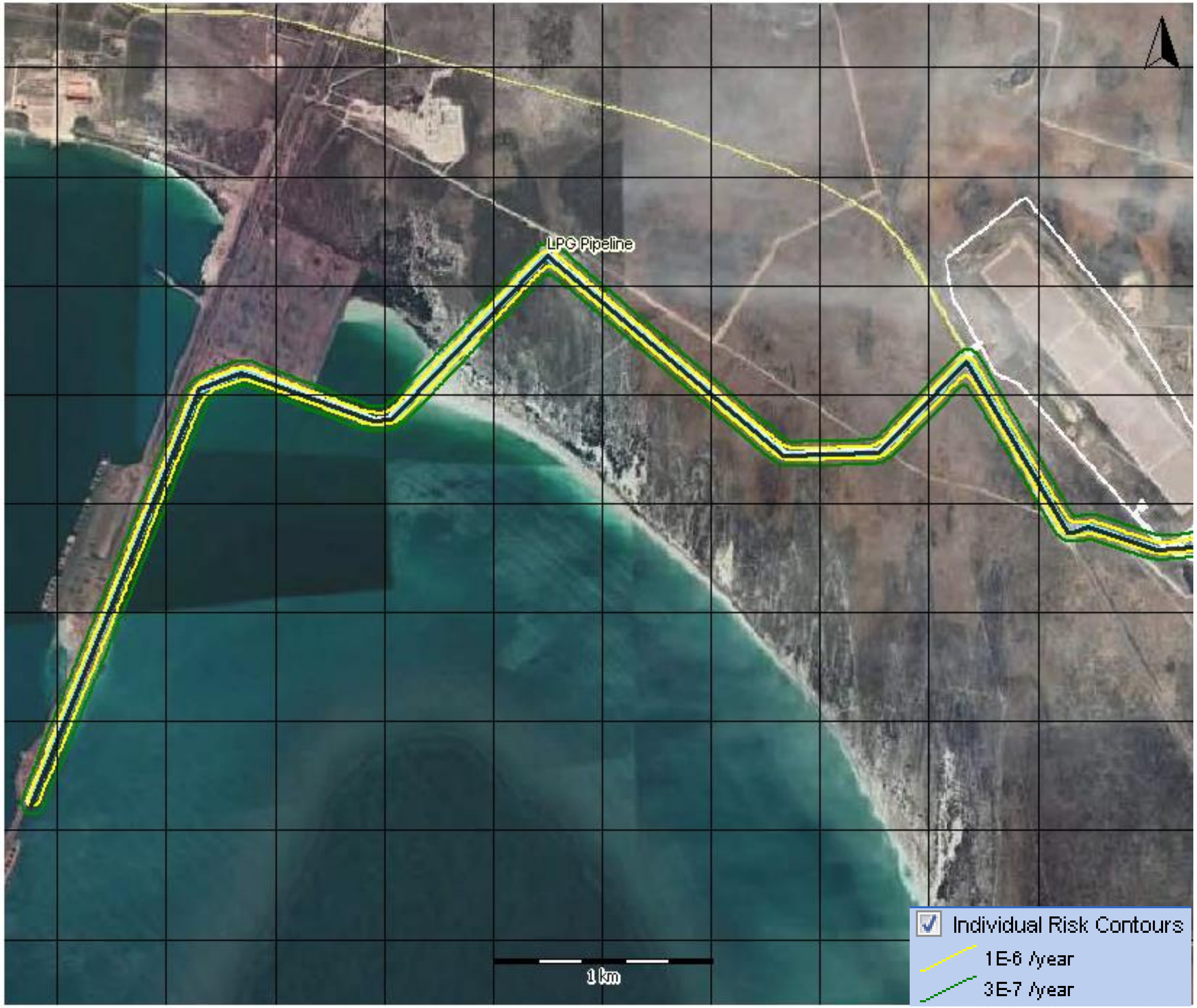
Individual Risk LPG Jetty



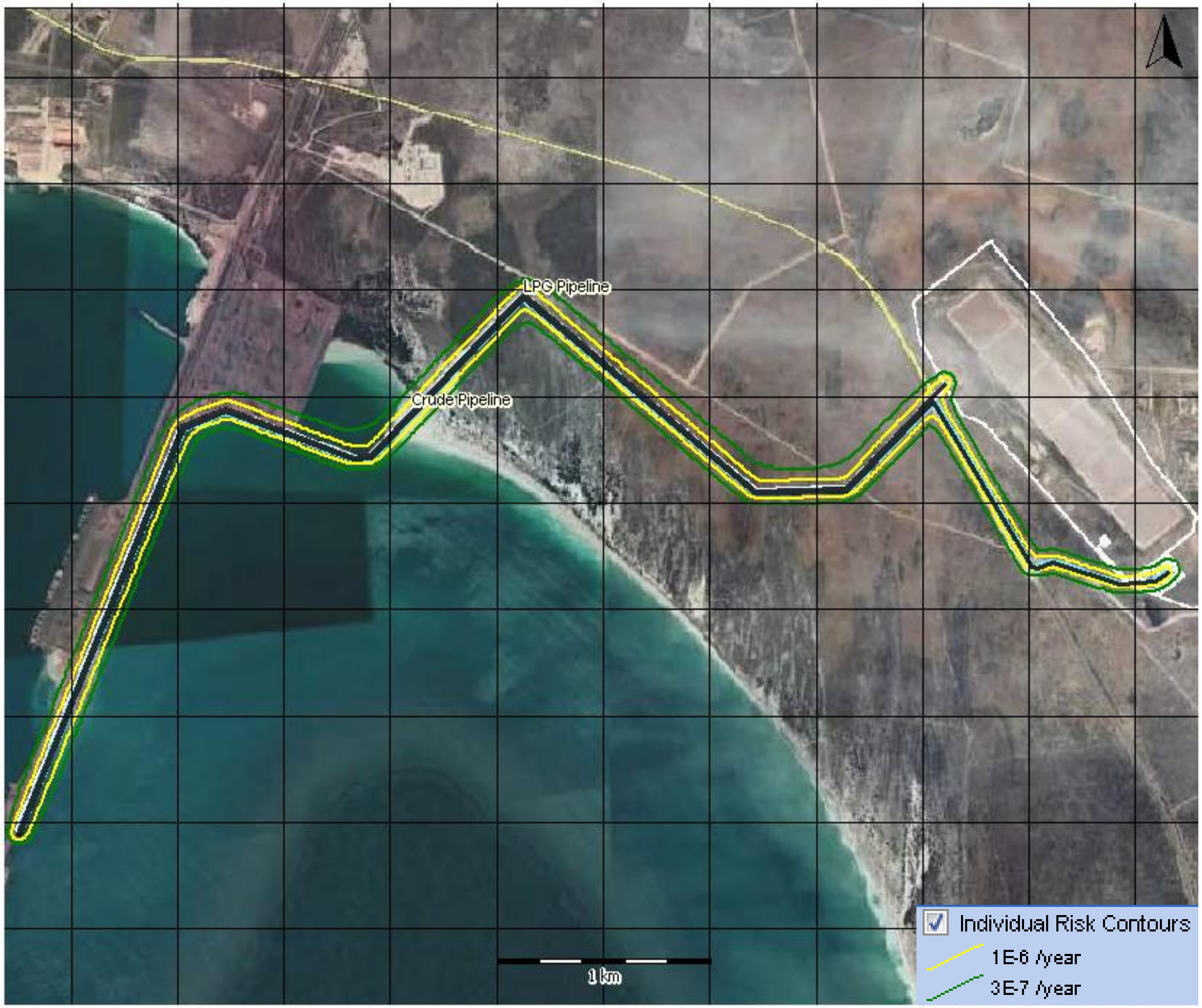
Individual Risk Crude Oil and LPG Jetty



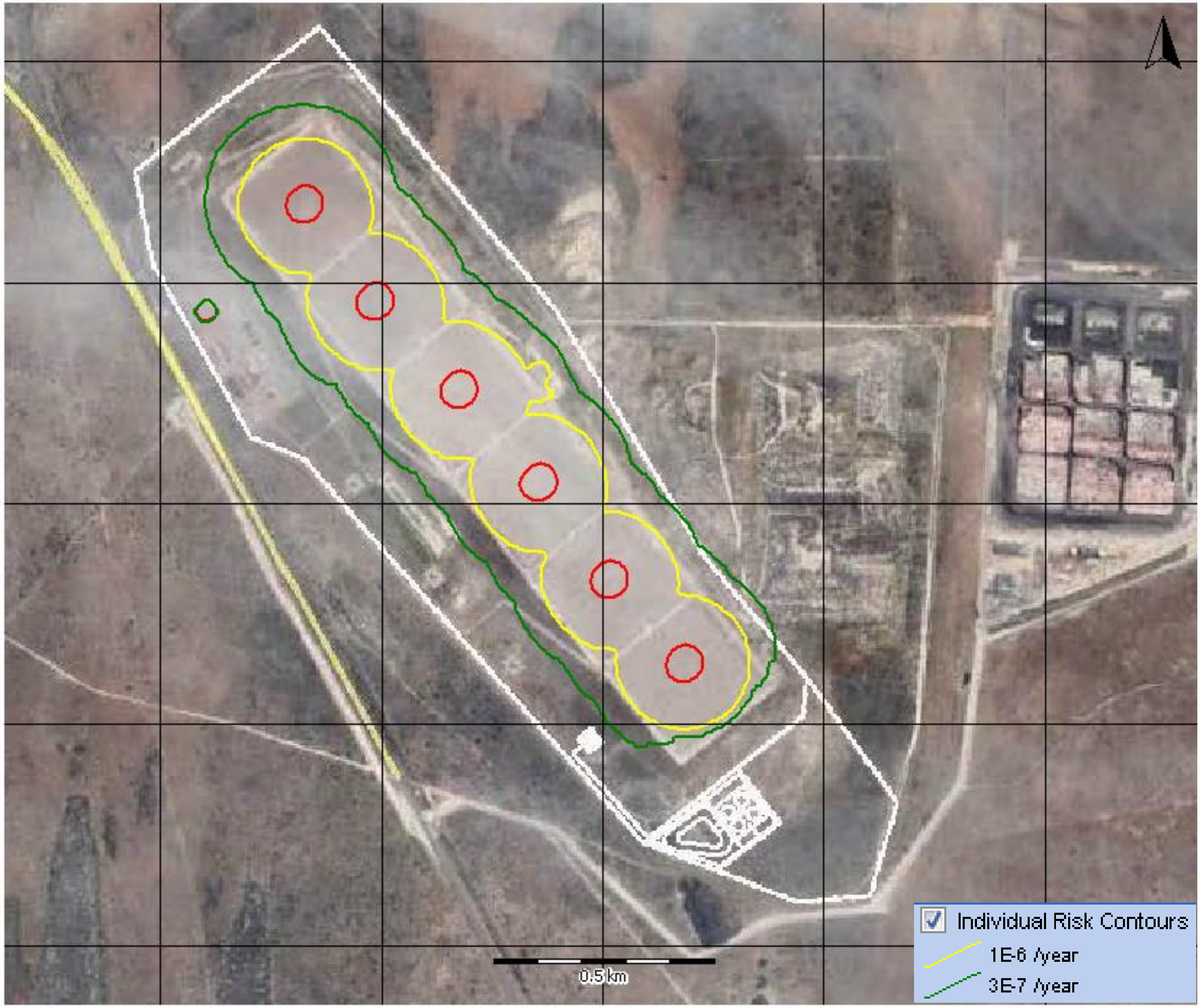
Individual Risk Crude Oil Pipeline



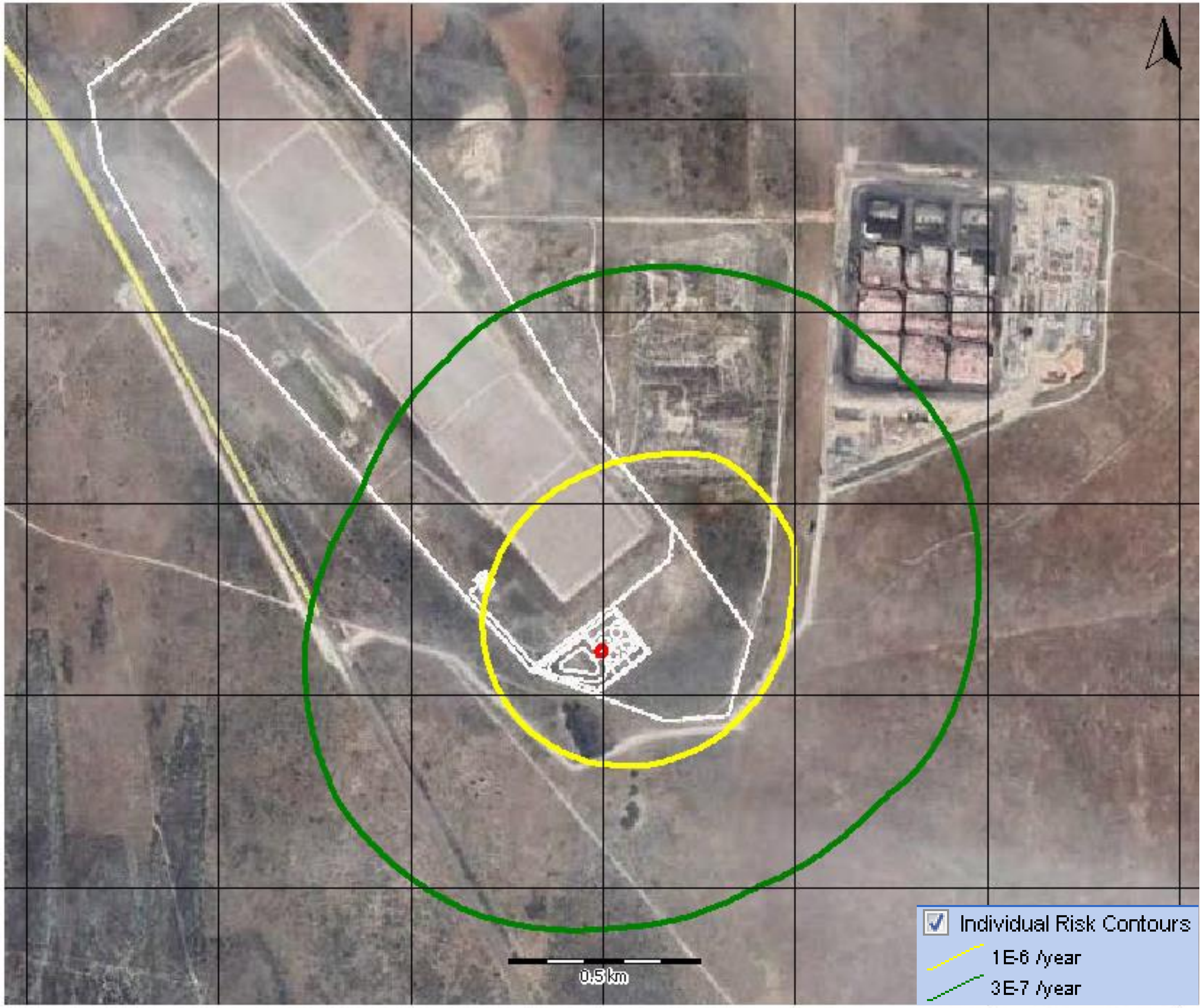
Individual Risk LPG Pipeline



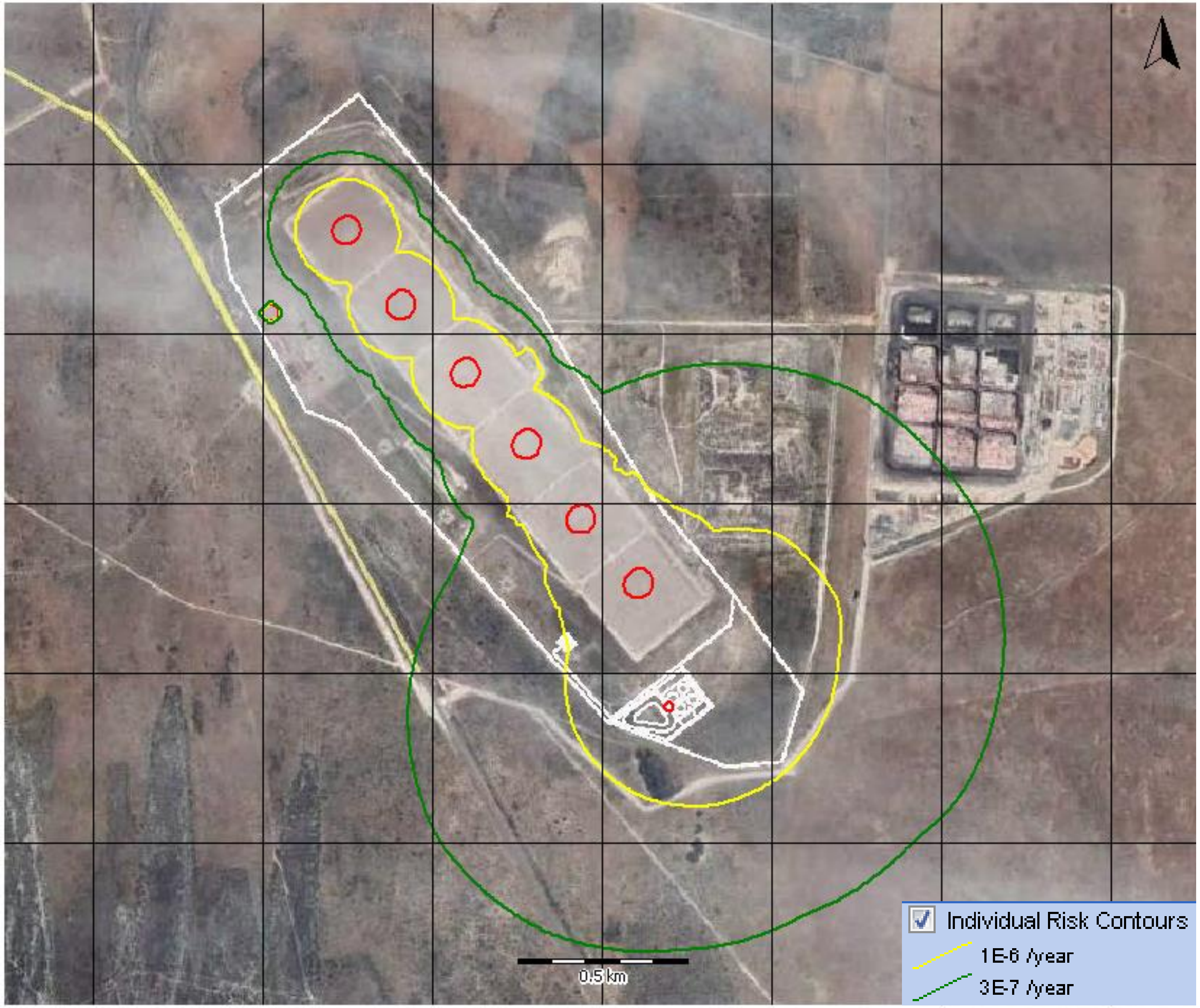
Individual Risk Crude Oil and LPG Pipeline



Individual Risk Crude Oil Installation



Individual Risk LPG Installation



Individual Risk Crude Oil and LPG Installation

3. Risk Levels and Ranking

Individual risk levels at several important points around the Jetty:

At the Ore Terminal

Scenario	Contribution %	Risk Value
LPG Pipe Rupture	64.2	1.13E-06
Crude Oil Pipe Leak	28.8	5.08E-07
Crude Oil Pipe Rupture	7	1.23E-07

Individual risk levels at several important points around the Pipelines:

25m Away from the Pipelines

Scenario	Contribution %	Risk Value
LPG Pipe Rupture	44.7	8.48E-07
Crude Pipe Rupture	38.3	7.46E-07
Crude Pipe Leak	16	3.03E-07

50m Away from the Pipelines

Scenario	Contribution %	Risk Value
LPG Pipe Rupture	40.2	1.54E-07
Crude Pipe Rupture	39.7	1.52E-07
Crude Pipe Leak	20.1	7.70E-08

Individual risk levels at several important points around the Import Terminal:

At Moggs Property

Scenario	Contribution %	Risk Value
Emptying in 10 Minutes Sphere 2	18.5	1.15E-07
Emptying in 10 Minutes Sphere 1	18.2	1.13E-07
Emptying in 10 Minutes Sphere 3	15.1	9.37E-08
Emptying in 10 Minutes Sphere 4	15	9.32E-08
Vessel Failure Sphere 4	8.36	5.18E-08
Vessel Failure Sphere 3	8.34	5.16E-08
Vessel Failure Sphere 2	8.21	5.08E-08
Vessel Failure Sphere 1	8.19	5.07E-08

At SFF Offices

Scenario	Contribution %	Risk Value
Vessel Failure Sphere 3	15.9	8.56E-09
Vessel Failure Sphere 1	15.5	8.36E-09
Vessel Failure Sphere 4	13.8	7.44E-09
Vessel Failure Sphere 2	13	7.03E-09
Emptying in 10 Minutes Sphere 3	10.8	5.82E-09
Emptying in 10 Minutes Sphere 1	10.6	5.74E-09
Emptying in 10 Minutes Sphere 4	10.3	5.57E-09
Emptying in 10 Minutes Sphere 2	10.1	5.47E-09

At Main Road

Scenario	Contribution %	Risk Value
Emptying in 10 Minutes Sphere 3	17.5	8.02E-08
Emptying in 10 Minutes Sphere 4	17.1	7.84E-08
Emptying in 10 Minutes Sphere 1	16.3	7.49E-08
Emptying in 10 Minutes Sphere 2	15.8	7.22E-08
Vessel Failure Sphere 4	8.58	3.93E-08
Vessel Failure Sphere 3	8.55	3.92E-08
Vessel Failure Sphere 1	8.36	3.83E-08
Vessel Failure Sphere 2	7.82	3.58E-08

4. RECOMMENDATIONS

The main risk contributing parts of the total risk at the facility is the bulk crude oil installations.

The risks were found to be acceptable for the remote area in which the installations are located.

The recommendations are as follows:

- Good housekeeping always needs to be observed on site;
- The Emergency Plan must be updated to include the scenarios described in this report;
- The Emergency Plan must comply with SANS 1514;
- Once the designs for the LPG Import Terminal has been finalised a final MHI must be conducted and the updated MHI report must be distributed to Local, Provincial and National Government as per MHI Regulations;
- Only suitably qualified people must be used for all work on the proposed installations;
- Check alarms and emergency procedures regularly;
- Do a full pressure test to ensure that there are no leaks prior to commissioning the installations;
- Hazard Area Classification must be done as per SANS 10108;

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15.4. Design and Drawing for the proposed LPG Installations
15.5. Frequency Analyses
15.6. HSE Development Sensitivity Tables

DEFINITIONS

As Low as Reasonably Practicable (ALARP)

Risks in this range are risks that the public are generally prepared to tolerate to secure certain benefits. A risk in the ALARP range risk means that for new installations or modifications/ expansions to existing installations, the risk assessment shall not advise against the development. For existing installations (without modifications/ expansions) a broadly acceptable risk means that risk should continue to be monitored and all reasonably practicable risk reduction measures shall be implemented. A level of risk that is tolerable and cannot be reduced further without expenditure at costs that are disproportionate to the benefit gained, or where the solution is impractical to implement.

Broadly Acceptable

Risks which are broadly acceptable are generally regarded as insignificant and adequately controlled. Risk in the region would usually not require further action to reduce risks unless reasonably practicable measures are available. A broadly acceptable risk means that for new installations or modifications/ expansions to existing installations the risk assessment shall not advise against the development. For existing installations (without modifications/ expansions) a broadly acceptable risk means that risk should continue to be monitored and reduction implemented if necessary. For either new or existing installations, if reasonably practicable risk reduction measures are available, then these should be implemented.

BLEVE

Boiling liquid expanding vapour explosion.

Containment System

One or several devices, any parts of which are continuously in open contact with one another and are intended to contain one or several substances.

Critical Scenarios

Intended to mean:

- The scenarios that when added together define at least 90% of the location-specific risk for the 1.0e-6 contour (i.e. the ‘remainder’ that has not been defined in detail is added together as < 10%);
- The scenarios that are added together define at least 90% of the societal risk in the intervals 10 – 100 and 100 – 1000.

Informal Residential Area

A residential area where the structures are not formally approved.

Inspection

An examination or measurement to verify whether an item or activity conforms to specified requirements.

Intolerable

Risks in this range are generally regarded as unacceptable whatever the level of benefits associated with the activity. An intolerable risk means that for new installations or modifications/ expansions to existing installations the risk assessment shall advise against the development. For existing installations (without modifications/ expansions) an intolerable risk means that risk reduction shall be implemented until the risks fall within the ALARP range or the broadly acceptable range.

Location Specific Individual Risk

The probability that during a period of one year a person will become the victim of an accident, in which case this person is in a particular location permanently and without protection and without means of escape.

Major Hazard Installation

The Operational Health and Safety Act defines a Major Hazard Installation as the following:

- where more than the prescribed quantity of any substance is or may be kept, whether permanently or temporarily; or
- where any substance is produced, used, handled or stored in such a form and quantity that it has the potential to cause a major incident.

Maximum Capacity

For equipment this is the total amount of material that can be accommodated in that equipment in the absence of equipment inventory control. For example, the volume of a cube vessel would be the product of the width, length and height of the vessel.

Occupied Building

Permanent or temporary structures/ buildings within a major hazard installation that are occupied by employees and/or contractors or that contain critical process control equipment (e.g. control rooms).

Procedure

Description of how to perform an activity, usually in the form of a document.

Recommendations

Suggestions put forward by the AIA, within the scope of the accreditation of the AIA, for consideration by the owner/ user of an MHE/ MHI.

Regulations

Regulations promulgated under the relevant Act.

Regulatory Authority

Body authorised to make Regulations or to control the application of such Regulations, in the field of Major Hazard Installations (see 3.1.22) which includes the Occupational Health and Safety Act, 1993 (Act 85 of 1993) and the South African National Accreditation System.

Restricted Development Distance

The maximum distance from an MHI/ MHE where land use planning restrictions should be considered. This is defined as the 3.0e-7 fatalities / person / year location specific individual risk contour.

Safety Report

A report which addresses major incident prevention and safety management systems at the installation/ establishments.

Sensitivity Level

The sensitivity levels of a proposed development take into consideration the structure of the development and the characteristics of the population occupying the development. The larger the development and the more vulnerable the occupying population, the higher the level of sensitivity.

Societal Risk (F-N Curve)

Societal risk is a measure of the risk posed on a society and an F-N Curve is a tool to indicate societal risk. They are plots of the cumulative frequency (F) of various accident scenarios against the number (N) of fatalities associated with the modelled incidents. The plot is cumulative in the sense that, for each frequency, N is the number of fatalities that could be equalled or exceeded.

Verification

The act of reviewing, inspecting, testing, checking, auditing or otherwise determining and documenting whether items, processes, services or documents conform to specified requirements.

Vulnerable Groups/ Populations

The elderly, children, persons in hospitals/ clinics and people with certain disabilities are considered particularly vulnerable and may need special attention. In the South African context, concentrations of homeless persons and persons occupying informal settlements should also be considered vulnerable.

ABBREVIATIONS

The following are key abbreviations used in this document:

ACDS	Advisory Committee on Dangerous Substances
AIA	Approved Inspection Authority
ALARP	As Low As Reasonably Practicable
API	American Petroleum Institute
BEVI	Besluit Externe Veiligheid Inrichtingen (Dutch safety legislation)
BLEVE	Boiling Liquid Expanding Vapour Explosion
BP	Boiling Point (usually at 101.325 kPa)
CAS	Chemical Abstracts Service
CASRN	Chemical Abstracts Service Registry Number
RDD	Restricted Development Distance
CFD	Computational Fluid Dynamics
CIA	Chemical Industries Association
DTL	Dangerous Toxic Load
ERPG	Emergency Response Planning Guideline
F – N (cumulative)	Frequency - Number
FMECA	Failure Mode Effect and Criticality Analysis
FP	Flash Point

HAZID	HAZard IDentification
HAZAN	HAZard ANalysis
HEL	Higher Explosive Limits
IBC	Intermediate Bulk Container (typically $1m^3$ capacity)
IDLH	Immediately Dangerous to Life and Health
IEC	International Electro-technical Commission
ISO	International Standards Organisation
IZ	Inner Zone
kPa	Kilo-Pascal
kW/m2	Kilo-watts Per Square Meter
L/D	Length/ Diameter
LEL	Lower Explosive Limits
LFL	Lower Flammable Limit
LOC	Loss Of Containment
LOPA	Layer Of Protection Analysis
LPG	Liquefied Petroleum Gas
MAHPs	Major Accident Hazard Pipelines
MAPP	Major Accident Prevention Policy
mg/m^3	Milligram Per Cubic Meter
MHI	Major Hazard Installation
MZ	Middle Zone
OHS	Occupational Health and Safety
OZ	Outer Zone
PAC	Protective Action Criteria
PAHDI	Planning Advice for Developments near Hazardous Installations
PFD	Process Flow Diagram
P&ID	Piping and Instrumentation Diagram
ppm	Parts-per-million (volume basis)
PSM	Process Safety Management
QRA	Quantitative Risk Assessment
UFL	Upper Flammable Limit

A QUANTITATIVE RISK ASSESSMENT OF THE TOXIC AND FLAMMABLE INSTALLATIONS ON THE PREMISES OF SFF IN SALDANHA AND THE PORT OF SALDANHA

1. INTRODUCTION

The Strategic Fuel Fund Association (SFF) was established in 1964 as a Section 21 company. The first crude oil it ever owned was stored in its Milnerton tanks in 1967. In 1975, a levy on fuel payable into the Central Energy Fund was introduced and managed by CEF (Pty) Ltd and a portion of this was paid to SFF.

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In 1979, the world suffered the second major international oil crisis. That was the year when the procurement of crude oil by SFF on behalf of the oil companies began and when the first Saldanha oil tank was completed. The whole Saldanha project was completed in 1982.

SFF's oil storage installation at Saldanha is the largest facility of its kind in the world. It comprises six in-ground concrete storage tanks which have a combined capacity of 45 million barrels of oil. The tanks are linked by pipeline to an oil terminal at the Port of Saldanha where tankers either discharge or backload oil. A second 107 km long pipeline, belonging to the refinery in Milnerton links the oil storage facilities with the refinery. Crude oil is also stored for other companies and delivered to ships via the pipeline to the port.

LPG has been declared a strategic fuel and SFF is proposing to build a bulk LPG import storage terminal on their premises linked by a pipeline to a proposed LPG import terminal at the Port of Saldanha.

As crude oil and LPG have the potential to cause onsite and offsite incidents, Major Hazard Risk Consultants cc was commissioned to conduct a Risk Assessment in accordance with the Major Hazard Installation Regulations to determine the impact of the facilities on the surrounding area.

This investigation would serve as a basis for the notification of the facility in accordance with the Major Hazard Installation Regulations. The purpose of this report is to convey the essential details, including a short description of the hazards, the receiving environment, the design, the risks and consequences of an accident.

The main aim of the investigation was to quantify the risks to employees and neighbours regarding the facility in Saldanha Bay.

Risk is the severity of the consequence of a hazardous event and the probability of the event occurring.

This report summarises the results of the Risk Assessment conducted by MHR Consultants.

This Assessment is based on the best possible information and expertise and MHR Consultants cannot be held liable for any incident which may occur at this facility which directly or indirectly relates to the work in this report.

4.6. Legal Framework

The Occupational Health and Safety Act (OHS Act) defines an Approved Inspection Authority (AIA) in Section 1(1)(i) as *“An inspection authority approved by the Chief*

Inspector: Provided that an inspection authority approved by the Chief Inspector with respect to any particular service shall be an approved inspection authority with respect to that service only."

The Major Hazard Installation Regulations (MHI Regulations), which were promulgated under the OHS Act provides more specifically for an AIA in terms of MHI Regulation 5 (5)(a) as *"An employer, self-employed person and a user shall ensure that the assessment contemplated in Sub-regulation (1), shall be carried out by an Approved Inspection Authority which is competent to express an opinion as to the risks associated with the major hazard installation."*

This Risk Assessment was conducted as per SANS 1461:2018.

The report also includes the Regulations according to the local by-laws.

The report is intended as a Specialist Report for the Environmental Impact Assessment.

4.7. Purpose and Scope of Investigation

The purpose of this investigation was to quantify the risks to employees and neighbours regarding the facility in Saldanha Bay.

This Risk Assessment was conducted in accordance with the Major Hazard Installation Regulations and is intended as a Specialist Report for the Environmental Impact Assessment. The Risk Assessment includes the following:

- Identifying likely hazards associated with the processes of the installations including the causes, consequences and their effects;
- Quantifying the likely hazards in terms of their magnitude;
- Quantifying the consequences for each hazard (thermal radiation, domino effect, toxic cloud formation, etc.);
- Determining the lethality of the effects of the consequences;
- Determining the frequency of all the hazardous events;
- Calculating the individual risk values considering all accidents, meteorological conditions and lethality;
- Using the population density around the facility to determine the societal risk posed by the facility;
- Reporting on the risks in terms of internationally acceptable criteria;
- Providing an assessment of the adequacy of emergency response programmes, fire prevention and fire-fighting measures;
- Proposing measures to reduce or eliminate the risks.

4.8. Methodologies

Methodologies and techniques used for this Assessment are as follows.

- Site visits and meetings were conducted to collect as much technical information to accurately determine all the processes, materials, etc.;

- It was accepted that the process and storage installations were designed using the correct codes of practice and design specifications, and that the installations were built by qualified professionals;
- For this report the public refers to all people outside the boundaries of the facility, including neighbouring facilities and everyone inside the facility is regarded as employees, including visitors;
- The hazards were identified at the site visits and meetings and analysed using international reference publications;
- The consequences were calculated using the computer software *'Effects'* by TNO in the Netherlands;
- The risk calculations were made using the computer software *'Risk Curves'* by TNO in the Netherlands.

5. COMPANY, SITE AND INSTALLATION DESCRIPTION

5.6. Main Activity

The main activity of SFF is the storage of crude oil and LPG.

5.7. Site Address

The sites are located in located at the Port of Saldanha and the SFF import facility at:

Extension of Camp Road
Saldanha Bay.

5.8. Site Installations

The site at the Port of Saldanha consists of the following:

- Existing crude oil loading/off-loading jetty just beyond the ore loading facility
- Three loading arms for connecting to the crude tanker ships
- A 700m above ground crude oil 1067 diameter steel pipeline on the jetty to the underground transfer station.
- A 700m above ground water 508 diameter steel pipeline on the jetty to the underground transfer station.
- A Pipeline End Manifold (PLEM) for a 300mm steel pipeline with flexible hoses for connecting to an LPG tanker ship.
- A 700m above ground 300mm steel LPG pipeline from the PLEM to the underground transfer station.

The 10m wide pipeline servitude site consists of the following:

- Existing underground 10067mm crude oil pipeline from the jetty at the Port of Saldanha to the SFF import facility.
- Existing underground 508mm water pipeline from the jetty at the Port of Saldanha to the SFF import facility.
- Existing Pig Traps for the crude oil pipeline.
- Proposed underground 300mm LPG pipeline from the jetty at the Port of Saldanha to the SFF import facility.

The site at the SFF facility consists of the following:

- Existing Offices;
- Existing Bulk Nitrogen Vessels;
- Existing Bulk Crude Oil Caverns;
- Existing Booster Pump Station;
- Existing Pig Traps;
- Existing Bulk LPG Vessel for the Flare;
- Existing Tank Flare Installation;
- Existing Fire Water Pump Station;
- Existing Bulk Fire Water Tanks;
- Proposed Bulk LPG Spheres;
- Proposed LPG Pump Station;
- Proposed LPG Road Tanker Gantry;

5.9. Process Flow

Crude Oil Process:

The crude oil is pumped from a tanker ship at the jetty in the Port of Saldanha to the SFF import facility via a 10067mm pipeline into bulk crude oil tanks. The crude oil can then be pumped via pipeline to Milnerton or back to the Port of Saldanha to a tanker ship. See diagram below.

Crude Oil Process Flow

LPG Process:

The LPG is pumped from a tanker ship at the jetty in the Port of Saldanha to the SFF import facility via a 300mm pipeline into bulk LPG spheres. The LPG can then be pumped via or back to the Port of Saldanha to a tanker ship. See diagram below.

LPG Process Flow

5.10. Installations at the Port of Saldanha

5.10.4. Crude Oil Jetty

The existing loading/offloading installation on the Saldanha side of the jetty consists of three loading arms and a pipe and pump manifold in a bunded area on the jetty. There are two pumps connected to the pipe manifold onto the two pipelines. The section of pipelines on the jetty is above ground. The pipelines are aboveground until just before the Multi-Purpose quay at the underground transfer station. The ships pumps will be used to pump the crude oil to the tank farm and the tank farm pumps will be used to pump crude oil to the ship.

The maximum design flow rate is 12 000 m³/hour from jetty to the tank farm using the ships pumps. The average flow rate is 8 800 m³ / hour from jetty to tank farm using the ships pumps. The maximum loading flow rate from tank farm to the ship using the tank farm pumps is 8 000 ³m / hour and the average flow rate using the tank farm pumps is 7 800 m³ / hour.

5.10.5. LPG Jetty

The proposed loading/offloading installation on the Langebaan side of the jetty will consist of Pipeline End Manifold (PLEM) on the jetty. The manifold will be connected to the 300mm pipeline. The section of pipeline on the jetty will be above ground. The pipeline will be aboveground until just before the Multi-Purpose quay at the underground transfer station. The ships pumps will be used to pump the crude oil to the tank farm and the tank farm pumps will be used to pump crude oil to the ship.

The maximum design flow rate is 750 m³/hour from jetty to the tank farm using the ships pumps. The average flow rate is 500 m³ / hour from jetty to tank farm using the ships pumps. The average flow rate using the tank farm pumps is 400 m³ / hour.

During import operations, the vessel lined up to receive product will be isolated from all the liquid and vapour connections. The 12” liquid import line will then be opened to enable the receipt of product. During operations, the 12” liquid import line will be filled with product and liquid will be pumped via the cargo carrier pumps to the storage vessel.

After the receiving vessel has been filled to a certain level, the cargo carrier pumps will be stopped. The liquid line will be emptied by means of a vapour push. The cargo carrier will utilise on board compressors to achieve the vapour push. The current SFF facility is located roughly 10km from the jetty and the 12” liquid import /export line will contain an approximate volume of 820 cubic meters of product. Considering that each storage vessel will have a volume of 8181 cubic meters, the tank filling during liquid pumping will be limited to ensure that sufficient storage space is available for the receipt of the liquid in the import line.

The line will then be emptied by means of the vapour push. Once the vapour push is complete, and a custody transfer surveyor has confirmed quantity and quality of product, the cargo carrier may disconnect.

During export operations, both vessel pumps can be utilised to increase transfer rates. A maximum of 400m3/hr can be achieved with the use of both pumps. The vessel under export mode will be lined up such that the liquid export connection is opened to the suction header. The liquid export pumps will draw product off the suction header and discharge product into the export line via the 12” import/export line.

The vapour push required to remove liquid from the import/export line will be performed by cargo carrier vapour compressors and the liquid remaining in the import/export line will be transferred back to storage.

There will be a Control Station is provided on the Jetty to provide the following functions:

- Operate electrical isolating flanges;
- Operate Fire Safe, Fail Safe, Nitrogen (or Instrument Air) Operated, Flow Control Valves;
- Monitor Coriolis Flow Meter providing flow, temperature, and density monitoring;
- Monitor Pressure Transmitter;

The Control Station will contain a segregated Emergency Shutdown System in a separate cubicle to take care of the following functions:

- Monitor Fire detection / CCTV combination monitoring at berth;
- hut off the LPG flow in the event of fire (Flame detector);
- Shut off the LPG flow on detection of LPG at the Gas Detector;
- Shut off the LPG flow when the manual ESD button located at the operations is activated;
- Shut off the LGP flow on command from the PCMS;
- All shut-downs are reported to the PCMS.

Jetty Installations

5.11. Pipelines from the port of Saldanha to the SFF Import Terminal

5.11.4. Crude Oil Pipelines

There are overland two pipe lines located in a10m servitude from a causeway in the Port of Saldanha to the bulk crude oil storage facility. The main pipeline is a 1067mm diameter pipeline and the other pipeline is 508mm for seawater.

Pipe Outside Diameter	1067mm	508mm
Material	Steel	Steel
Wall Thickness	14mm	9.5mm

The pipeline starts at the jetty with three loading arms and ends at the tank farm in a bunded pipe manifold and two pumps. The product can be directed to any of the bulk oil tanks.

Corrosion protection consists of three coats of N-oxide paint and bitumen wrapping as well as cathodic corrosion protection. The pipeline depth is between 1 and 2 meters.

The pipelines are fitted with cathodic corrosion protection test points, pressure sensor manholes and air vent manholes.

The pipeline is designed with pigging operations for maintenance and statutory inspection (See pig launching and receiving procedures in Appendix 1).

The maximum design flow rate is 12 000 m^3/hour from jetty to the tank farm using the ships pumps. The average flow rate is 8 800 m^3 / hour from jetty to tank farm using the ships pumps. The maximum loading flow rate using the tank farm pumps is 8 000 ^3m / hour and the average flow rate using the tank farm pumps is 7 800 m^3 / hour.

The pipeline has an automation system that is a custody-metering system at the tank farm where the pipeline is constantly monitored and it includes various levels of alarms and blocking systems. Flow rates and pipeline pressures are constantly monitored. The pipeline is cleaned and inspected for condition and stresses by regular 'smart pigging'. The pipeline corridor is marked and maintained.

It is assumed that the pipeline has been designed to reduce the likelihood an incident to as low as reasonably possible. The primary code for the pipeline is assumed to be ASME B31.4 Pipeline Transportation Systems for Liquid Hydrocarbons and Other Liquids. The code applies to piping transporting products which are predominately liquid between wells, plants, and terminals, and within terminals, pumping, regulating, and metering stations. B31.4 is intended to be applied to pipeline transportation of liquids such as crude oil.

The pipeline route is shown on the satellite image below followed by the pipeline design diagrams.

5.11.5. LPG Pipeline

The risk assessment was based on the following proposed pipeline design:

• Type of pipeline	Carbon Steel ASTM A-333 Grade 6
• Hydrostatic Test Pressure:	30 bar(g) run over approx. 4 hours[3]
• Nominal pipe diameter Maximum Allowable Operating Pressure	12" (305mm) 20 bar(g)
• Wall thickness	12 mm
• Pipe coating	3-Layer Polyethylene or Polypropylene: FBE: min 350µm Polyethylene or Polypropylene sheath: min 2.5 mm
• Pipe joints	Clean weld, apply primer, apply bitumen wrap/ denso-tape

It is assumed that the design code will be ASME B31.4 *Pipeline Transportation Systems for Liquid Hydrocarbons and Other Liquids*. The code applies to piping transporting products which are predominately liquid between wells, plants, and terminals, and within terminals, pumping, regulating, and metering stations.

Crude and LPG Pipeline Routes

5.12. SFF Crude Oil Import Terminal

5.12.4. Bulk Crude Oil Tank Farm

The tank farm consists of six square, semi-underground concrete tanks with a capacity of one million m3 each, as shown in the picture below. The total storage capacity of the storage facility is 45 million barrels (six million m^3).

Tank Farm

These concrete tanks are retained by earthen embankments, which are planted with indigenous vegetation. No bunds are required for secondary containment.

Each tank has a fixed concrete roof supported with pillars with a conservation vent that allows vapours to be removed from the tank during filling. The vent also keeps the tank's vapour pressure constant with the inclusion of nitrogen injection. This prevents the formation of flammable vapours above the liquid level, reducing fire and explosion hazards.

During filling operations, the hydrocarbon gasses will be directed to a flare situated on the north-western side of the tanks to be burned off.

A fire-water pump station, incorporating foam, is located at each tank in the event of a fire.

There is an oily-water treatment system in place, i.e. an oil-water separator and evaporation ponds for the treatment of water drained from the storage tanks. The evaporation ponds are water proof lined lagoons to prevent any possible soil and groundwater contamination. (see layout details on the satellite image below).

Tank Farm Details

5.12.5. Booster Pump Station

There are two main canned booster pumps rated at 8800m^3 per hour located in a bunded area with the pipe manifolds with manual and solenoid valves and controls for pumping crude oil. There are backup diesel pumps in a pump house.

These pumps are used to pump crude oil to ships in the harbour as well as pumping crude oil to Milnerton via the 107km pipeline.

5.12.6. Pig Traps

A pig is a device inserted into a pipeline which travels freely through the pipeline, driven by the product flow, to do a specific task within the pipeline. These tasks fall into a number of different categories:

- Utility pigs perform a function such as cleaning, separating products in-line or dewatering the line.
- Inline inspection pigs are used to provide information on the condition of the pipeline and the extent and location of any problem.
- Special-duty pigs act as plugs for isolating pipelines.

Pigs are inserted into the pipeline by pig launchers and are removed at pig receivers, these combined are referred to as pig traps.

There is a pig trap for the pipelines at the jetty as well as at the bulk tank farm.

5.12.7. Fire Pump Station

Next to the booster pump station is the fire pump station consisting of pumps and pipe manifolds for the control of fire water to the different tank foam pump stations. There are water reservoirs and also a test water return pump connected to the smaller pipeline.

5.12.8. Nitrogen Storage and Process

The bulk crude oil tanks use nitrogen as a blanket to prevent air ingression into the tanks and thereby preventing an explosive atmosphere within the tanks.

Liquid nitrogen is delivered by road and stored in cryogenic tanks having a total capacity of 550 m^3 in 5 x 50 m^3 vessels.

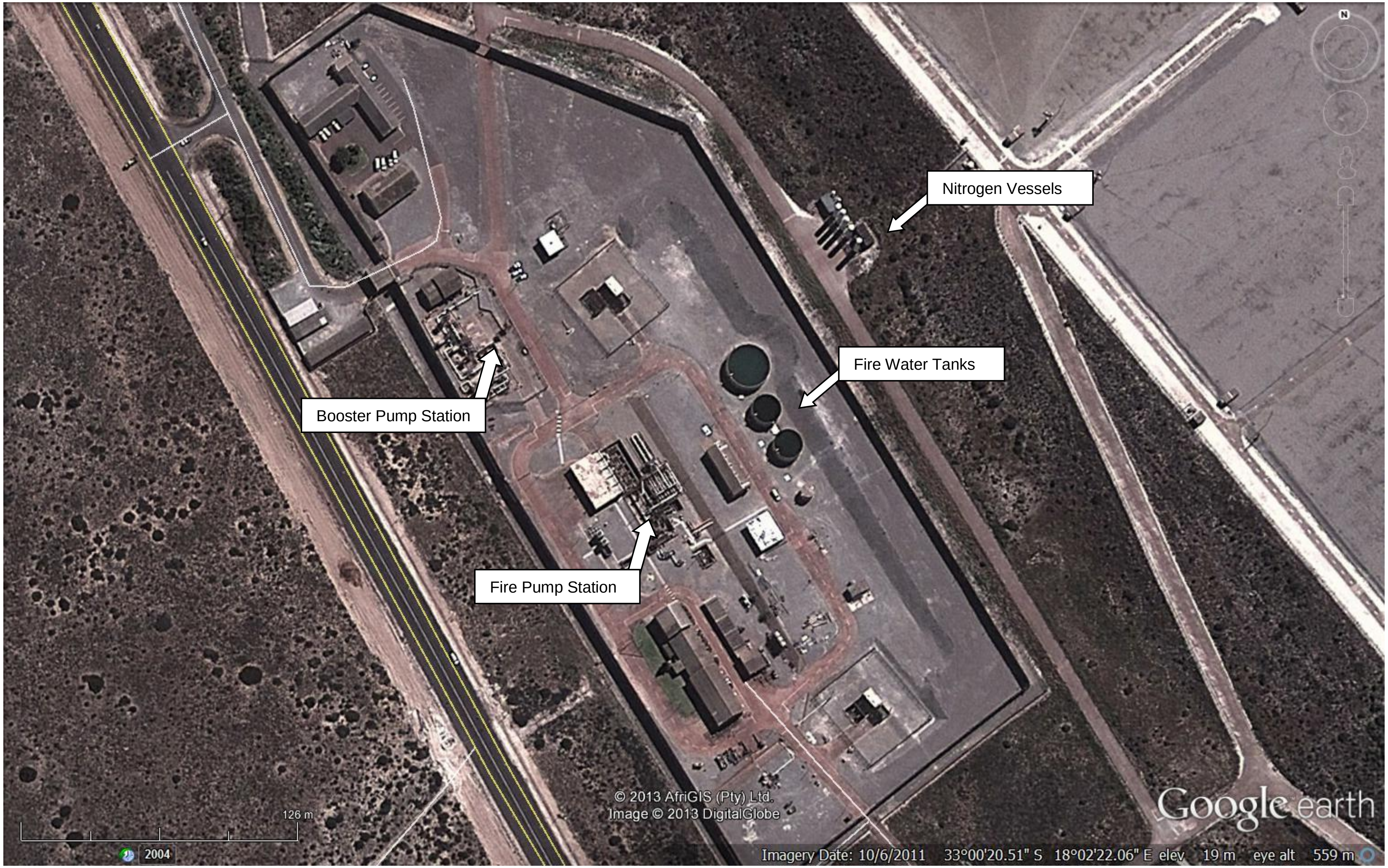
As no off-site fatalities would be expected from asphyxiation from a nitrogen release, no further analysis was required.

5.12.9. Tank Flare Installation

For the flare, LPG is used to fuel the pilot flame. At the flare installation there is a 9m^3 LPG vessel. The vessel is filled from a road tanker.

During filling operations, approximately 8 800 m3/hour of crude-oil vapours mixed with nitrogen is displaced from the bulk storage tanks. These vapours are sent to a flare located at the north eastern side of the tanks.

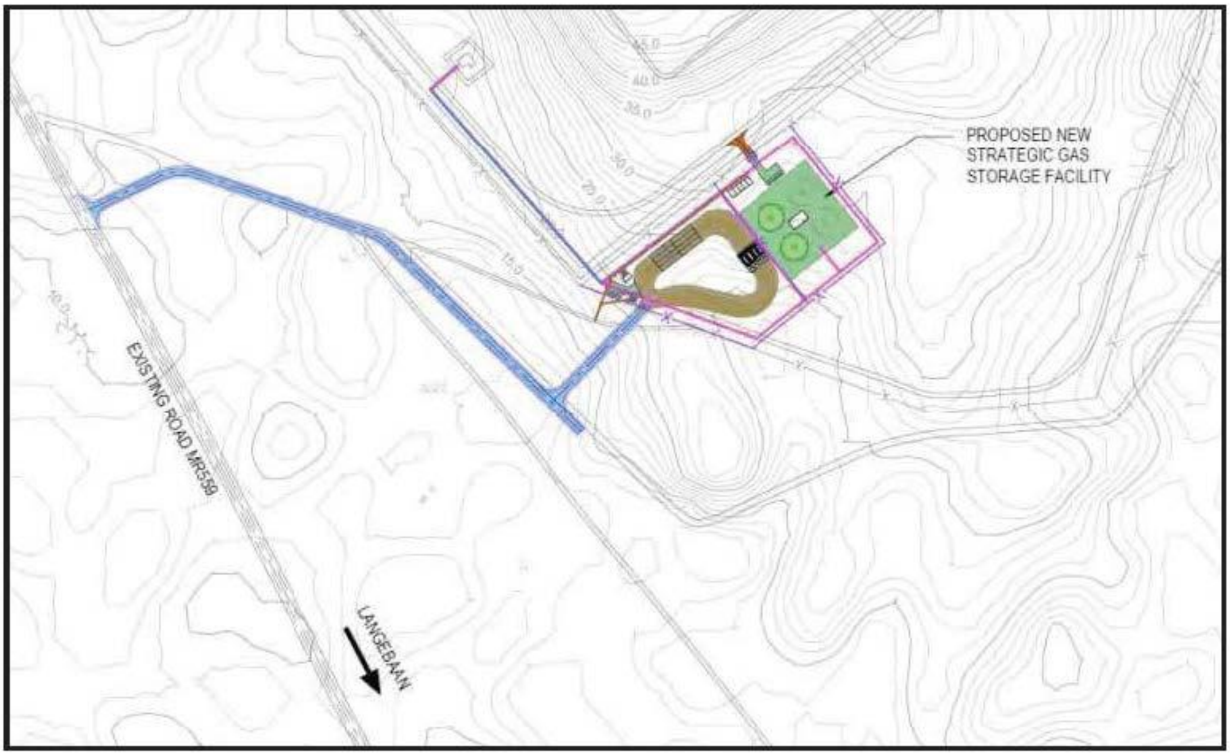
The flare is in operation during the offloading of a ship using a pilot flame fuelled from the LPG vessel.



Tank Farm Details 2

5.13. Proposed SFF LPG Import Terminal

The proposed facility will have 4 LPG spheres with a pump station and a road tanker loading gantry.



Proposed LPG Import Terminal

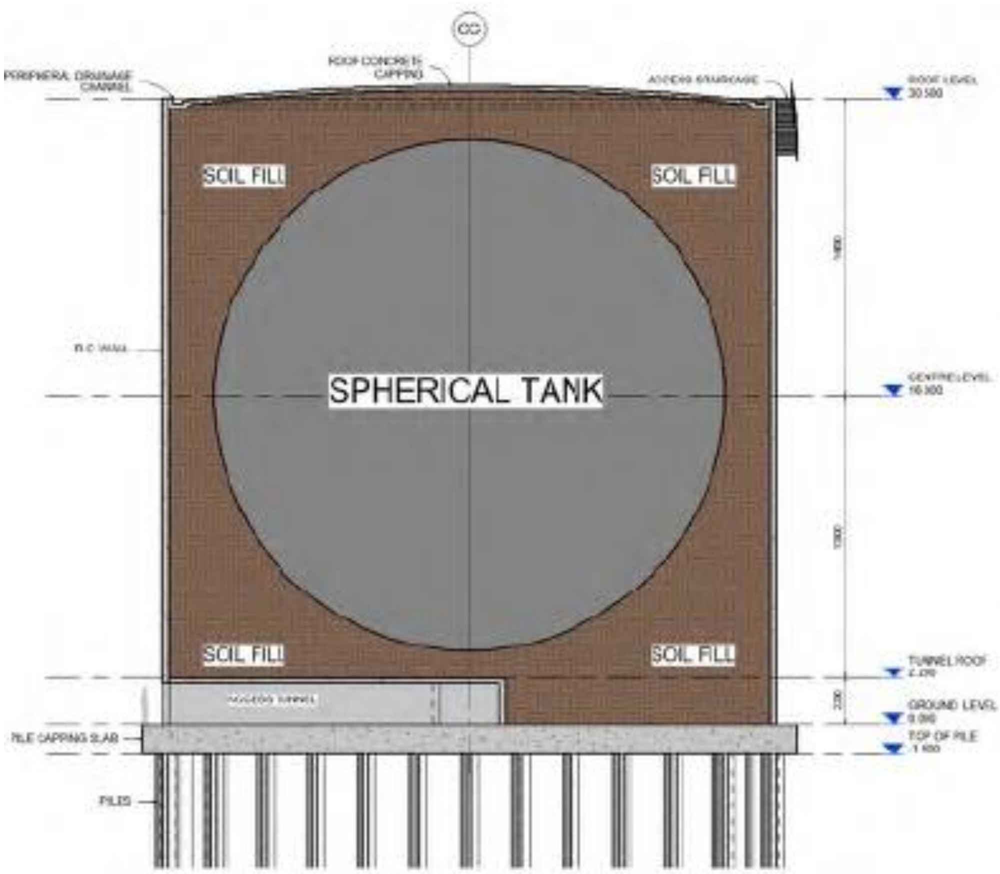
5.13.4. Bulk LPG Spheres

There will initially be two 4000ton mounded spheres for storage with an option for expansion of the facility by the addition of another two spheres. Each storage sphere is fitted with the following connections:

- 1 x 12” liquid receipt connection
- 1 x 4” emergency liquid recovery connection
- 1 x 10” liquid export connection
- 1 x 4” liquid minimum flow connection
- 1 x 6” vapour vent header connection (compressor discharge)
- 1 x 4” vapour transfer connection (compressor suction)

The vessels are also fitted with the necessary inspection man holes, tank gauging equipment as well as temperature and pressure monitoring equipment. The vessel over pressure protection provisions are catered for by the inclusion of Pressure Relief Valves.

The proposed mounds will be a sealed dry concrete cylindrical structure with an approximate radius of 15m and a height of 30m. The outer structural skin will be an in-situ reinforced concrete wall with compacted soil/sand filling between the wall and spherical tank. The mound roof is finished off with a pumpable soilcrete cap with a gravel top to serve as additional insulation and protection. A spiral staircase will provide access to the top and a reinforced concrete tunnel to the liquid discharge connection at the bottom. The tunnel will also be utilised for the reticulation of services.



Sphere Mound Design

The LPG storage shall normally be at a chilled condition of +5°C to +10°C. This provides for increased fill levels based on the lower thermal expansion of the stored contents and faster transfer times to road tankers and cargo carrier export. The chiller unit shall comprise of two air cooled glycol closed loop commercial chiller pack system, each chiller will be fitted to a storage vessel and will recirculate product to ensure stable operating temperatures

The use of product chilling allows for the terminal to maximise on the quantity of product that can be held in the vessels. Each vessel will be connected to a chiller unit which will be sized to maintain product temperatures between +5 to +10°C. The chillers will be connected to the vapour space of the vessel and flow will be induced by the vessel vapour pressure. The chiller unit will decondense the vapours and return the liquid to the vessel. Each unit will be located in a safe zone and will be based on a water-glycol intermediate configuration.

A Main Control Room is provided at the LPG Terminal. The control system incorporates the following functions:

- SCADA system as MMI (Man-Machine-Interface) to monitor and control LPG Terminal and Jetty operations;
- PCMS to monitor and control the following systems:
    - Ethyl Mercaptan injection;
    - Tank Farm Operations including Loading Gantries;
    - LPG Chiller Plant;
    - Storage Spheres;
    - Transfer Pumps;

    - Pipeline Evacuation System;
- Vehicle Booking System.

The Control Station will contain a segregated Emergency Shutdown System in a separate cubicle to take care of the following functions:

- Monitor Fire detection / CCTV combinations located throughout the LPG Terminal;
- Shut off the LGP flow in the event of fire (Flame detector);
- Shut off the LPG flow on detection of LPG at the Gas Detector;
- Shut off the LPG flow when the manual ESD button located at the operations is activated;
- Shut off the LGP flow on command from the PCMS;
- Any shut-down is reported to the PCMS.

5.13.5. Product Transfer

To allow for product flexibility, inventory management and maintenance/inspections requirements, the terminal supports inter-tank transfer operations. Inter-tank transfer is achieved as follows:

- The sphere that is to be evacuated shall be lined with its liquid export connection opened to the 10" suction header.
- The sphere vapour connections (6" vent header) will be opened to allow for pressure equalisation between the spheres.
- The receiving sphere will be lined up to allow for product receipt via the minimum flow liquid line.
- Once the vessel pressures have been equalised and the required vessel line ups confirmed, the discharging vessel pump will be started to commence liquid product transfer.
- On completion of the liquid discharge, the vessel to be evacuated will be connected to the drain tanker compressor to draw down the tank pressure and recover vapour product.

5.13.6. Pipeline Evacuation System

To provide a safe manner of evacuating pipelines for maintenance purposes the evacuation system allows liquid LPG to be drained to a drain tank, which is then transferred back to storage tank. The residual LPG vapour pressure remaining in pipelines can also be recovered to a safe level where it can be released through a strategically placed vent stack. Two identical systems shall be provided and shall consist of the following;

- 2 x LPG reciprocating compressors with liquid trap, directional change valve for liquid transfer and vapour recovery, instrumentation monitoring of low oil pressure, low suction pressure, high discharge pressure, high temperature and DOL motor control.
- Drain tank complete with level, pressure, and temperature monitoring.
- Fail Safe Fire Safe Spring Return air operated flow control valve.
- Fire safe manually operated flow control valves.
- 2 x 4m3/Hr Multistage Mag Drive Pump complete with 4kW VFD motor control.
- Auto and manual operation options.
- Local Stop/Start Station for compressor and pumping systems.
- Local lock and turn ESD control are locally provided.

- Modbus integration with a SIL 2 rated PCMS, SCADA monitoring with segregated Emergency Shutdown Systems.

Line evacuation will be achieved by initially draining liquid into the drain tanks. Thereafter, the compression of vapour from the line into the drain tank vapour space will enable the reduction of the vapour line pressure to safe levels at which time venting can take place. The compressed vapours drawn from the line will be directed back to storage.

5.13.7. Road Tanker Gantry

The road tanker gantry initially with 2 loading bays will be located just west of the spheres. The positioning of the gantry is such that it allows the road tankers to travel into the gantry and exit without any reversing or manoeuvring.

The road tanker gantry will be underneath a steel canopy fitted with an automatic deluge system for fire-fighting purposes. The design includes vapour lines and bottom loading arms on stand post structures with quick release couplings, flow meters and emergency shutdown valves. Fusible loop switches will be fitted, which when fused will activate an automatic shutdown process and the deluge system and fire water pumps will be activated. Metering will be by way of a weighbridge.

During road tanker filling, the dispensing storage vessel will be lined up with the liquid export connection opened to the suction header line. The pump used for transfer operations will be lined up so as to discharge product into the 4” road tanker loading line.

The tanker vapour space will be equalised with the storage vessel through the vapour connections situated at the loading bay. The vapour lines will connect to the storage vessel via the 4” vapour transfer connection. The liquid loading line will be connected to the road tanker via a 2” connection with suitable dry-break coupling to ensure minimum product loss in case of disconnection.

The liquid delivery line will also be fitted with two flow regulators which enable the control of road tanker filling rate. These control valves are utilised to manage the start and completion of transfer with liquid loading rates lowered during road tanker top off.

The transfer pump in operation will be started up once all valve alignment and road tanker connections have been made. Transfer rates of 60 m3/hr will be supported per road tanker bay. This transfer rate can be accomplished using a single transfer pump. Once the road tanker has reached a fill weight of 100 kg lower than the set point, the loading valve shall auto set to final slow fill transfer to maintain accuracy of loading. On the loading value being reached the loading valves shall close and if there are no other road tankers being loaded the pumps will stop.

In the event of a road tanker requiring product recovery, the road tankers mandatory pump will be used to discharge product back to storage. The road tanker vapour connection to storage will be opened to allow for tanker – storage pressure equalisation. This will be achieved by the use of the vapour transfer line. The road tanker will commence product transfer via its on-board pump and liquid will be transferred back to the terminal storage via the minimum flow return line. Once liquid transfer is complete, the vapour space of the road tanker can be evacuated using the terminal drain tank compressor. The compressor will draw the vapours off the road tanker via the 4” vapour transfer connection and dispense compressed vapour back to the terminal storage tanks.

5.13.8. Pump Manifold and Transfer Pumps

There will be two Low NPSH Multistage centrifugal LPG transfer pumps for Road Tanker Loading and Ship Export. The pumps will be integrated with a SIL 2 rated PCMS, SCADA monitoring with segregated Emergency Shutdown Systems.

5.13.9. Fire Fighting Systems

There will be a fire water ring main to provide water to:

- Fire Hydrants
- Fire Hose Reels and
- Road Tanker Gantry Deluge System

The firefighting systems will be integrated with a SIL 2 rated PCMS, SCADA monitoring with segregated Emergency Shutdown Systems.

5.14. Receiving Environment

The installations are located as indicated on the satellite image on the following page.

5.14.4. Topography of the Surrounding Area

The area surrounding the facility is flat industrial land with no water bodies close by.

5.14.5. Population Information

Area	Daytime Persons/Hectare	Night-time Persons/Hectare
Low Density Industrial	5	1

There are no sensitive population groups close to the site.

5.14.6. Surrounding Facilities and Other MHIs

The surrounding area consists of heavily industrial properties. There are 5 MHIs located to around the site which are as follows:

- Moggs Crude Oil Terminal 630m North-East
- Sunrise Energy 2.8km West North-West
- Avedia Energy 3.4km West
- AcelorMittal 2.4km North North-West

The following residential areas are located surrounding the industrial area of Saldanha Bay:

- Mykonos 3.2km South
- Bluewater Bay 6.5km West

The surrounding MHI's are indicated on the image on the following page.

Location of the SFF Import Terminal

5.15. Meteorological Information

The report describes the typical weather for the Saldanha Bay area in Gauteng over the course of an average year. It is based on the historical records from November 2014 to January 2017.

Saldanha Bay has a mild temperate climate with dry winters and warm summers.

2.10.1 Wind Directions

The wind is most often from the north-northwest (13.7% of the time), north-west (10.4% of the time), and east (3.8% of the time). The wind is least often from the east-southeast (2.5% of the time).

Over the course of the year typical wind speeds vary from 3m/s to 4m/s (light air to moderate breeze), rarely exceeding 10m/s (fresh breeze).

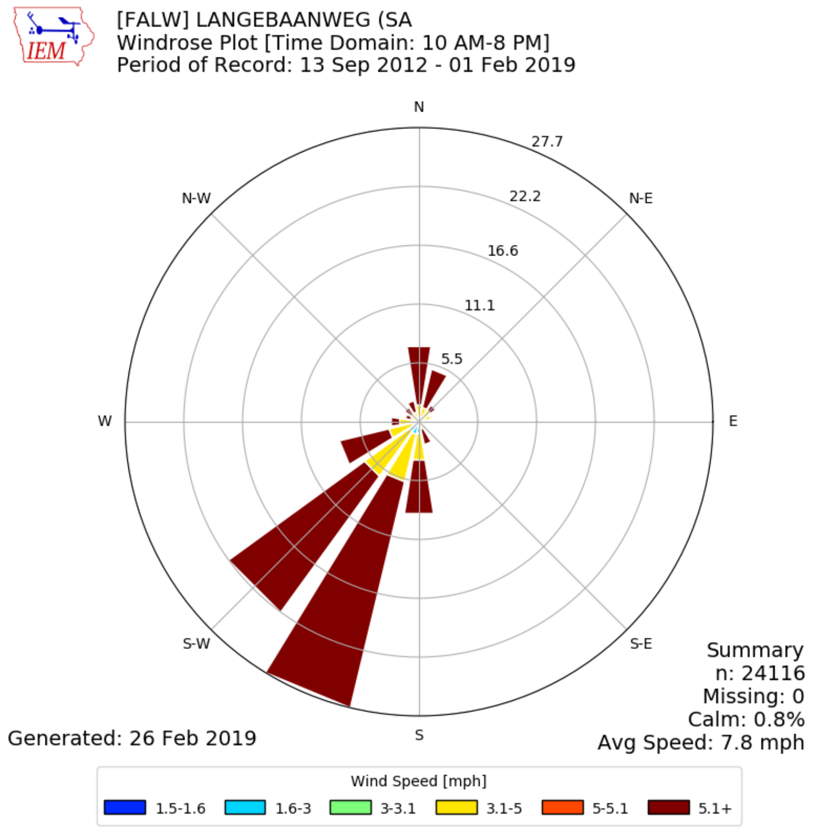
The highest average wind speed of 4m/s (gentle breeze) occurs from September to November, at which time the average daily maximum wind speed is 4m/s (moderate breeze). The lowest average wind speed of 3m/s (light breeze) occurs from December to August, at which time the average daily maximum wind speed is 3m/s (gentle breeze).

Month of year	Jan 01	Feb 02	Mar 03	Apr 04	May 05	Jun 06	Jul 07	Aug 08	Sep 09	Oct 10	Nov 11	Dec 12	Year 1-12
Dominant wind direction													
Wind probability >= 4 Beaufort (%)	15	32	19	24	13	28	22	43	45	42	57	29	**30**
Average Wind speed (kts)	8	9	8	8	7	9	9	10	11	11	13	9	**9**
Average air temp. (°C)	24	24	22	21	18	15	14	19	22	23	23	25	**20**

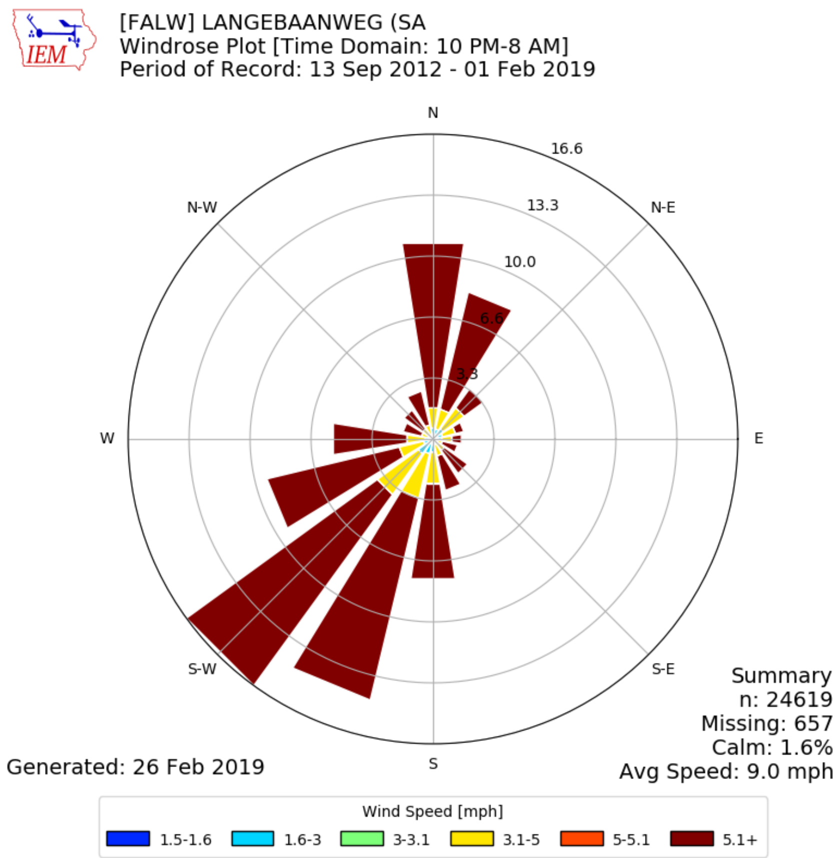
Saldanha Bay Yearly Weather Summary

2.10.2 Wind Rose

The annual wind rose for the area is as follows:



Wind Rose Day



Wind Rose Night

Dispersion models also require the atmospheric condition to be categorised into one of six stability classes, namely:

Stability Category	Meteorological Conditions	Occurrence
A	Very Unstable	Hot daytime conditions, clear skies, calm wind
B	Unstable	Daytime conditions, clear skies
C	Slightly Unstable	Daytime conditions, moderate winds, slightly overcast
D	Neutral	Day and night, high winds or cloudy conditions
E	Stable	Night-time, moderate winds, slightly overcast conditions
F	Very Stable	Night-time, low winds, clear skies, cold conditions

2.10.3 Summary

Based on the above information the meteorological information extracted for the modelling of scenarios was as follows:

- Wind and stability information:
  - B 1.5m/s meaning a stability class of B (moderately unstable conditions) where the wind speed is 1.5m/s.
  - D 5m/s meaning stability class of D (neutral conditions) where the wind speed is greater than 5m/s. D 5 gives a conservative daytime weather condition.
  - F 2m/s meaning a stability class of F (moderately stable) where the wind speed is less than or equal to 2m/s. This class is often used by the US Environmental Protection Agency for determining worst case scenarios for vapour cloud dispersion consequence analysis.
- The atmospheric temperature was set to be 24^{o}C, a typical summer temperature.
- The relative humidity was set to be 0.7.
- The solar radiation flux was set to be 0.5KW/m^2.
- The *Pasquill stability* was selected instead of the mixing layer height.

6. HAZARD IDENTIFICATION

This is the process of examining each work area and work task for the purpose of identifying all the hazards which are inherent to the job.

Hazard analysis is used as the first step in a process used to assess risk. The result of a hazard analysis is the identification of different types of hazards. A hazard is a potential condition and exists or not (probability is 1 or 0). It may be in single existence or in combination with other hazards (sometimes called events) and conditions become an actual Functional Failure or Accident (mishap). Once a hazard has been identified, it is necessary to evaluate it in terms of the risk it presents to the employees and the neighbouring community. In principle, both probability and consequence should be considered, but there are occasions where if either the probability or the consequence can be shown to be sufficiently low or sufficiently high, decisions can be made on just one factor.

During the hazard identification process the complete system of assets, materials, human activities and process operations within the boundaries of the site should be clearly defined and understood, taking account of the original design, subsequent changes and current conditions. Typically, the system should be divided into distinct separate components or sections to enable manageable quantities of information to be handled at each stage.

Some key questions and issues could be:

- What is the design intent, what are the broad ranges of activities to be conducted, what is the condition of equipment, and what limitations apply to activities and operations?
- What are the critical operating parameters? What process operations occur, and how could they deviate from the design intent or critical operating parameters? This should consider routine and abnormal operations, start-up, shutdown and process upsets.
- What materials are present? Are they a potential source of major accidents in their own right? Could they cause an accident involving another material? Could two or more materials interact with each other to create additional hazards?
- What operations, construction or maintenance activities occur that could cause or contribute towards hazards or accidents? How could these activities go wrong? Could other hazardous activities be introduced into this section by error or by work in neighbouring sections of the facility?
- Could other materials, not normally or not intended to be present, be introduced into the process?
- What equipment within the section could fail or be impacted by internal or external hazardous events? What are the possible events?
- What could happen in this section to create additional hazards, e.g. temporary storage or road tankers?
- Could a section of the facility interact with other sections (e.g. adjacent equipment, an upstream or downstream process, or something sharing a service) in such a way as to cause an accident?

6.6. Site Layout Details

The Site Plans and PID’s are included in the Appendices.

6.7. Significant Incidents at the Site and Related Sites

The crude oil installations are existing, and there have been some incidents involving the crude oil installations.

The LPG installations are proposed, but incidents at similar sites have been loss of containment of LPG resulting in fires. There have been no incidents of explosions and BLEVEs at bulk LPG installations in South Africa.

The few incidents that have occurred at similar installations were mainly caused by a lack of maintenance or negligence.

6.8. Preventative Measures to be Taken

A good Maintenance Plan must be compiled with a Maintenance Register. An Emergency Plan must be updated to include the risks as described in this report. At the Crude Oil Terminal, the systems are integrated with a SIL 1 rated PCMS, SCADA monitoring with segregated Emergency Shutdown Systems.

At the LPG Terminal the systems will be integrated with a SIL 2 rated PCMS, SCADA monitoring with segregated Emergency Shutdown Systems.

6.9. Hazard Details

6.9.4. Hazardous Materials

The materials on site were categorised as per SANS 10228:2003 classes of dangerous substance, shown in the table below:

Class	Description
1	Explosives (Not included in MHI Regulations)
2	Gases (Flammable or Toxic gases only)
3	Flammable Liquids
4	Flammable Solids
5	Oxidising Substances and Peroxides
6	Toxic and Infectious Substances
7	Radioactive Materials (Not included in MHI Regulations)
8	Corrosives
9	Combustible Materials

6.9.5. Hazardous Materials on Site

SFF uses hazardous products on site, categorised as per the table below:

Substance	Gases	Flammable Liquids	Flammable Solids	Potential for Buncefield	Potential for an MHI
Class	**2**	**3**	**4**		
LPG	Yes			Yes	Yes
Crude Oil		Yes		Yes	Yes
Nitrogen	Yes			No	No

As this Assessment deals only with the Crude Oil, LPG and Nitrogen the detailed properties of these products are included in the appendices.

6.10. Containment and Safety Systems in Design

The following containment and safety systems have been incorporated in the design of the proposed installation:

- A SIL 1 rated system is in place for the crude oil installations;
- A SIL 2 rated system will be used for the LPG installations;

- All relevant national and international codes, standards, regulations, legislation and guidelines will be used;
- All required fire-fighting and control systems will be in place;
- The auto shutdown systems are detailed in the appendices.

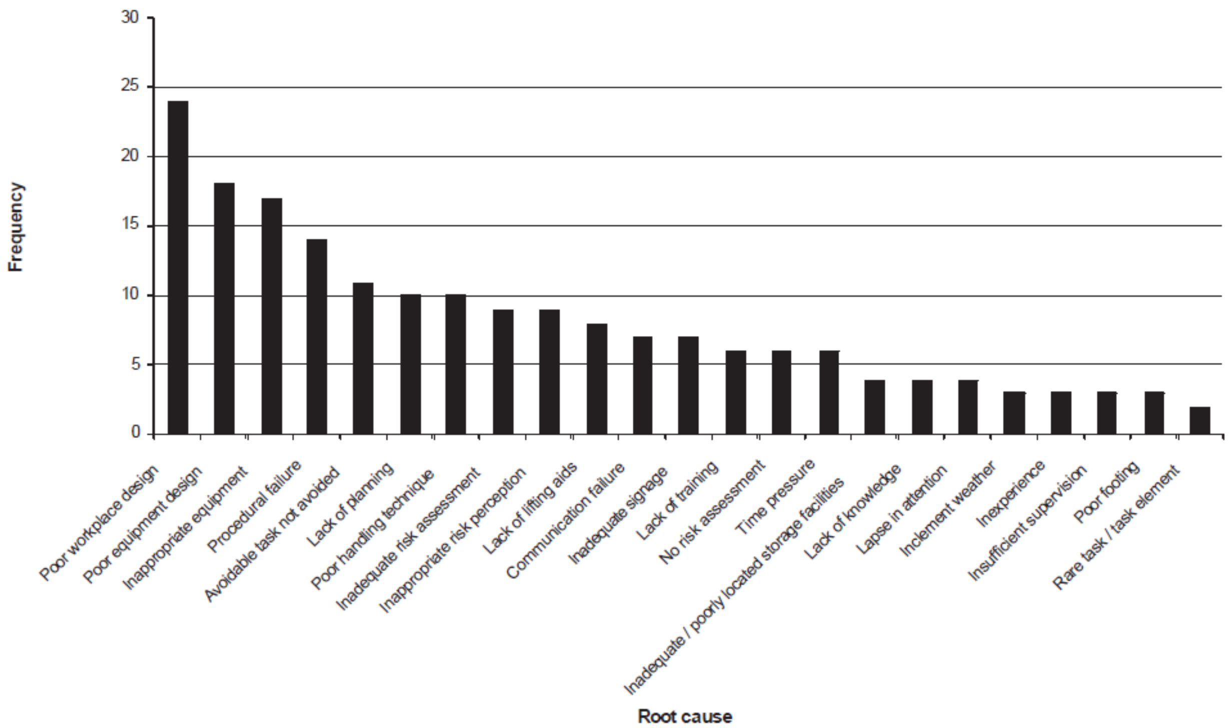
6.11. Environmental Hazards

Environmental Hazards are not included in the MHI Regulations and were not included in this report.

7. HAZARD ANALYSIS

7.6. Incident Root Causes

One hundred and twenty-six incidents were recorded an HSE report database in the UK. A greater number were reviewed but were not taken forward for analysis. The graph below shows the frequency with which each root cause was identified for the 126 incidents analysed.



The most common causes shown above are linked to the workplace and equipment available:

- Poor workplace design (representing 13%);
- Poor equipment design (10%);
- Inappropriate equipment (9%);
- Procedural failure (7%).

The next most commonly found issues are more closely linked with day-to-day organisation and management:

- Avoidable task not avoided (6%);
- Lack of planning (5%);
- Poor handling technique (5%);
- Inadequate risk assessment (5%);
- Inappropriate risk perception (5%).

The report mentions more than one root cause could be present in the same incident. In the sample analysed, 78 incidents were attributed to a single root cause; the remaining 48 had two or more root causes.

Most incidents are due to a mismatch between the operators' requirements or expectations and workplace or equipment design. If the root causes were principally to do with training or risk assessment (i.e. linked to risk perception and avoidance), it would imply that personnel were failing to use their experience and prior training to predict and avoid manual handling risks. Where an individual has unintentionally harmed themselves or others, it follows that the task carried risks which the operator(s) had to avoid by using safe working procedures and their skill and knowledge. The root cause in fact lies with one or more risky elements of the task that the operator then has to deal with. Training and experience help only to avoid the background risks.

The findings suggest that operators are mostly being injured because of poor equipment, task or workplace design, and to a lesser extent misunderstanding the level of risk. Failure to avoid an avoidable task is similar to a lack of planning as both indicate that an overview of the work was not held that could have highlighted alternatives to risky manual handling. 'Procedural failure' is linked to planning and overview too as this root cause indicates that agreed procedures inadvertently placed operators at risk of injury.

7.7. Events Following a Loss of Containment

Where no Boiling Liquid Expanding Vapour Explosion (BLEVE) and fireball occur following an instantaneous release with direct ignition, a liquid pool is formed, and a vapour cloud will expand to atmospheric pressure. The direct ignition of the vapour cloud is modelled as a flash fire (probability 0.6) and explosion (probability 0.4).

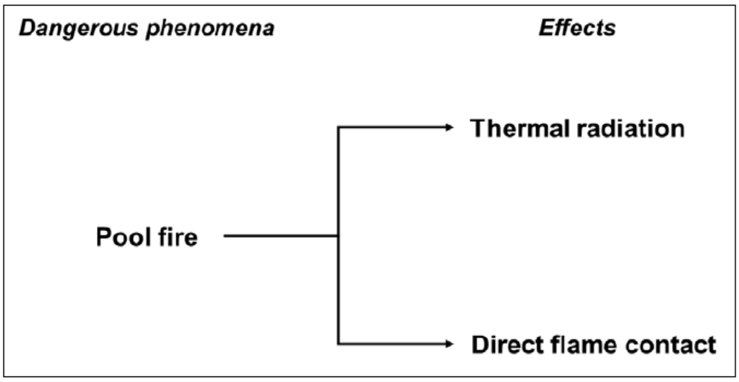
For an above-ground storage vessel (or road tanker), a BLEVE or fireball may occur. A BLEVE can occur when a flame impinges on a vessel containing a material that is a gas at atmospheric pressure and temperature but is a liquid at storage temperature and pressure. It is assumed that a BLEVE occurs when the vessel or road/ rail tanker is full. While BLEVEs are possible because of catastrophic vessel failure and localised vessel failure, they typically occur outside of these two events. Should this not occur, a vapour cloud may form. The ignition of the vapour cloud is modelled as a flash fire and explosion.

The flash fire is modelled through simulating the expansion of the initial cloud to the lower flammability limit (LFL) with air entrainment. The damage area then corresponds to the LFL cloud footprint. The explosion is modelled using the total mass subject to the lower flammability limit (LFL).

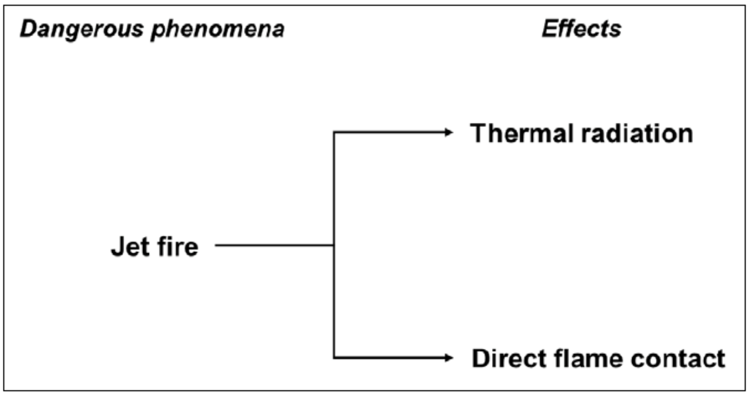
Accidental high velocity releases of ignited flashing liquids of pressurised flammable material at ambient temperature are classed as liquid jet fires. Jet fires occur when the jet of hydrocarbon can entrain air and burn at its edge. The jet remains ignited because the burning of the flame is greater than the velocity of the hydrocarbon jet, i.e. the flame can burn back towards the source of the jet. As a worst-case scenario, it is assumed that all failures occur in a horizontal position, i.e. the flame is orientated horizontally.

7.8. Event Trees

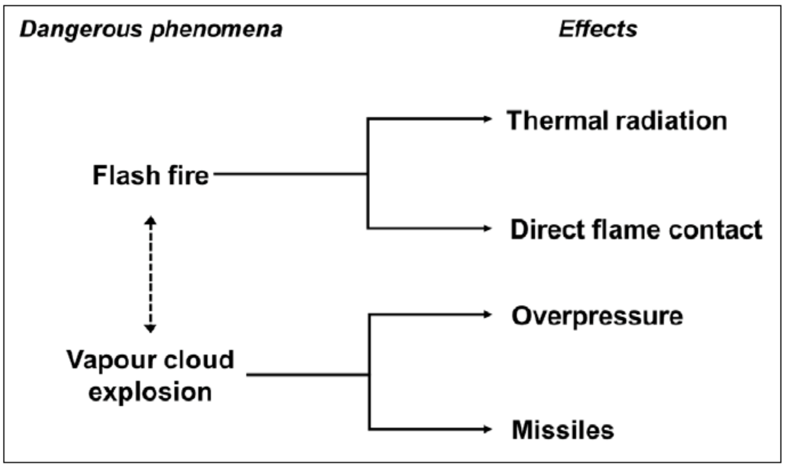
The probability of the flammable gas/ liquid release scenario identified above is represented as *event trees* for working daytime periods in the following diagrams.



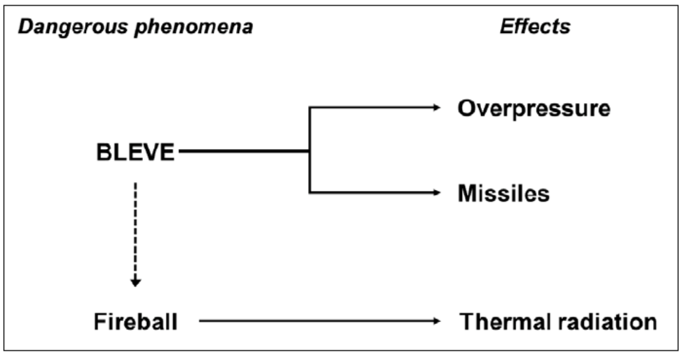
Physical Effects of a Pool Fire



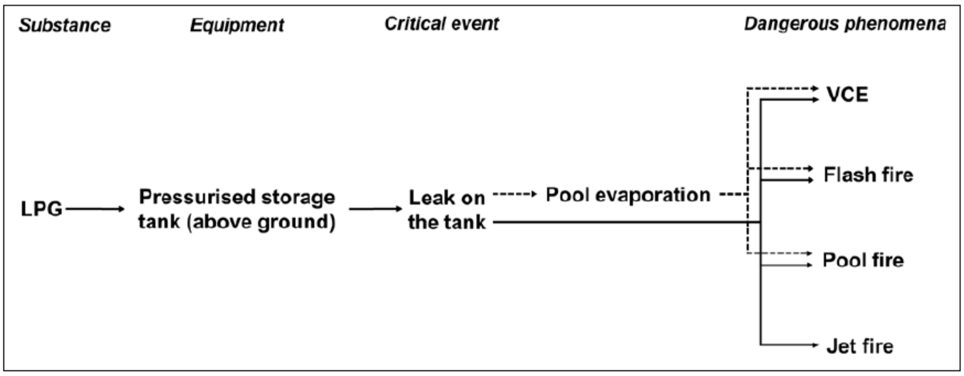
Physical Effects of a Jet Fire



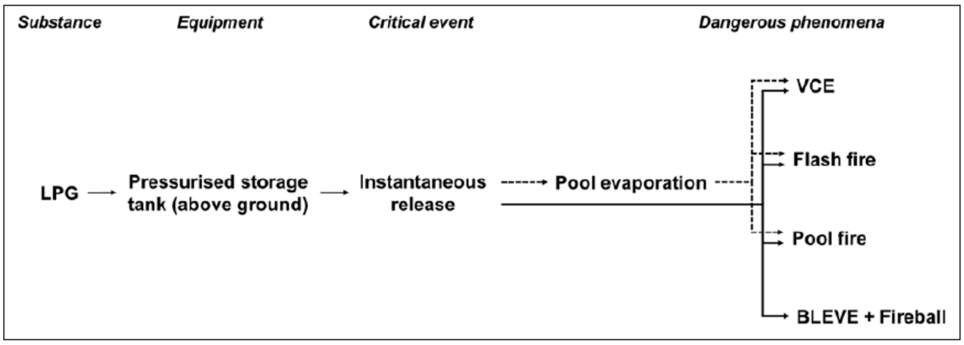
Physical Effects of a Flash Fire and Vapour Cloud Explosion



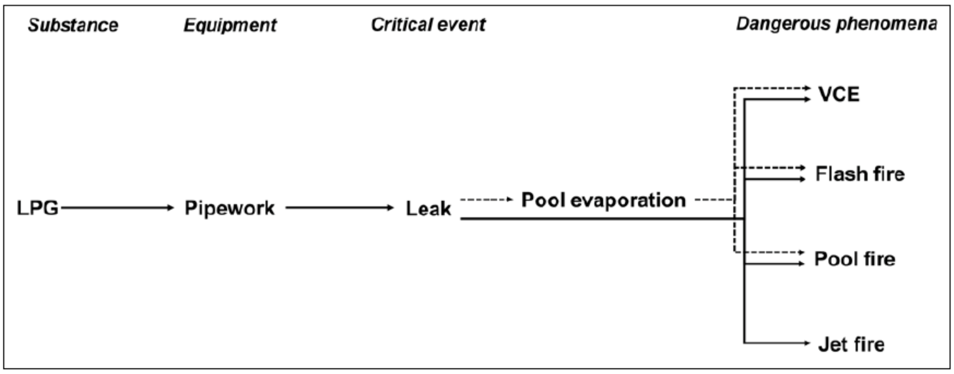
Physical Effects of a Fireball and BLEVE



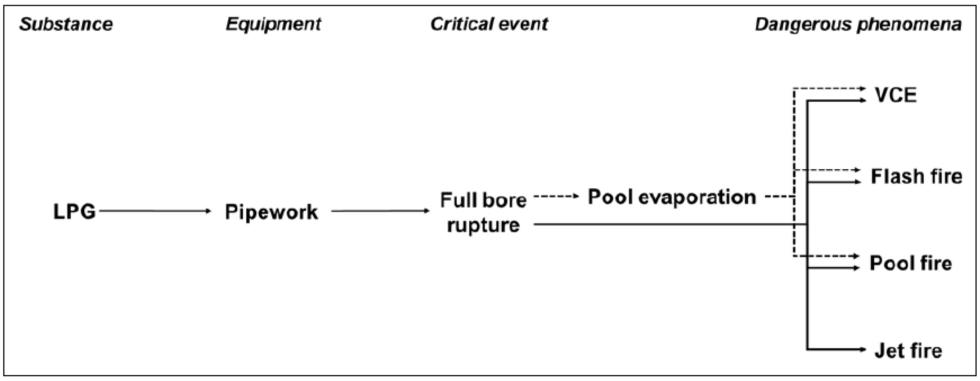
Event Tree of an LPG Vessel Leak



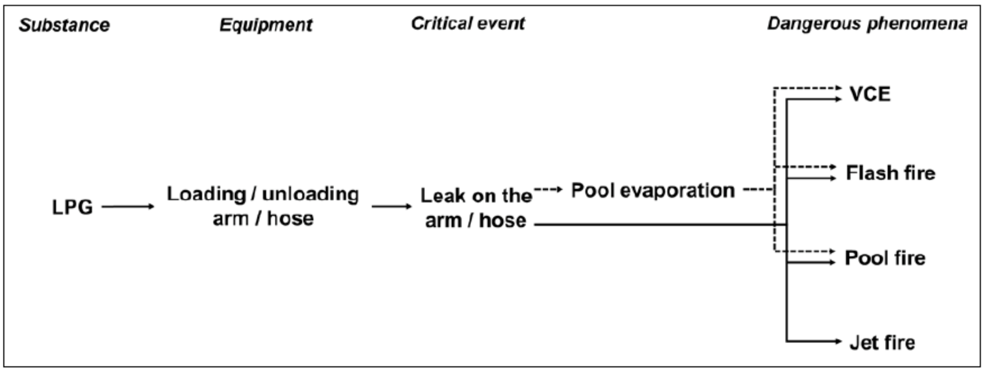
Event Tree of an Instantaneous Release of LPG Vessel



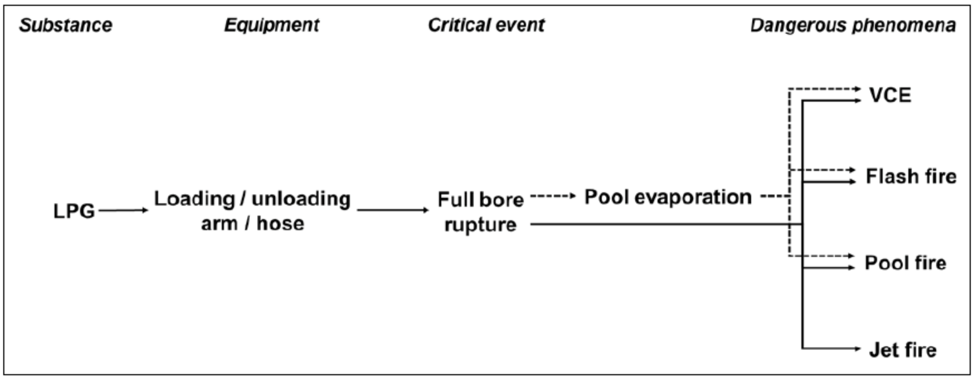
Event Tree of LPG Pipe Leak



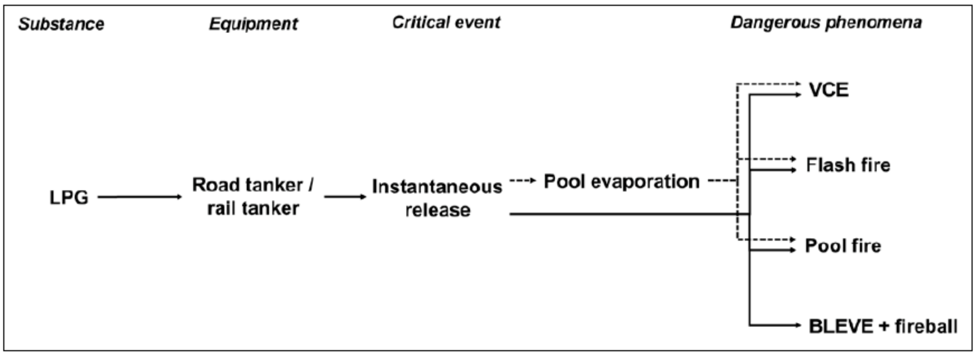
Event Tree of LPG Pipe Rupture



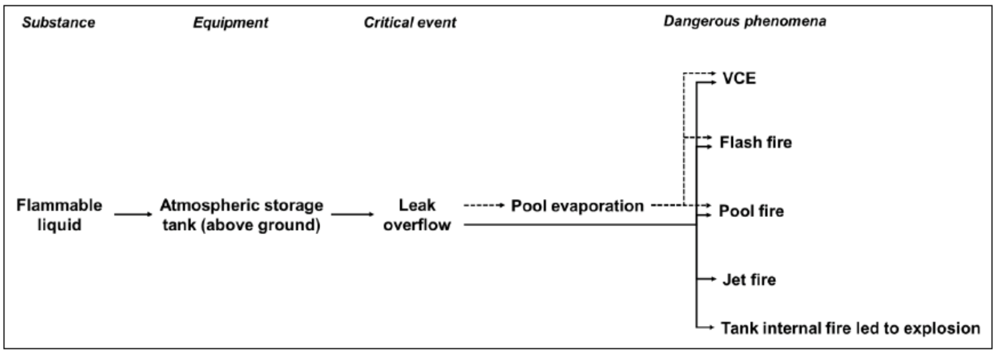
Event Tree of LPG Loading Hose Leak



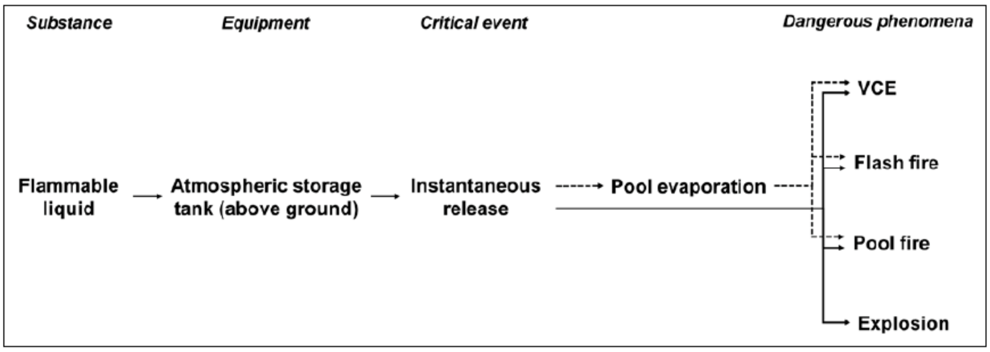
Event Tree of LPG Loading Hose Rupture



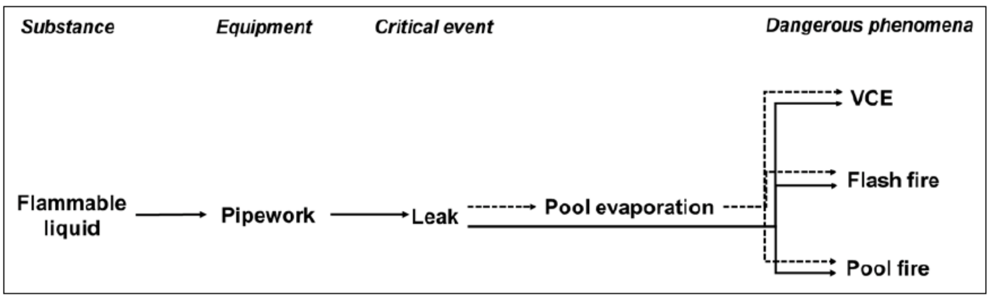
Event Tree of Instantaneous Release of a Road Tanker



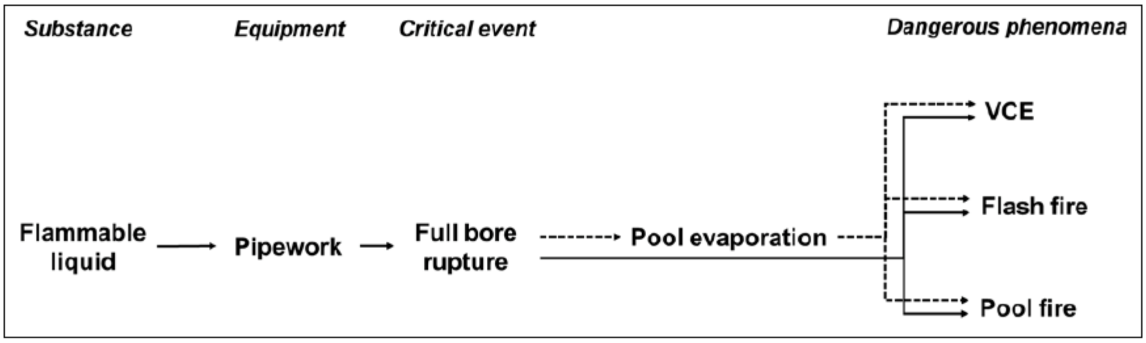
Event Tree of an Atmospheric Tank Leak



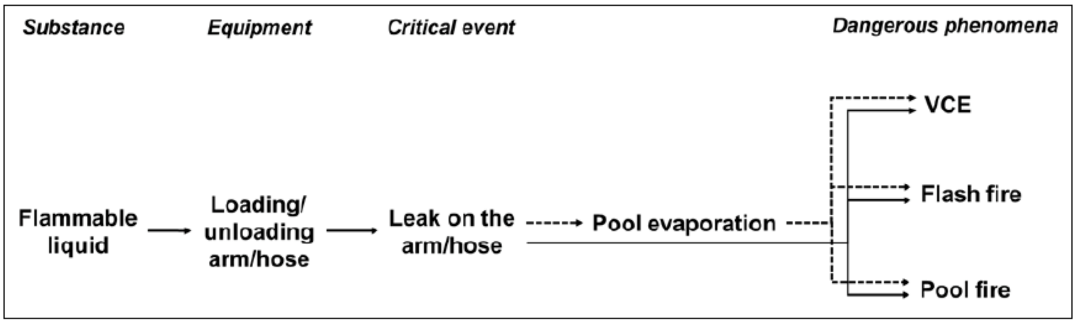
Event Tree of an Instantaneous Release of Atmospheric Tanks



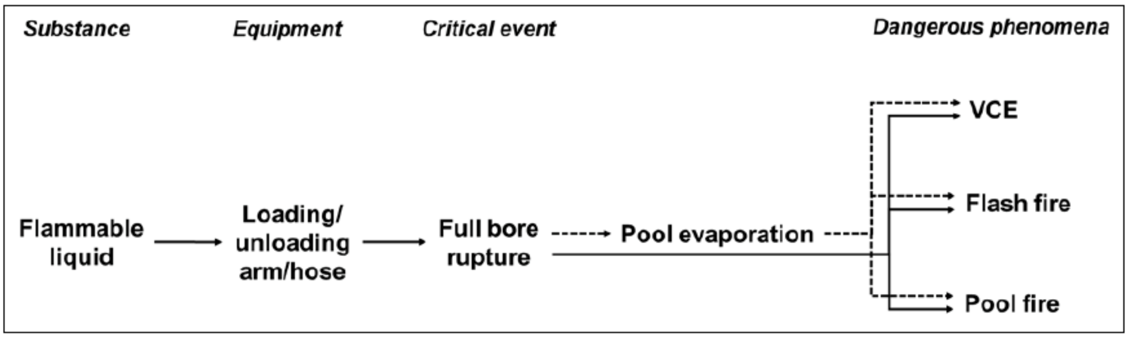
Event Tree of Flammable Liquid Pipe Leak



Event Tree of Flammable Liquid Pipe Rupture



Event Tree of Flammable Liquid Loading Arm Leak



Event Tree of Flammable Liquid Loading Arm Rupture

7.9. Scenarios to be Modelled

4.4.1. Jetty at the Port of Saldanha and the Pipeline to the Import Terminal

The following scenarios will be modelled for the Jetty and the Pipelines:

Pool Fire Scenarios

- Crude oil pool fire as the result of a 30mm leak in a loading arm;
- Crude oil pool fire as the result of a loading arm rupture;
- Crude oil pool fire as the result of a 30mm leak in the pipeline or a pig trap;
- Crude oil pool fire as the result of a pipeline rupture;

Jet Fire Scenarios

- LPG jet fire as a result of a loading hose leak;
- LPG jet fire as a result of a loading hose rupture;
- LPG jet fire as a result of a pipeline leak;
- LPG jet fire as a result of a pipeline rupture;

Flash Fire Scenarios

- Flash fire as a result of a crude oil pipeline rupture;
- Flash fire as a result of an LPG pipeline rupture;

Vapour Cloud Explosions

- VCE as a result of a crude oil pipeline rupture;
- VCE as a result of an LPG pipeline rupture;

4.4.2. Crude Oil Tank Farm

The following were modelled for the crude oil tank farm:

Pool Fire Scenarios

- Pool fire as a result of a crude oil pump leak;
- Crude oil tank top fire;

Flash Fire Scenarios

- Flash fire as a result of a crude oil loss of containment;

Explosion Scenarios

- VCE as a result of a crude oil loss of containment;
- Fixed roof tank explosion;

4.4.3. Crude Oil Import Terminal Flare LPG Installation

The following were modelled for the crude oil import terminal LPG installation:

Jet Fire Scenarios

- Jet fire as a result of an LPG vessel PRV leak;
- Jet fire as a result of an LPG road tanker loading hose leak;
- Jet fire as a result of an LPG road tanker loading hose rupture;

Flash Fire Scenarios

- Flash fire as a result of an LPG road tanker loading hose rupture;
- Flash fire as a result of an LPG vessel catastrophic failure;
- Flash fire as a result of an LPG vessel emptying in 10 minutes;
- Flash fire as a result of an LPG road tanker catastrophic failure;

Explosion Scenarios

- VCE as a result of an LPG vessel catastrophic failure;
- VCE as a result of an LPG road tanker catastrophic failure;

BLEVE Scenarios

- BLEVE of an LPG vessel;
- BLEVE of an LPG road tanker;

4.4.4. Crude Oil Terminal Nitrogen Installation

The following scenarios were modelled for the Crude Oil Terminal Nitrogen Installation:

- The catastrophic failure of a nitrogen vessel;

4.4.5. LPG Import Terminal

The following scenarios will be modelled for the LPG import terminal:

Jet Fire Scenarios:

- Jet fire as the result of a 10mm hole in one of the bulk vessels;
- Jet fire as the result of a loading arm shear;
- Jet fire as the result of a 10mm hole in a loading arm;
- Jet fire as the result of a pump failure;

Flash Fire Scenarios:

- Flash fire as the result of a loading arm failure;
- Flash fire as a result of an LPG vessel catastrophic failure;
- Flash fire as a result of an LPG vessel emptying in 10 minutes;
- Flash fire as a result of an LPG road tanker catastrophic failure;

Explosion Scenarios:

- VCE as the result of a catastrophic vessel failure;
- VCE as the result of a catastrophic road tanker failure;
- BLEVE of a road tanker.

8. CONSEQUENCE ANALYSIS

8.6. Background

The consequence analysis describes the extent of impacts from major events. The results of this analysis are used as input to the risk analysis section as well as providing guidance to emergency planning.

In order to establish the impact following an accident, it is necessary to first estimate the physical process of the spill (i.e. rate and size), spreading of the spill, the evaporation from the spill and the subsequent atmospheric dispersion of the airborne cloud or, in the case of ignition, the burning rate, the resulting thermal radiation or the overpressures from an explosion.

The second step is to estimate the consequences of a spill on humans and structures. For humans this is normally expressed as a probability of fatality at distances from the release point.

The consequence analysis as documented in the Risk Assessment report is to provide enough process data, calculations etc. to allow for a reasonable verification of key consequence modelling results.

8.7. Source Term Analysis

When determining the amount of materials possibly released or involved in an incident, the following aspects should be considered:

- The amount of material available for release from each item should be at least the full inventory of the piece of equipment when it is filled to its maximum capacity. The maximum capacity of equipment is the total amount of fluid that can be accommodated in that equipment in the absence of equipment inventory control. For example, the volume of a cube vessel would be the product of the width, length and height of the vessel.
- When a component fails, such as a vessel, subsequent delivery of other system components which are connected with the vessel may take place. If the quantity that is subsequently delivered is significant, the combined volume/flows need to be taken into consideration.
- If in the case of an on-site pipeline failure an increased pumping rate occurs, this is modelled by increasing the flow rate to that of 1.5 times the pumping rate.
- The effects of measures affecting outflow, such as shutting off valves can be considered.
- In the case of a 'long pipeline' rupture scenario the outflow is calculated based upon the content of the pipeline and a pumping rate. This means that the outflow from a reservoir that may be connected is not included. The 'long pipeline' scenario can therefore only be used when the pumping rate and the content of the transport pipeline is critical for the outflow. It is also important that the condition that L/D > 1000 is complied with, where L is the (total) length of the pipeline and D is the diameter of the pipeline.
- In the case of a line rupture, outflow occurs from both ends of the rupture. There are several possibilities:
  - If the outflow mainly takes place from one end, the scenario can be modelled as a rupture of one pipeline ('line rupture').
  - If the rupture occurs in a long transport pipeline, the various contributions from both ends of the rupture are included in the calculation of the outflow.
  - If the contributions from both ends of the line rupture are relevant to the outflow, one effective pipeline diameter must be used in the calculation, for which the outflow rate matches the outflow rate from both ends added together.

8.8. Fires

Flammable liquids and gases may ignite and burn if ignited. This normally occurs as a result of a loss of containment and ignition. Fires include pool fires, jet fires and flash fires.

The consequence of a fire will be thermal radiation.

It is expected that an individual either in pain from a thermal dose received or suffering from first degree burns should escape rapidly as the injury should not be sufficient to impede movement, yet the pain will be too uncomfortable to bear standing still.

An individual with second degree burns will have even greater motivation to escape, commonly referred to as the fight or flight response. However, at this level of injury, any exposed skin will be very uncomfortable and difficult to use in contact with another surface. Simple tasks, such as turning door handles or dressing in survival equipment will take longer, if at all possible. Depending on the location and extent of injury, more difficult tasks such as operating control panels or turning valves may be impossible.

With third degree burns an individual will be in severe pain and will realise that they are in immediate danger of losing their life. Individual response is hard to predict. Fine control of injured extremities will be impossible and other functions will be severely impaired. Escape will probably incur further injury as skin may fall away from the wound. Individuals with third degree burns should be considered as casualties who cannot evacuate unaided.

Thermal radiation levels used in this report are as follows:

- 4.5 kW/m^2 is the radiation that would cause pain and second degree burns within 20 seconds (Yellow Contour).
- 12.5 kW/m^2 represents a 1% fatality for people exposed to the fire for 20 seconds (Orange Contour).
- 37.5 kW/m^2 indicates the lower limit of damage to steel equipment and represents a 100% fatality for people exposed to the flame (Red Contour).

8.8.4. Thermal Radiation

The effect of thermal radiation is dependent on the type of fire and duration exposed to the thermal radiation. Codes such as API 520 and 2000 suggest the maximum heat absorbed on vessels for adequate relief designs to prevent the vessel from failure due to overpressure. Other codes such as API 510 and BS 5980 give guidelines for the maximum thermal radiation intensity as a guide to equipment layout.

The effect of thermal radiation on human health has been widely studied and it has been found that injuries developed due to the exposure and intensity of the radiation. Two values normally quoted are 1.5kW/m^2 or 'safe' value where people can be exposed for a long period of time and 5kW/m^2 for people performing an emergency operation for short periods of time.

Thermal Radiation Guidelines (BS 5980-1990)

Thermal Radiation Intensity (kW/m^2)	**Limit**
1.5	Will cause no discomfort for long exposure
2.1	Sufficient to cause pain if unable to reach cover within 40 seconds
4.5	Sufficient to cause pain if unable to reach cover within 20 seconds
12.5	Minimum energy required for piloted ignition of wood and melting of plastic tubing
25	Minimum energy required to ignite wood at indefinitely long exposures
37.5	Sufficient to cause serious damage to process equipment

8.8.5. Pool Fires

On ignition of a flammable pool, the fire would extend to the limit of the pool but would shrink rapidly as the fuel within the pool is consumed.

A loss of containment of gas does not normally result in the formation of a flammable pool. Pool fires at the gas installations were therefore not modelled.

8.8.6. Jet Fires

Jet fires occur when flammable material of a high exit velocity ignites. Ejection of flammable material from a vessel, pipe or pipe flange may give rise to a jet fire and in some instances the jet flame could have substantial ‘reach’. Depending on wind speed, the flame may tilt and impinge on pipelines, equipment or structures. The thermal radiation from these fires may cause injury to people or damage equipment some distance from the source of the flame.

8.8.7. Flash Fires

A loss of containment of flammable materials if not immediately ignited, would mix with air and form a flammable cloud. This cloud could drift and if ignited could result in a flash fire or vapour cloud explosion.

The cloud of flammable material would be defined by the lower flammable limit (LFL) and the upper flammable limit (UFL). An ignition within a flammable cloud can result in an explosion if the front is propagated by pressure. If the front is propagated by heat, the fire moves across the flammable cloud at the flame velocity and is called a flash fire. In some instances, pockets of flammable clouds may extend beyond the LFL due to localised conditions. The ½ LFL endpoint assumes there are no isolated pockets and that ignition would not occur beyond this point.

8.9. Explosions

An explosion is a rapid increase in volume and release of energy in an extreme manner, usually with the generation of high temperatures and the release of gases. Supersonic explosions created by high explosives are known as detonations and travel via supersonic shock waves. Subsonic explosions are created by low explosives through a slower burning process known as deflagration.

Explosions associated with flammable gas installations are vapour cloud explosions (subsonic explosions), confined vapour cloud explosions (supersonic explosions) and boiling liquid expanding vapour explosions (BLEVE).

8.9.4. Vapour Cloud Explosions

A vapour cloud is formed by the release and mixing of a flammable vapour, gas or spray from an installation. The concentration of the material mixture within the vapour cloud must be in the explosive range to ignite and cause an overpressure. The rate of acceleration of the flames within the vapour cloud will lead to significant overpressure. Should the rate of ignition in the vapour cloud be instantaneous an explosion will occur. The rate of ignition will be influenced by the confinement of the vapour cloud. This will lead to a higher concentration of the flammable mixture. The results of a vapour cloud can be extensive property damage and injury or loss of life.

8.9.5. Unconfined Gas Explosions

An unconfined gas explosion is a flammable gas cloud that detonates within an area that is uncluttered and the expanding gases can easily escape. The maximum overpressure from an unconfined gas explosion is much lower than that of a confined explosion and hence the overpressure distance to safety is lower.

8.9.6. Confined Gas Explosions

Vapour cloud explosions are one of the most devastating events which can occur in the process industries. It was recognised that a facility design should include limiting explosion damage. The determination of peak overpressures from gas explosions and development of design criteria for structural support become more complex due to high pressure inventories in congested areas.

There are four key factors in an explosion. These are related to the overpressure which is the pressure rise above normal atmospheric pressure, the positive phase duration which is the time during which the pressure is above atmospheric pressure, the degree of confinement of the flammable mixture which causes turbulence and acceleration of the flame front and influences the overpressure, and the impulse (area under the pressure-time profile).

It is well established that it is not the size of the vapour cloud that matters when it comes to blast strength, but the degree of confinement of the vapour cloud and congestion in the path of the flame front. The energy of ignition source (e.g. naked flame) plays a dominant role in determining the blast strength, although a well-designed facility with strict implementation of hazardous area classification requirements in terms of hardware and safety management system can reduce the strength of a potential ignition source significantly.

The Multi-Energy Model (MEM) for rapid assessment of explosion overpressure has been developed by TNO (1997). It is based on the concept that significant overpressures can be generated by the ignition of a vapour cloud only in the presence of partial confinement or obstacles in the path of the flame front. This model, however, requires assumptions on the initial blast strength, which significantly influences the predictions. CFD models used in offshore modules have shown that rapid assessment models can underestimate the blast overpressures.

8.9.7. Boiling Liquid Expanding Vapour Explosion (BLEVE)

Boiling liquid expanding vapor explosions (BLEVEs) are one of the most severe accidents that can occur in the process industry or in the transportation of hazardous materials. Strictly speaking, these explosions do not necessarily imply thermal effects. However, in most cases the substance involved is a fuel that causes a severe fireball after the explosion. Usually BLEVE refers to the combination of these two phenomena, BLEVE and fireball, i.e., to an accident simultaneously involving mechanical and thermal effects.

While the explosion of a tank containing a pressurised flammable liquid will almost always lead to a fireball, the explosion cannot always be considered strictly a BLEVE. To qualify as this type of explosion, the following conditions must be met:

- **Significant superheating of the liquid.** Most liquefied gases under fire attack (GAS, ammonia, chlorine) fulfil this condition; it can also be fulfilled by other liquids contained in closed containers that undergo anomalous heating, for example due to a fire; and, as stated before, water can also be at this condition upon instantaneous depressurisation.
- **Instantaneous depressurisation.** This phenomenon is usually related to the type of failure of the vessel. The sudden pressure drop in the container upon failure causes the liquid superheat. If the liquid superheat is significant, the flashing may be explosive.

When these two conditions are met, a practically instantaneous evaporation of the contents takes place, with the formation of a large number of boiling nuclei in all the liquid mass (homogeneous nucleation). In these conditions the velocity at which the volume increases are extraordinary and the explosion is therefore very violent. Strictly speaking, this is the phenomenon associated with the BLEVE explosion.

When a BLEVE explosion involves a flammable substance, it is usually followed by a fireball, intense thermal radiation will be released and fragments from the shattered vessel. The thermal energy is released in a short time, usually less than 40 seconds (although this time is a function of the mass in the tank). The phenomenon is characterised from the first moments by strong radiation; this eliminates the possibility of escaping for the persons nearby (who also will have suffered the effects of the blast).

A blast wave from a BLEVE is fairly localised but can cause significant damage to immediate equipment.

8.10. Site Specific Consequence Analysis

At the flammable and toxic installations, the impacts of a loss of containment has been calculated without taking the probability of it occurring into account. This is done to show the consequence of the incident and how it will impact on the site and the surrounding area. Domino effects are also investigated in this section.

In the following sections various scenarios are calculated for the installation.

8.10.4. Jetty and the Pipeline

The following are the consequences of pool fires, jet fires, flash fires and explosions at the jetty and the cross-country pipeline.

a) Pool Fires

When a crude oil tanker ship is being offloaded or loaded a floating bund is built around the loading arms area to contain any spills. A loss of containment at the loading area would result in a pool. The pool will be a contained pool in the bunded area.

On ignition of a contained flammable pool, the fire would extend to the limit of the pool but would shrink rapidly as the fuel within the pool is consumed.

The following mitigation will be implemented to deal with consequences of a pool fire in the tank farm:

- Construction of temporary floating bund;
- Use of foam canons.

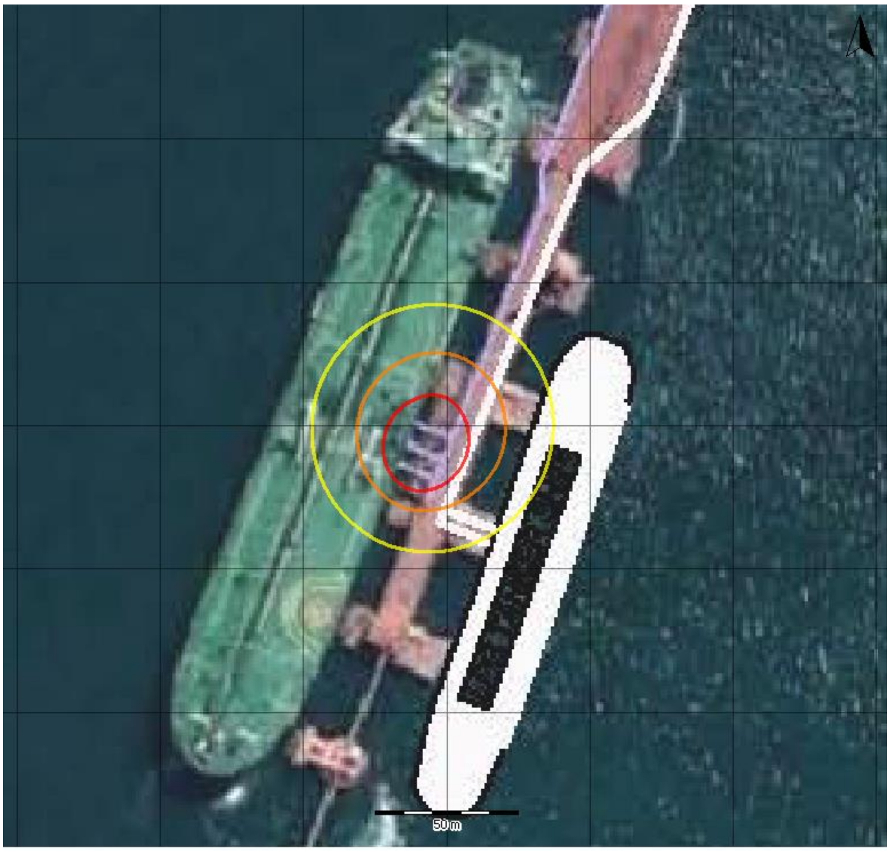
The consequences of a pool fire are as follows:

Crude Oil Loading/Offloading			
Loading Arm	**Radiation Contour 37.5kW/m^2**	**Radiation Contour 12kW/m^2**	**Radiation Contour 4.5kW/m^2**
Catastrophic Leak	84m	132m	193m
30mm Leak	22m	37m	53m

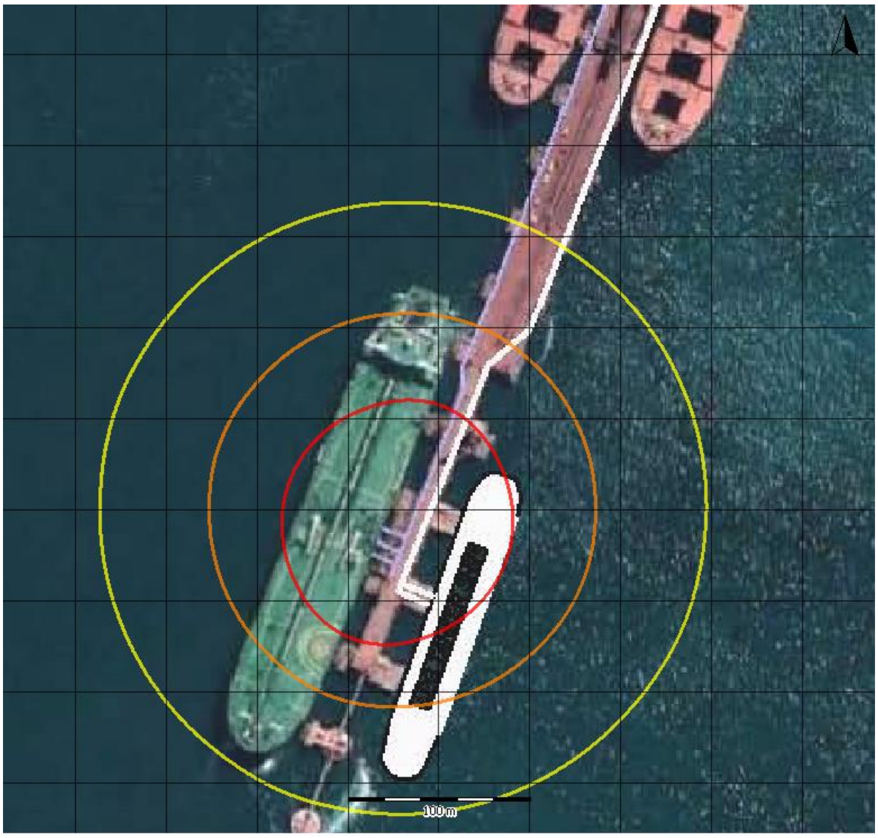
Crude Oil Pipeline			
Pipeline or Pig Trap	**Radiation Contour 37.5kW/m²**	**Radiation Contour 12kW/m²**	**Radiation Contour 4.5kW/m²**
Catastrophic Leak	43m	71m	104m
30mm Leak	22m	37m	54m

Thermal radiation from a pool fires at the crude oil jetty and pipeline are shown below.

- 4.5 kW/m² is the radiation that would cause pain and second degree burns within 20 seconds. (Yellow Contour)
- 12.5 kW/m² is the energy required for pilot ignition of wood. (Orange Contour)
- 37.5 kW/m² indicates the lower limit of damage to steel equipment and represents a 100% fatality for people exposed to the flame. (Red Contour)



Crude Oil Pool Fire as a result of a Loading Arm Leak



Crude Oil Pool Fire as a result of a Loading Arm Rupture

Crude Oil Pool Fire as a result of a Pipe Leak

Crude Oil Pool Fire as a result of a Pipe Rupture

b) Jet Fires

For this Assessment, jet fires from a 10mm leak and a pipe rupture was assumed. The worst-case scenario of the jet fire being horizontal and in the same direction of the wind was assumed.

Thermal radiation from jet fires is shown below.

- 4.5 kW/m^2 is the radiation that would cause pain and second degree burns within 20 seconds. (Yellow Contour)
- 12.5 kW/m^2 is the energy required for pilot ignition of wood. (Orange Contour)
- 37.5 kW/m^2 indicates the lower limit of damage to steel equipment and represents a 100% fatality for people exposed to the flame. (Red Contour)

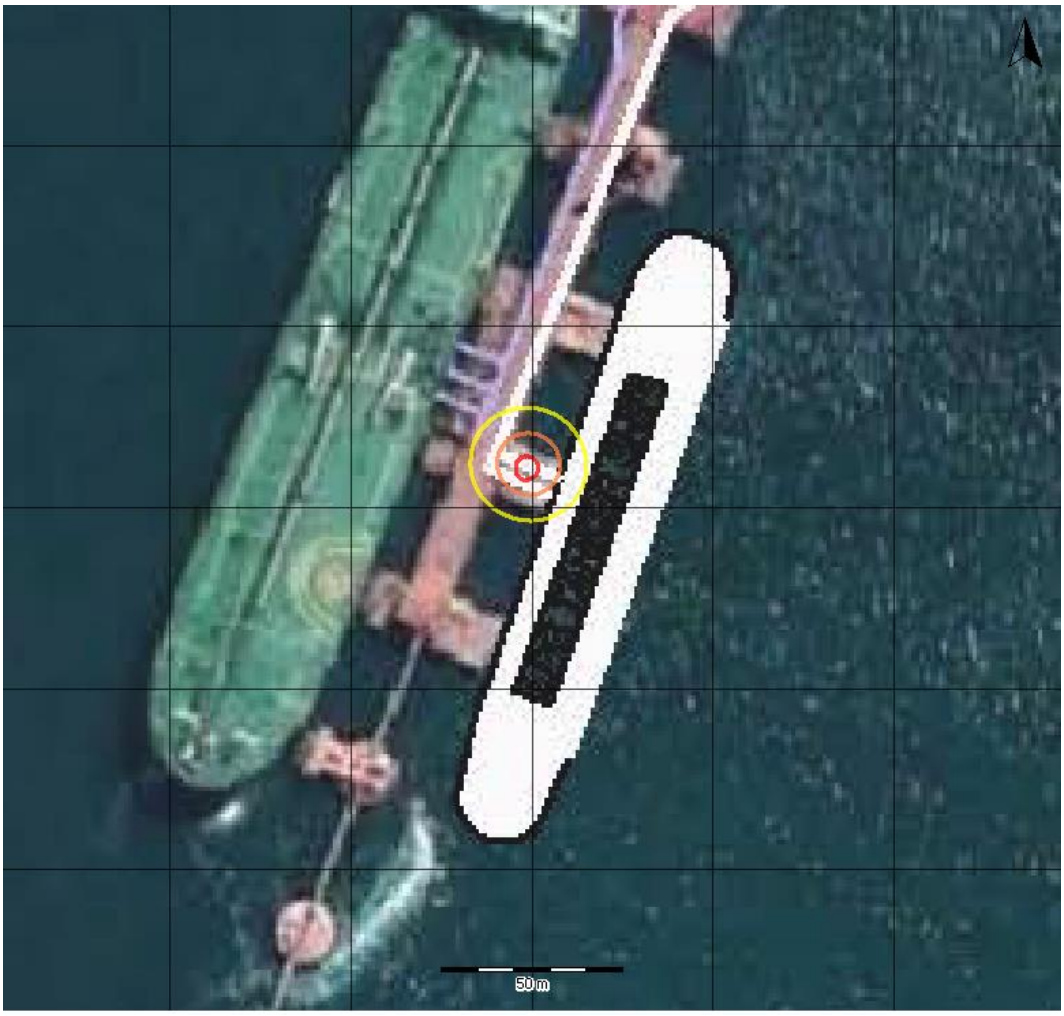
The following mitigation is currently installed and implemented to deal with consequences of an LPG release at the jetty and pipeline:

- All fire-fighting equipment will be in place;
- Emergency stop switch installed to stop pumps;
- Braided hoses will be used;
- Gas detection;

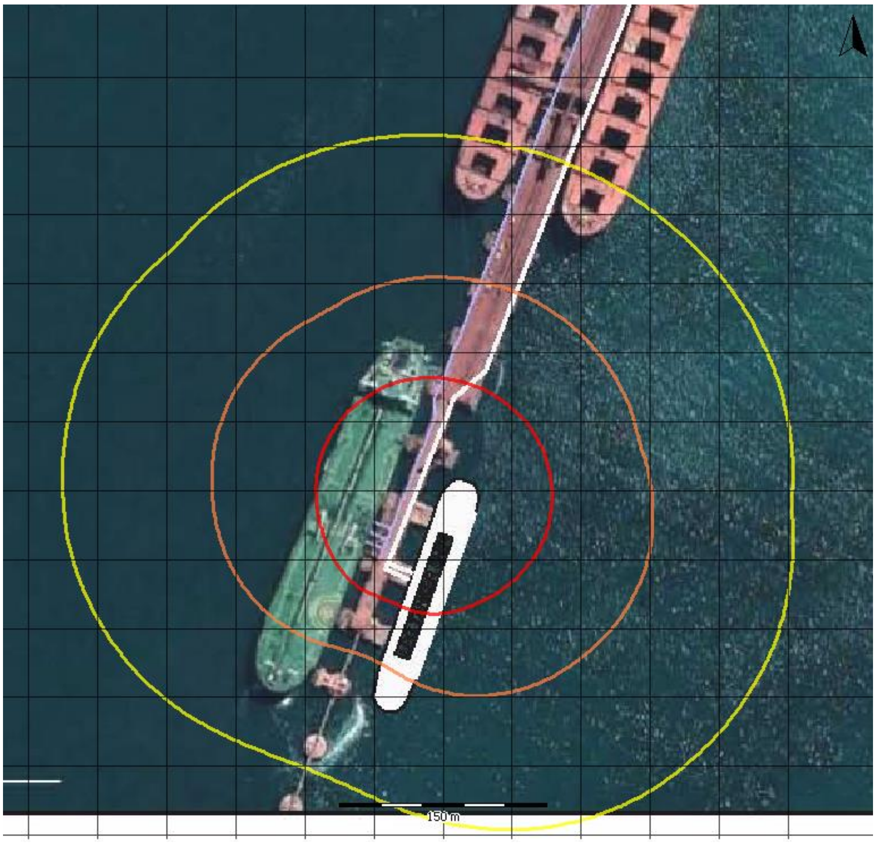
The consequences of jet fires are as follows:

LPG Loading Hose			
Loading Hose	**Radiation Contour 37.5kW/m²**	**Radiation Contour 12kW/m²**	**Radiation Contour 4.5kW/m²**
Catastrophic Hose Failure	148m	221m	322m
10mm Hose Leak	5m	11m	18m

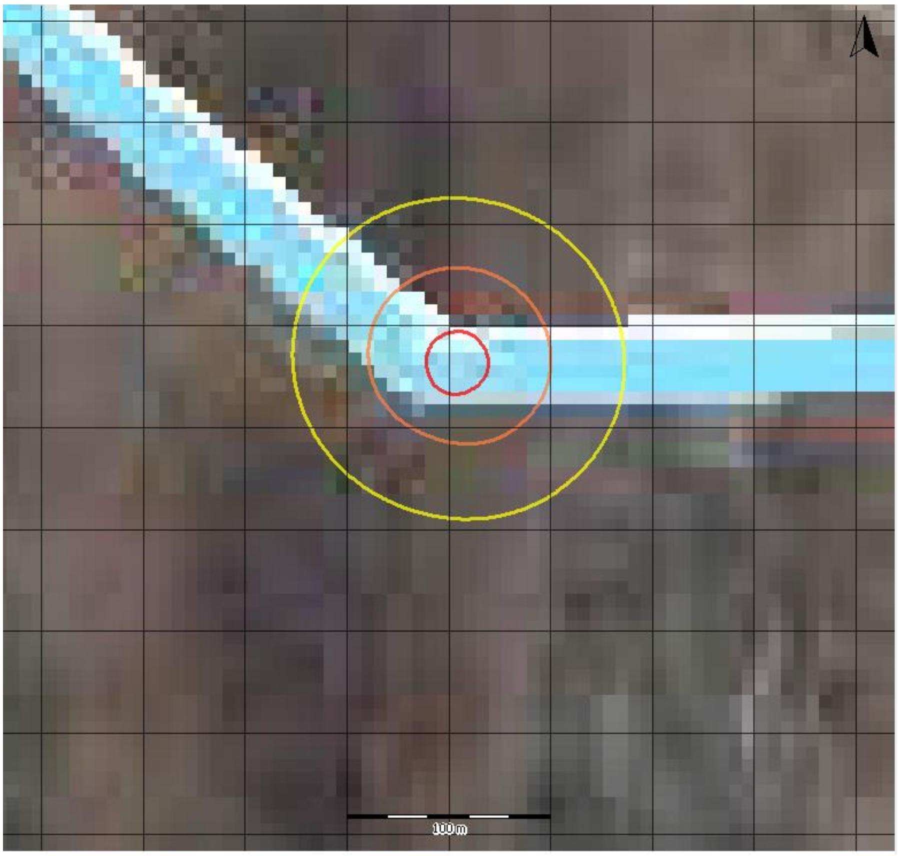
LPG Cross-country Pipeline			
Pipeline	**Radiation Contour 37.5kW/m²**	**Radiation Contour 12kW/m²**	**Radiation Contour 4.5kW/m²**
Catastrophic Pipe Failure	27m	58m	92m
10mm Pipe Leak	1m	2m	3m



10mm Jet Fire Loading Hose Leak



Jet Fire from a Loading Hose Rupture



Jet Fire from a Pipe Rupture

c) Flash Fires

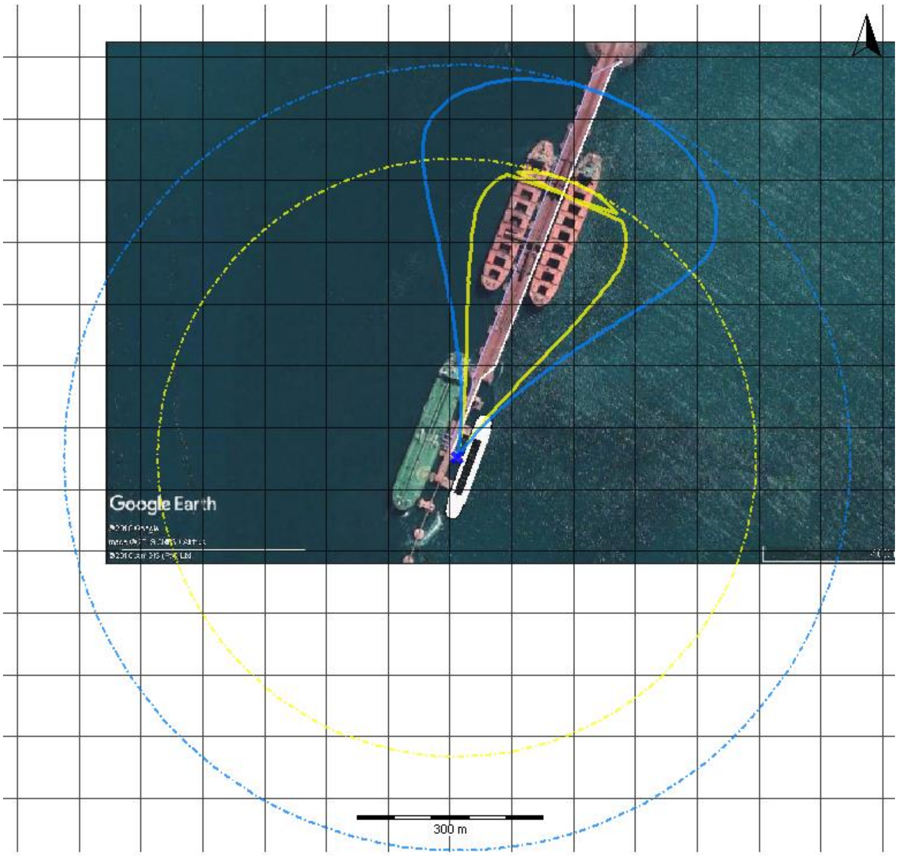
Flash fires as a result of catastrophic leaks at the jetty and at the cross-country pipeline were modelled with a 1.5m/s wind and the results are as follows:

Jetty Loading Arm/Hose	Max LFL Distance	Max ½LFL Distance
Crude Oil Loading Arm Rupture	77m	135m
LPG Loading Hose Rupture	484m	636m

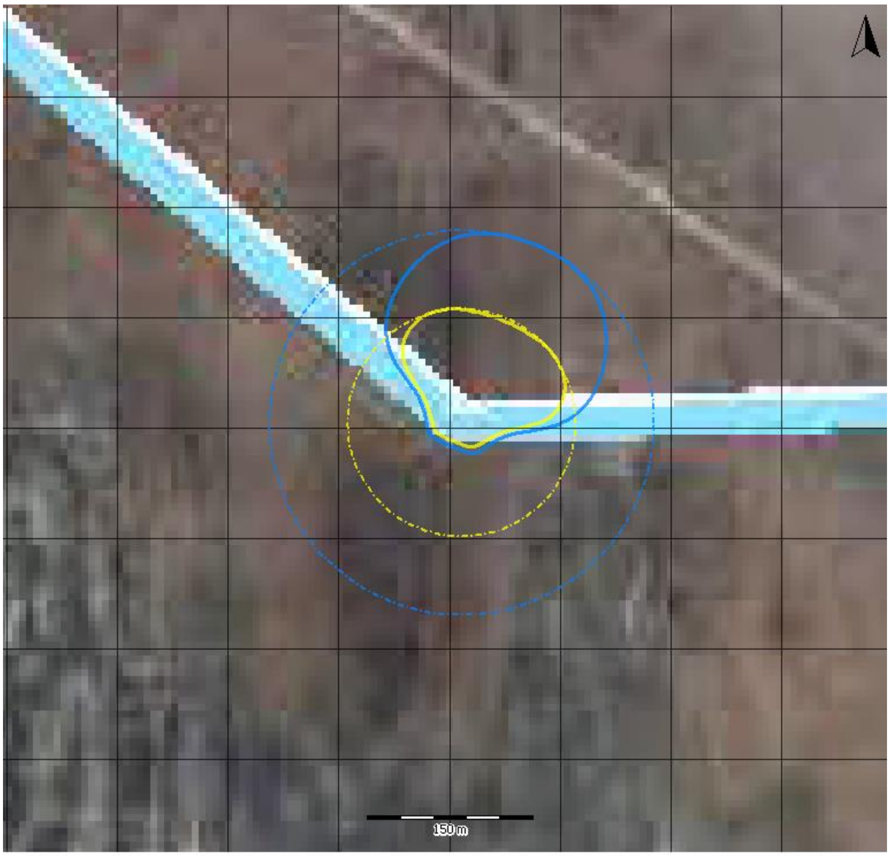
Pipeline	LFL Distance	½LFL Distance
Crude Oil Pipeline or Pig Trap Rupture	99m	174m
LPG Pipeline Rupture	112m	311m

For flash fires the LFL (yellow contour) and ½LFL (blue contour) are shown as follows:

Flash Fire as the result of a Crude Oil Load Arm Rupture



Flash Fire as the result of an LPG Loading Hose Rupture



Flash Fire as the result of a Crude Oil Pipeline Rupture

Flash Fire as the result of an LPG Pipeline Rupture

d) Vapour Cloud Explosions

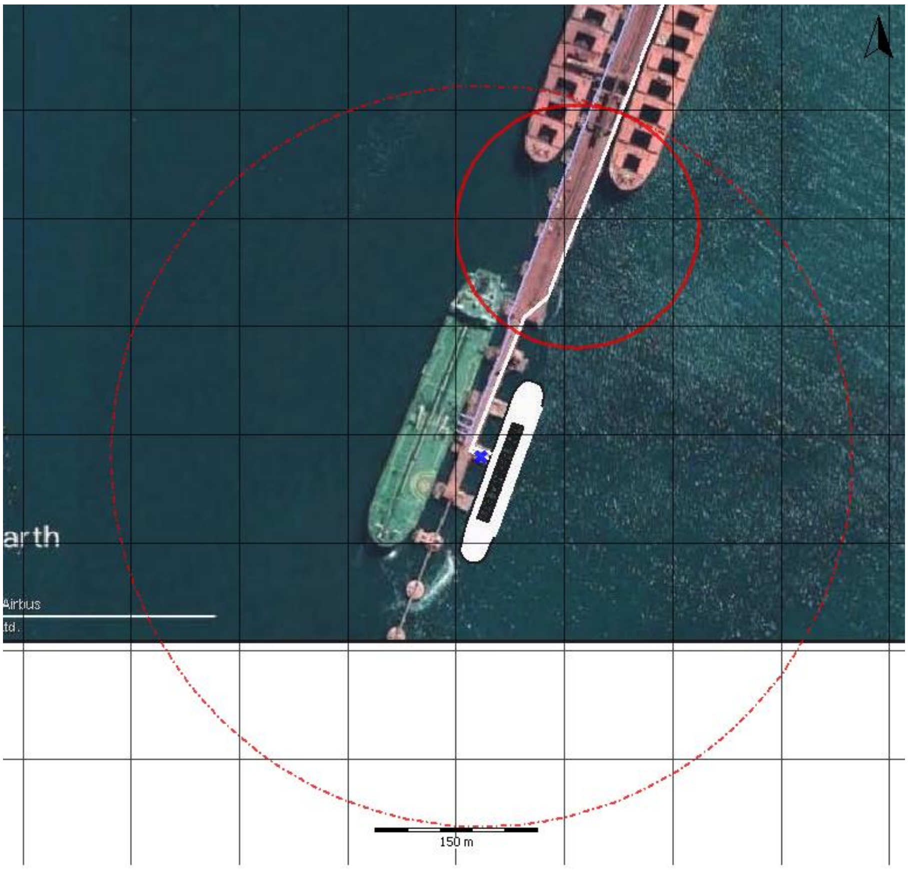
Flash fires as a result of catastrophic leaks at the jetty and at the cross-country pipeline were modelled with a 1.5m/s wind and the 10 kPa blast overpressure are as follows:

Jetty Loading Arm/Hose	Max Distance	Min Distance
Crude Oil Loading Arm Rupture	60m	3m
LPG Loading Hose Rupture	342m	118m

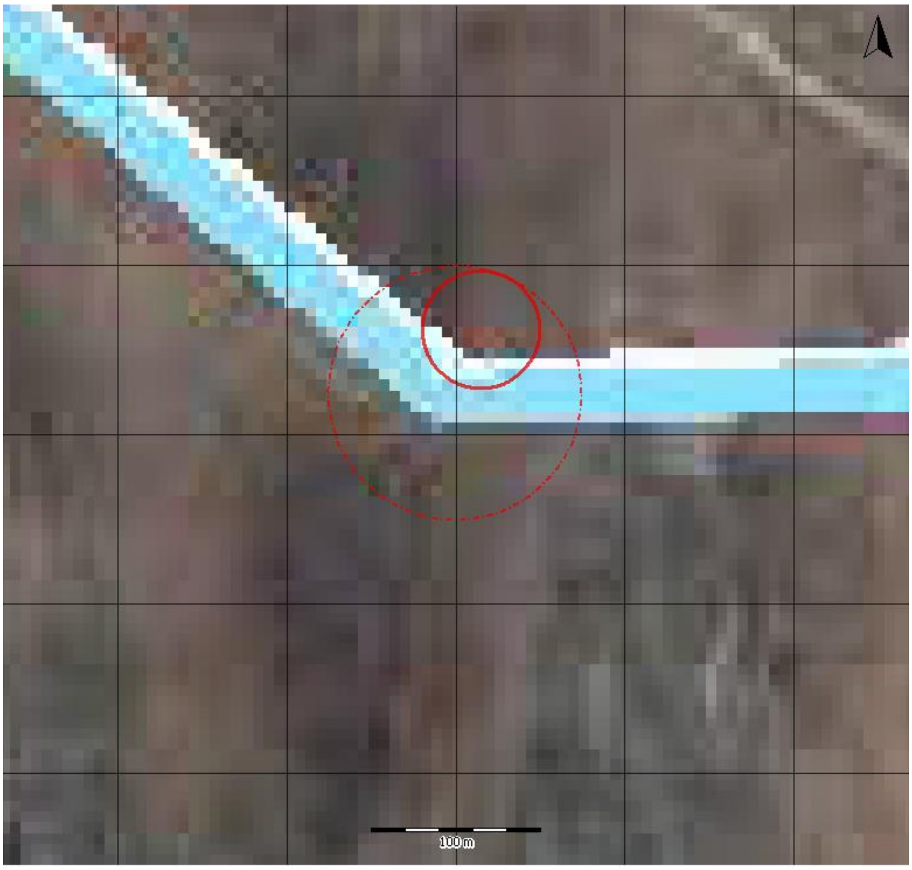
Pipeline	Max Distance	Min Distance
Crude Oil Pipeline or Pig Trap Rupture	75m	6m
LPG Pipeline Rupture	84m	29m

For vapour cloud explosions the 10kPa (red contour) are shown as follows:

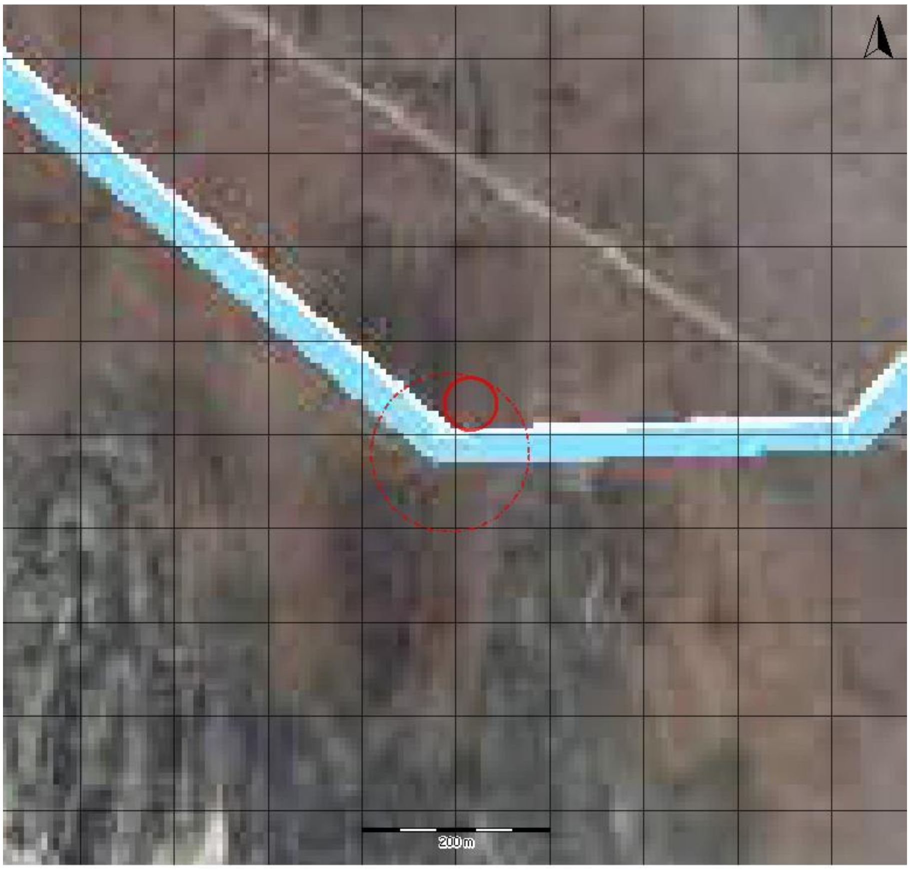
VCE as the result of a Crude Oil Load Arm Rupture



VCE as the result of an LPG Loading Hose Rupture



VCE as the result of a Crude Oil Pipeline Rupture



VCE as the result of an LPG Pipeline Rupture

8.10.5. Crude Oil Tank Farm

The following are the consequences of pool fires, jet fires, flash fires and explosions at the crude oil tank farm.

a) Pool Fires

At the crude oil tank farm, the loading pumps are located within a bunded area. A loss of containment at the pump station would result in a pool. The pool will be a contained pool in the bunded area.

On ignition of a contained flammable pool, the fire would extend to the limit of the pool but would shrink rapidly as the fuel within the pool is consumed.

The following mitigation will be implemented to deal with consequences of a pool fire at the pump station:

- Construction of a bund around the pump station;
- Use of foam canons.

A tank-top fire occurs when the flammable vapours above the stored liquid ignites. The resulting fire is contained within the tank. A faulty Pressure Valve could cause too much oxygen in the tank and a lightning strike for example could start a tank top fire inside the tank.

The following mitigation has been implemented to deal with prevention of a tank top fire:

- Installation of lightning protection;

The consequences of a pool fire are as follows:

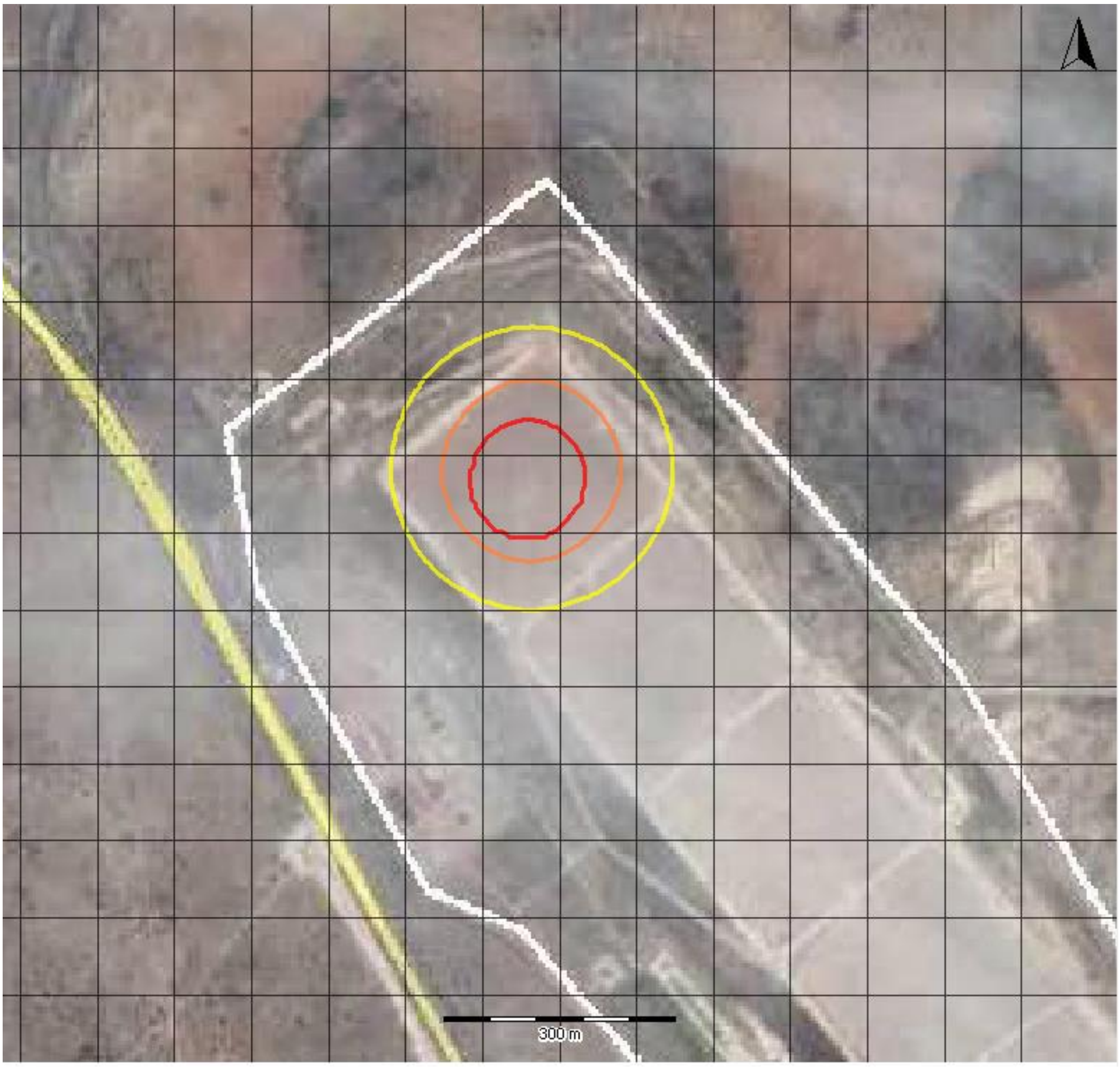
Crude Oil Pump Station			
Pump Leak	**Radiation Contour 37.5kW/m^2**	**Radiation Contour 12kW/m^2**	**Radiation Contour 4.5kW/m^2**
30mm Leak	22m	39m	60m

Tank Top Fire			
Pool Fire	**Radiation Contour 37.5kW/m^2**	**Radiation Contour 12kW/m^2**	**Radiation Contour 4.5kW/m^2**
Lightning Strike	77m	123m	191m

Thermal radiation from a pool fires at the crude oil tank farm are shown below.

- 4.5 kW/m^2 is the radiation that would cause pain and second degree burns within 20 seconds. (Yellow Contour)
- 12.5 kW/m^2 is the energy required for pilot ignition of wood. (Orange Contour)
- 37.5 kW/m^2 indicates the lower limit of damage to steel equipment and represents a 100% fatality for people exposed to the flame. (Red Contour)

Crude Oil Pool Fire as a result of a Pump Leak



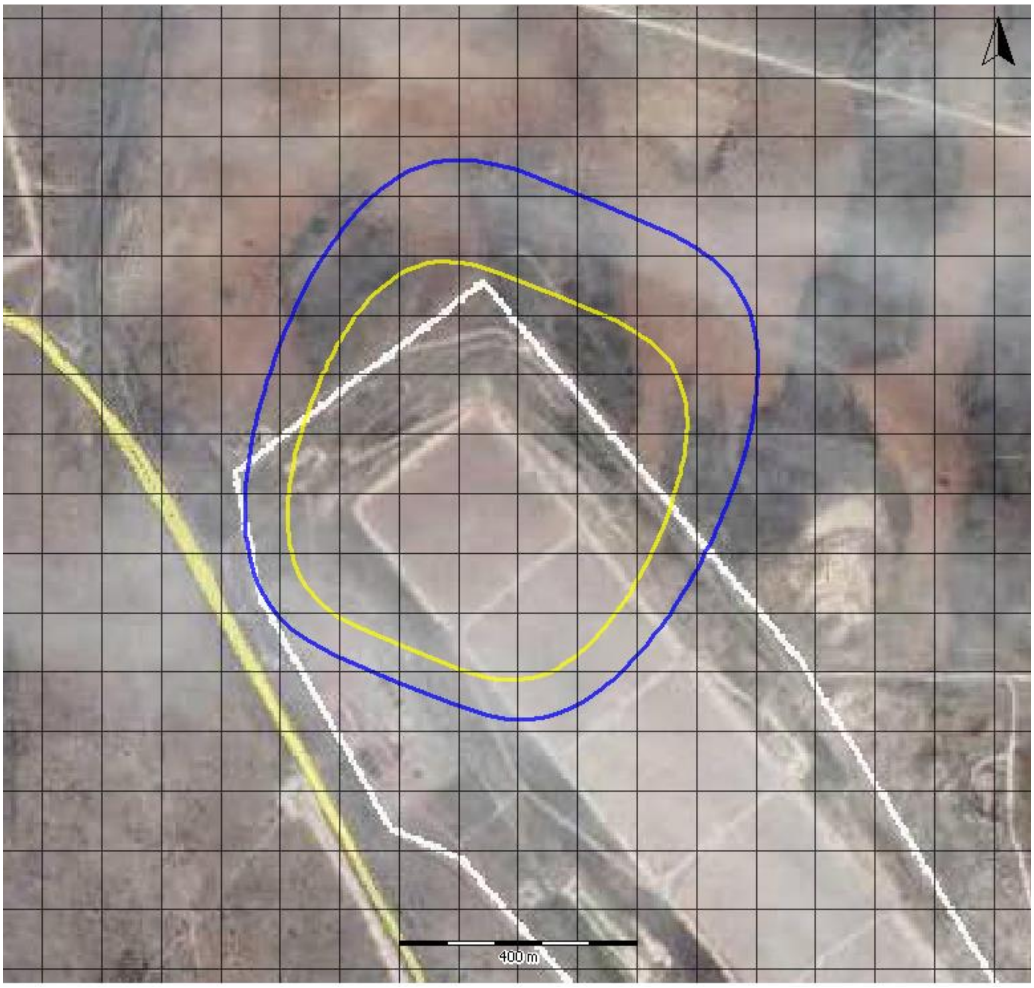
Crude Oil Tank Top Fire

b) Flash Fires

Flash fires as a result of a faulty Pressure Relief Valve during tank filling operations would result in a flammable cloud. The maximum extent of the flammable cloud for a single tank from all wind directions is shown below. Flash fires at the crude oil tank farm were modelled with a 1.5m/s wind and the results are as follows:

Tank Roof Failure	Max LFL Distance	Max ½LFL Distance
Flash Fire	77m	135m

For flash fires the LFL (yellow contour) and ½LFL (blue contour) are shown as follows:



Flash Fire as the result of a Tank Roof Failure

c) Vapour Cloud Explosions

A loss of containment of crude oil with an ignition source could form a flash fire, or a vapour cloud explosion. The blast overpressures of a single tank from the release of flammable vapours was modelled.

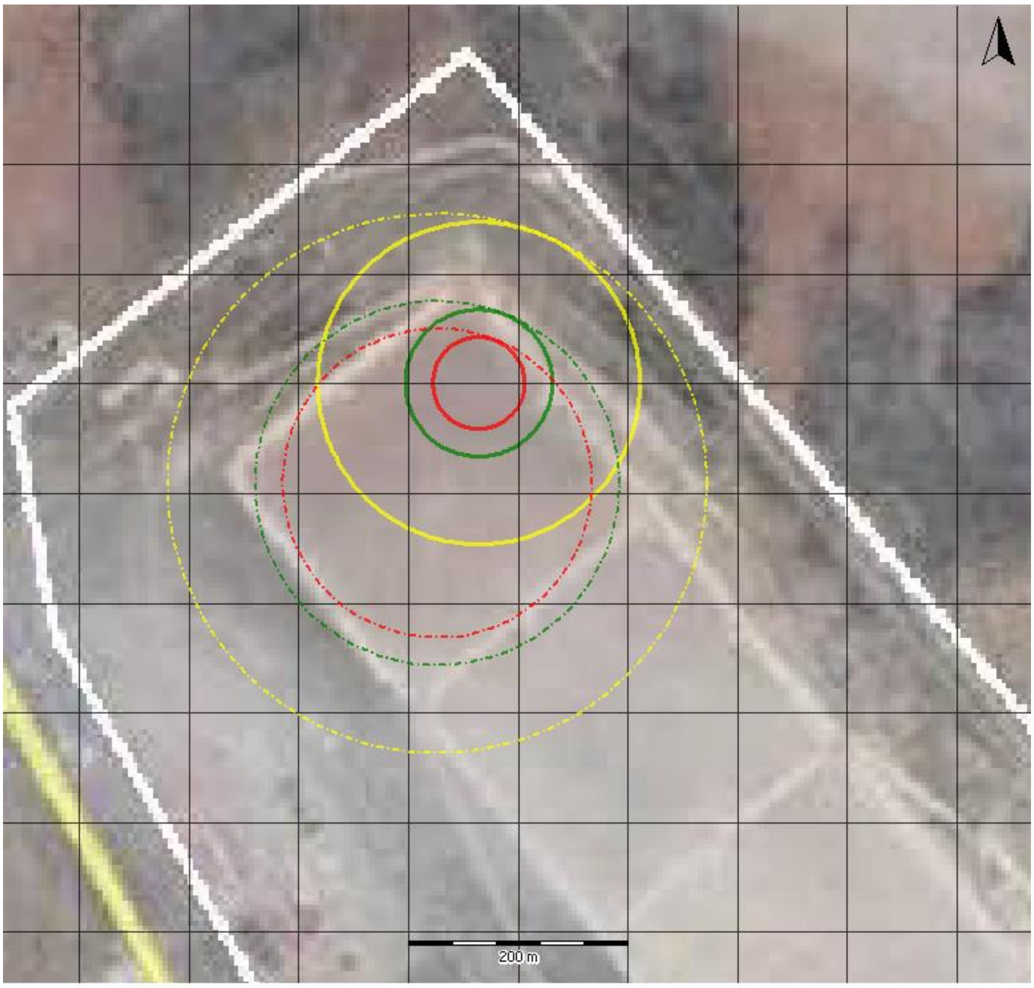
For a fixed roof tank explosion, the worst-case scenario would be when the tank is empty and flammable vapours fill the entire available space within the tank. The mass used in the explosion calculations is the volume of flammable material at its lower flammable limit. In this case, the material used for the explosion was butane with a calculated mass of 36526 kg.

For a large mass of explosive material to be included in the explosion is when the tanks have been empty and unused for a period of time and there is a large amount of air within the tank and a potential static build up, with liquid being introduced above liquid level. A good re-commissioning procedure would mitigate this. During continuous operation, nitrogen would be introduced to maintain the pressure in the tanks without introducing air. This would limit the amount of air in the vessel at any time and reduce the explosion consequences.

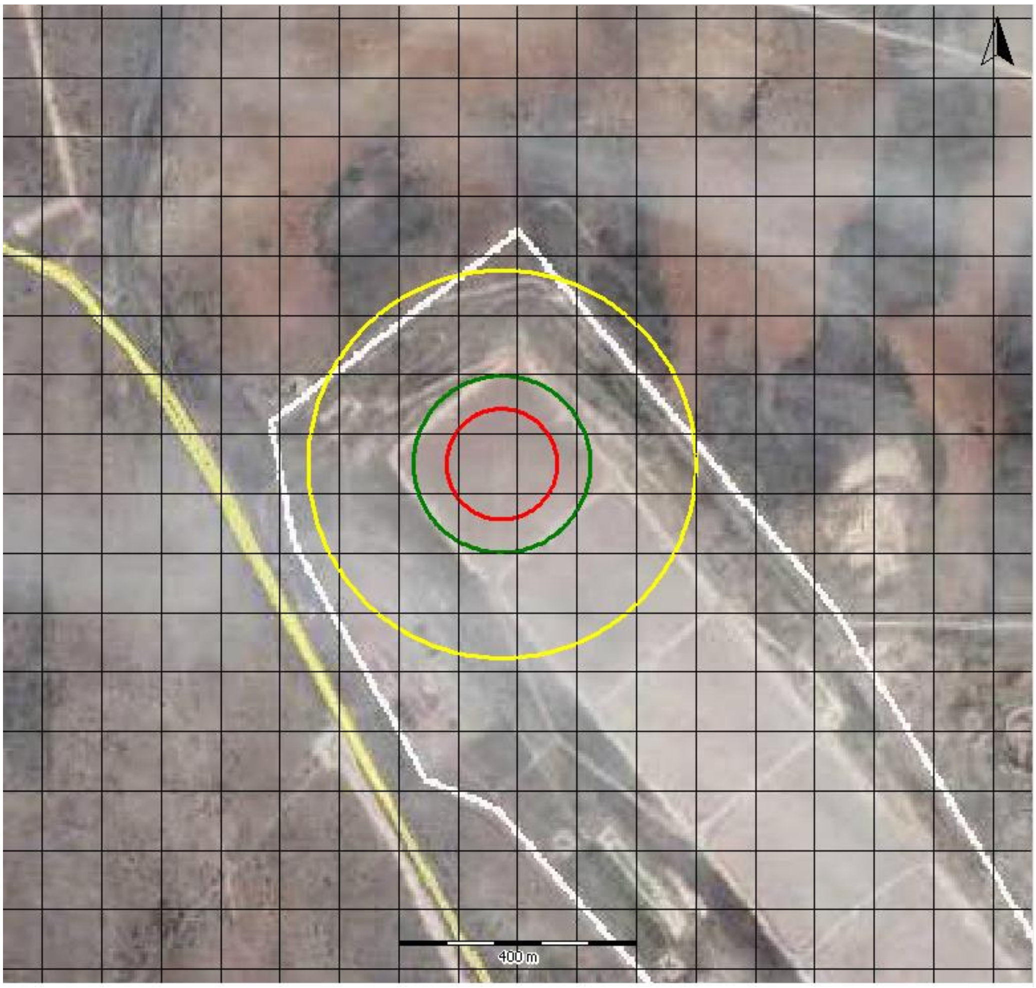
Explosions at the crude oil tank farm were modelled with a 1.5m/s wind and the results are as follows:

Crude Oil Tank Farm	10kPa		30kPa		70kPa	
	Max Distance	Min Distance	Max Distance	Min Distance	Max Distance	Min Distance
Crude Oil Loss of containment	246m	-48m	166m	32m	141m	57m
Fixed Roof Tank Explosion	326m	-326m	149m	-149	93m	-93m

For vapour cloud explosions the 10kPa (red contour), 30kPa (green contour) and 70kPa (yellow contour) are shown as follows:



VCE as the result of a Crude Oil Loss of Containment



Fixed Roof Tank Explosion

8.10.6. Crude Oil Terminal Flare LPG Installation

The following are the consequences of pool fires, jet fires, flash fires and explosions at the installation.

a) Pool Fires

A loss of containment of gas does not normally result in the formation of a flammable pool. Pool fires were not modelled.

b) Jet Fires

For this Assessment, jet fires from a 10mm leak, a pipe rupture and a PRV failure was assumed. The worst-case scenario of the jet fire being horizontal and in the same direction of the 1.5m/s wind was assumed except for the PRV.

Thermal radiation from jet fires is shown below.

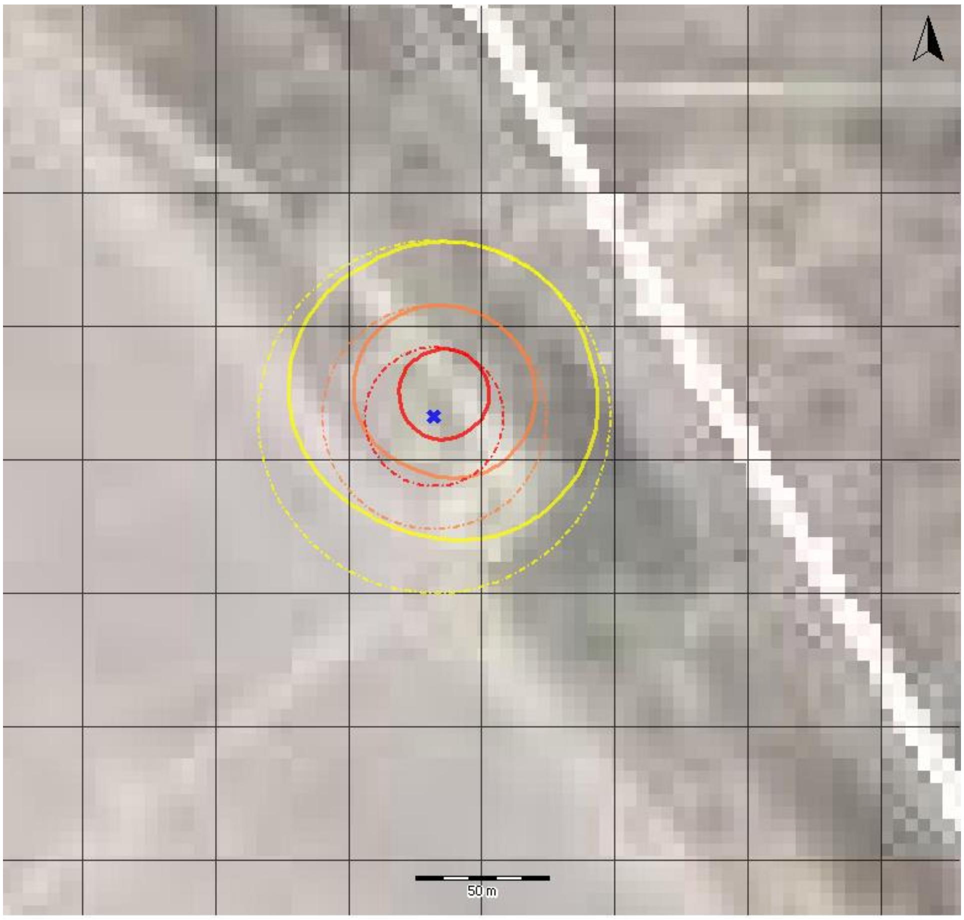
- 4.5 kW/m^2 is the radiation that would cause pain and second degree burns within 20 seconds. (Yellow Contour)
- 12.5 kW/m^2 is the energy required for pilot ignition of wood. (Orange Contour)
- 37.5 kW/m^2 indicates the lower limit of damage to steel equipment and represents a 100% fatality for people exposed to the flame. (Red Contour)

The following mitigation is currently installed and implemented to deal with consequences of an LPG release at the installation:

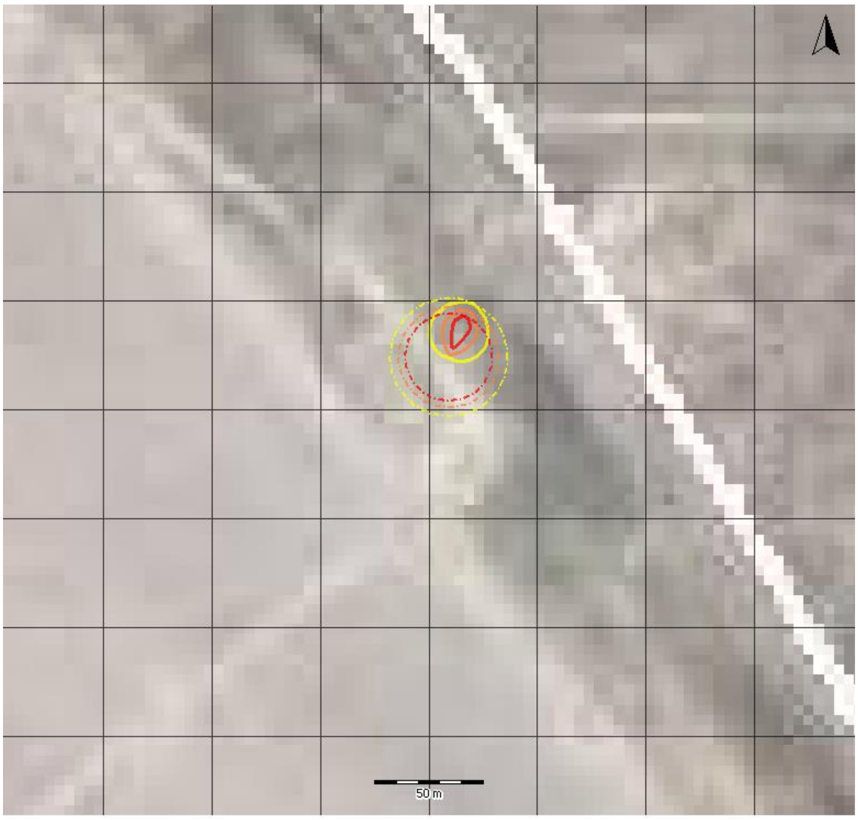
- All fire-fighting equipment will be in place;
- Emergency stop switch installed to stop pumps;
- Gas detection;
- Regular Maintenance;

The consequences of jet fires are as follows:

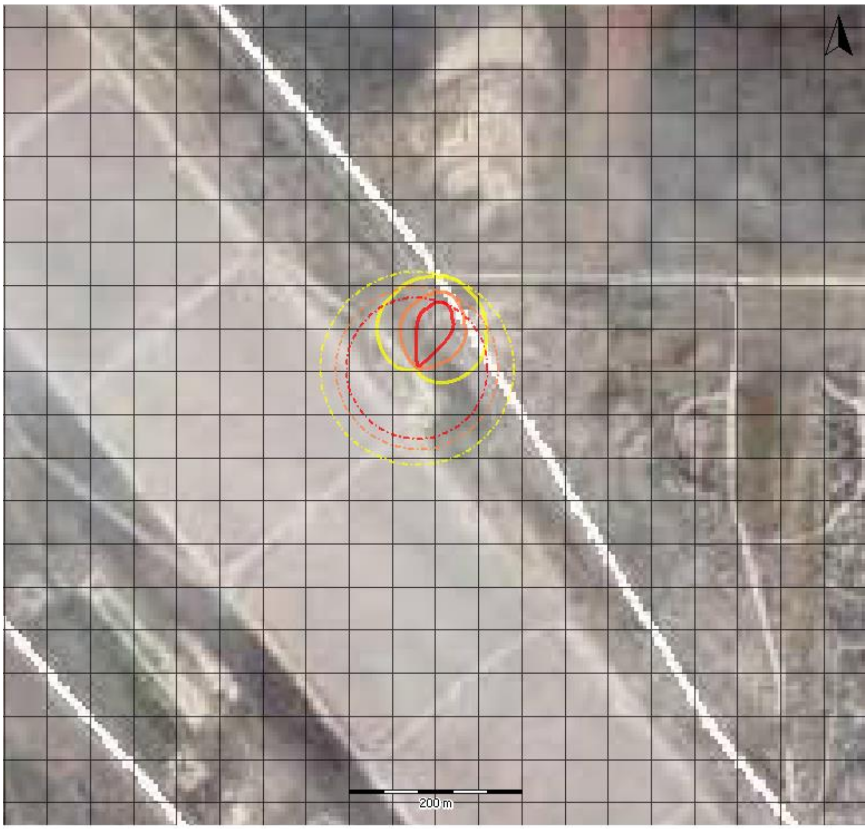
LPG Jet Fires				
Scenario	**Flame Length**	**Radiation Contour 37.5kW/m²**	**Radiation Contour 12kW/m²**	**Radiation Contour 4.5kW/m²**
PRV Failure	26.12m	26m	42m	66m
Loading Hose Rupture	67.64m	82m	94m	112m
10mm Loading Hose Leak	17.48m	20m	23m	27m



50mm Jet Fire PRV Failure



Jet Fire from a 10mm Loading Hose Leak



Jet Fire from a Loading Hose Rupture

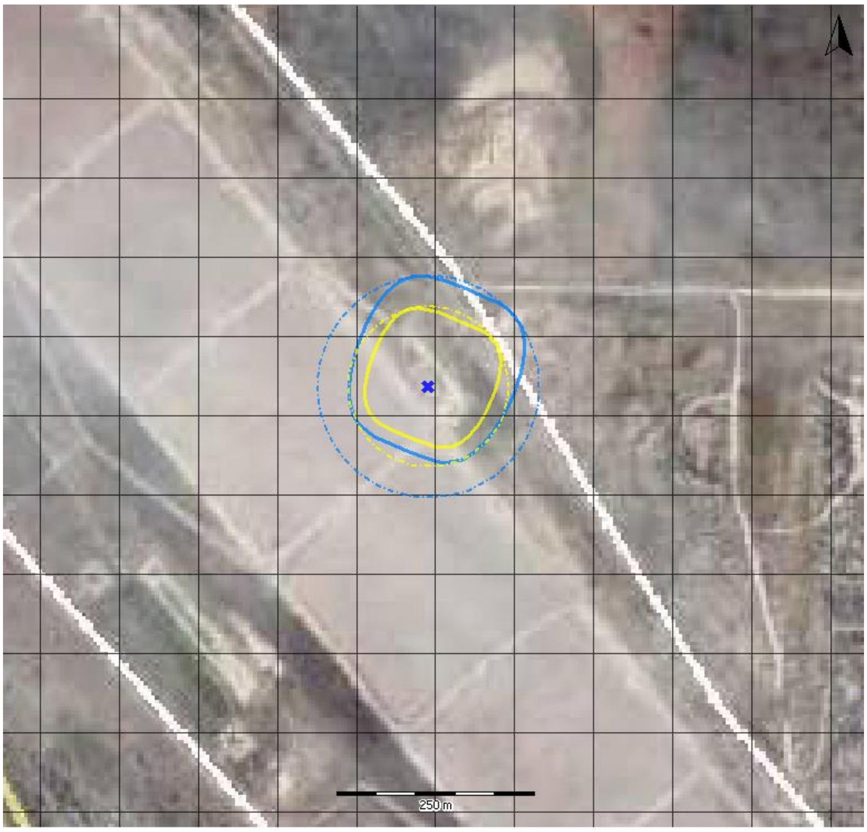
c) Flash Fires

Flash fires as a result of catastrophic leaks and vessel failures were modelled with a 1.5m/s wind and the results are as follows:

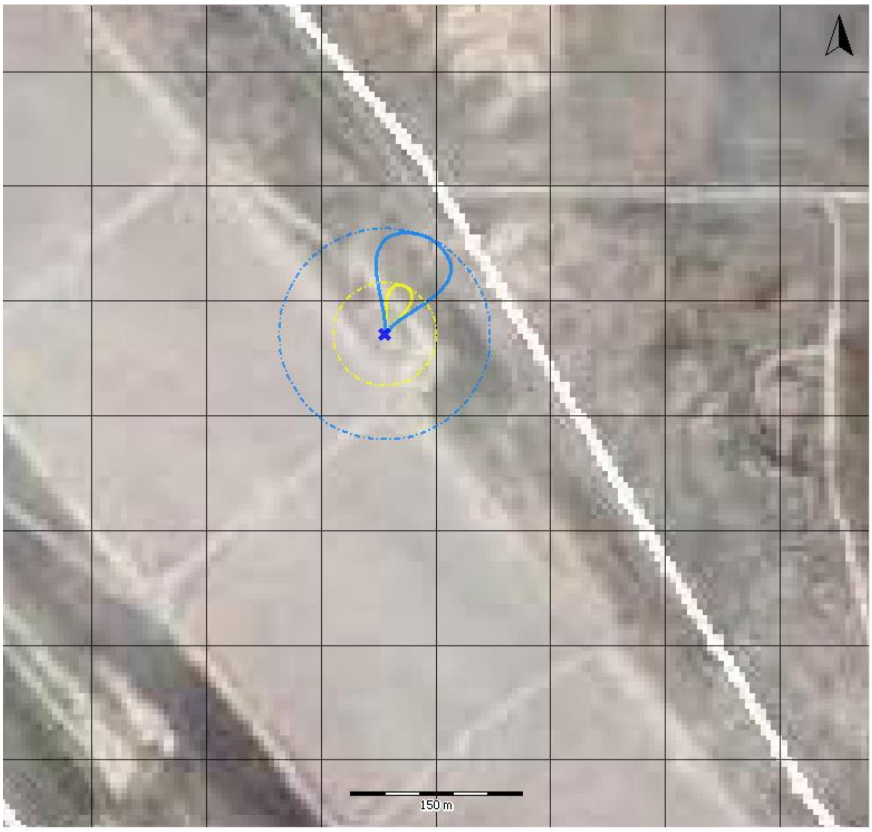
Scenario	Max LFL Distance	Max ½LFL Distance
Road Tanker Loading Hose Rupture	120m	161m
LPG Vessel Catastrophic Failure	92m	132m
LPG Vessel Empty in 10 min	45m	92m
Road Tanker Catastrophic Failure	189m	275m

For flash fires the LFL (yellow contour) and ½LFL (blue contour) are shown as follows:

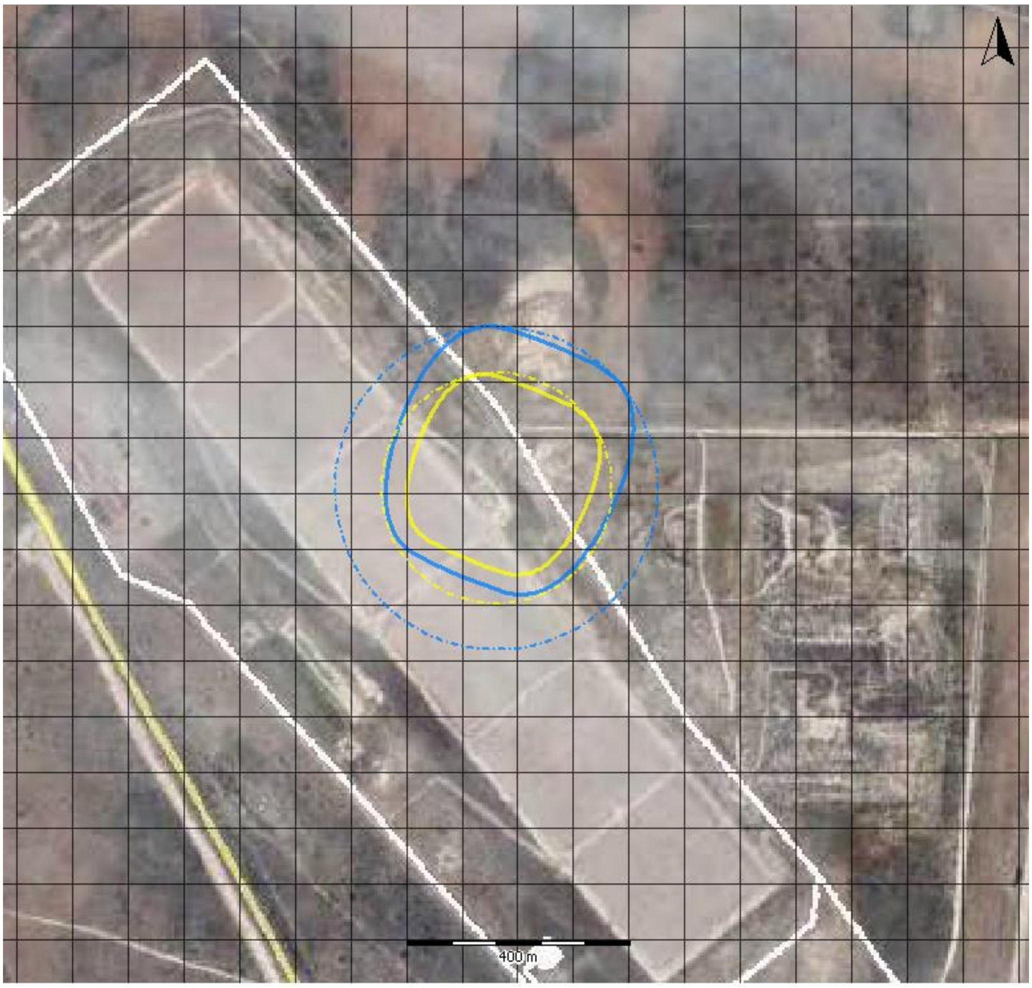
Flash Fire as the result of an LPG Loading Hose Rupture



Flash Fire as the result of an LPG Vessel Catastrophic Failure



Flash Fire as the result of an LPG Vessel Emptying in 10 minutes



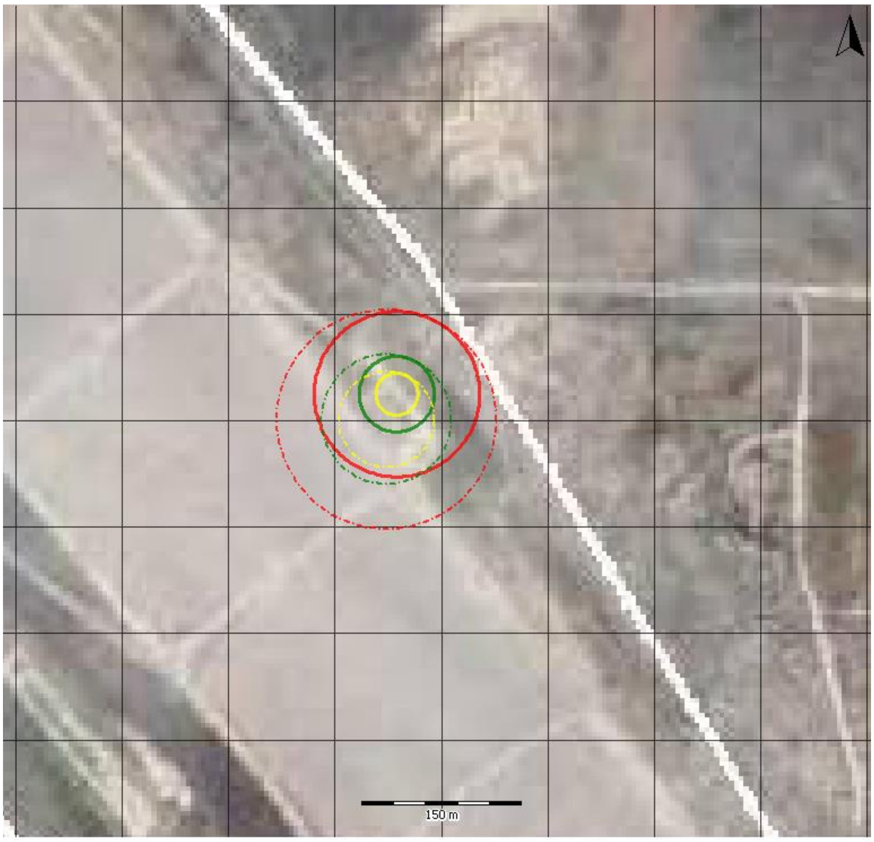
Flash Fire as the result of an LPG Road Tanker Failure

d) Vapour Cloud Explosions

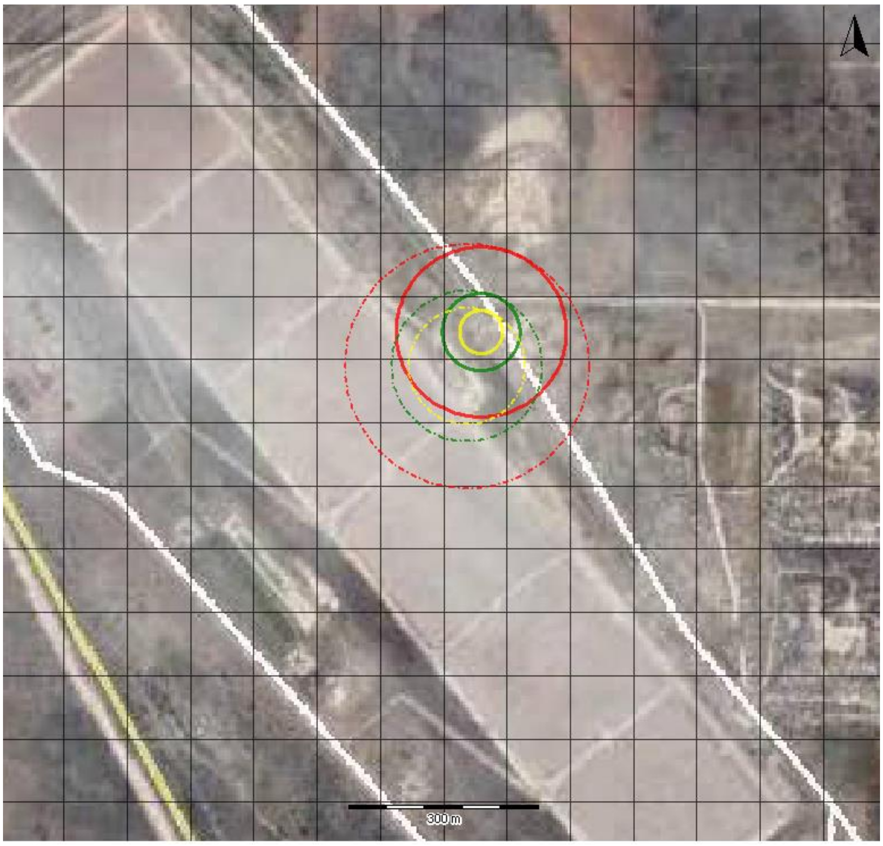
Explosions at the crude oil tank farm were modelled with a 1.5m/s wind and the results are as follows:

Flare LPG Installation	10kPa		30kPa		70kPa	
	Max Distance	**Min Distance**	**Max Distance**	**Min Distance**	**Max Distance**	**Min Distance**
LPG Vessel Catastrophic Failure	45m	6m	61m	-10m	103m	-52m
LPG Road Tanker Catastrophic Failure	326m	-326m	149m	-149	93m	-93m

For vapour cloud explosions the 10kPa (red contour), 30kPa (green contour) and 70kPa (yellow contour) are shown as follows:



VCE as the result of an LPG Vessel Catastrophic Failure



VCE as the result of an LPG Road Tanker Failure

e) **Boiling Liquid Expanding Vapour Explosion (BLEVE)**

The major consequences of a BLEVE are the intense thermal radiation from the fireball, a blast wave and fragments from the shattered vessel. These fragments may be projected to considerable distances. A blast wave from a BLEVE is fairly localised, but can cause significant damage to immediate equipment. In the case of a BLEVE formed from the LPG vessel, the radius of the fireball is estimated to be 93,13m. The vessel at this facility has a deluge system that will reduce the risk considerably.

The following figure shows the fireball diameter of a BLEVE in red.

LPG Vessel BLEVE

8.10.7. Crude Oil Terminal Nitrogen Installation

The consequences for the nitrogen installation was not modelled because the risks of a loss of containment of nitrogen would not have an impact on the on or off-site risks.

8.10.8. LPG Bulk Import Terminal

The following are the consequences of pool fires, jet fires, flash fires and explosions at the installation.

a) Pool Fires

A loss of containment of gas does not normally result in the formation of a flammable pool. Pool fires were not modelled.

b) Jet Fires

For this Assessment, jet fires from a 10mm leaks and pipe ruptures was assumed. The worst-case scenario of the jet fire being horizontal and in the same direction of the 1.5m/s wind was assumed except for the PRV.

Thermal radiation from jet fires is shown below.

- 4.5 kW/m^2 is the radiation that would cause pain and second degree burns within 20 seconds. (Yellow Contour)
- 12.5 kW/m^2 is the energy required for pilot ignition of wood. (Orange Contour)
- 37.5 kW/m^2 indicates the lower limit of damage to steel equipment and represents a 100% fatality for people exposed to the flame. (Red Contour)

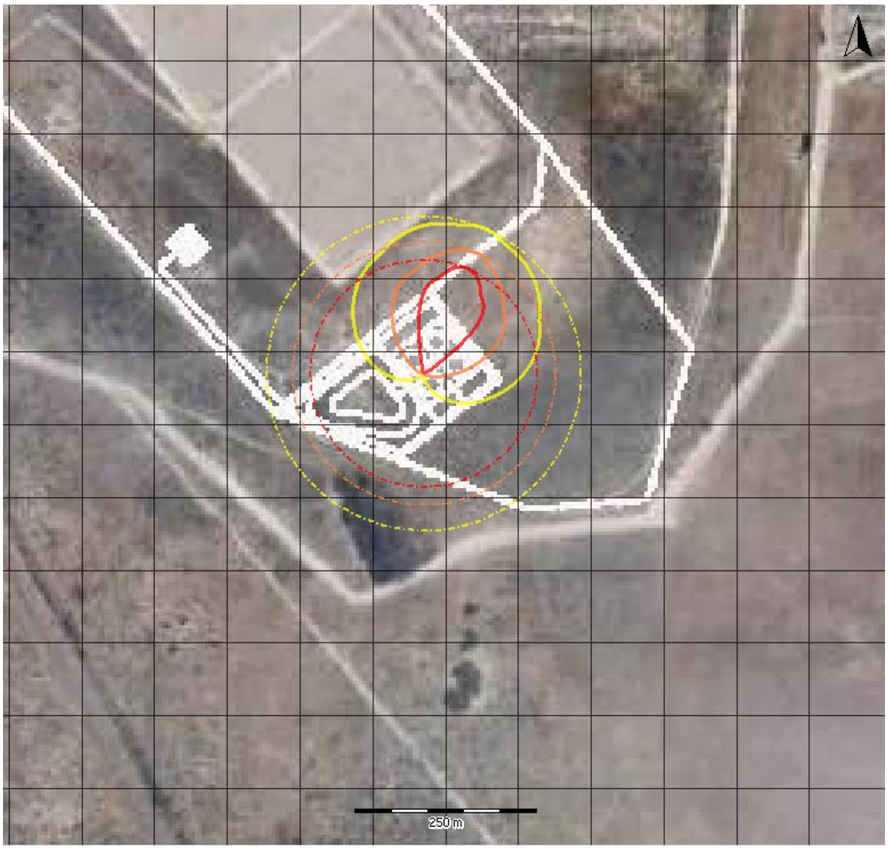
The following mitigation is currently installed and implemented to deal with consequences of an LPG release at the installation:

- All fire-fighting equipment will be in place;
- Emergency stop switch installed to stop pumps;
- Gas detection;
- Regular Maintenance;

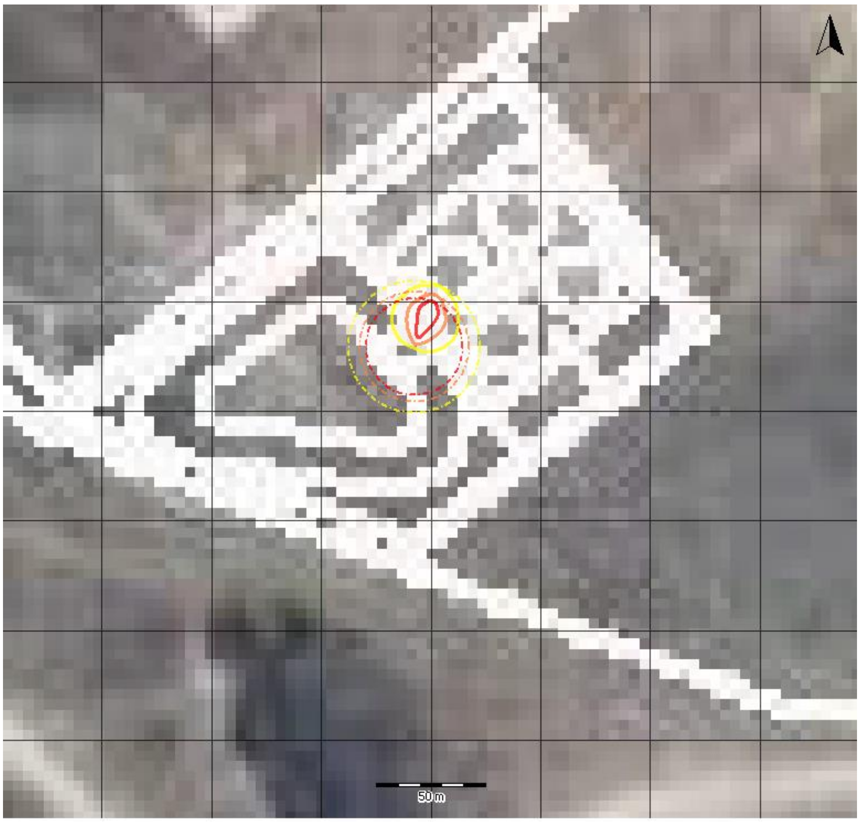
The consequences of jet fires are as follows:

LPG Jet Fires				
Scenario	**Flame Length**	**Radiation Contour 37.5kW/m^2**	**Radiation Contour 12kW/m^2**	**Radiation Contour 4.5kW/m^2**
Distribution Pipe 10mm Leak	19.35m	22m	25m	30m
Distribution Pipe Rupture	126.37m	156m	181m	216m
Loading Hose Rupture	126.37m	156m	181m	216m
10mm Loading Hose Leak	19.35m	22m	25m	30m

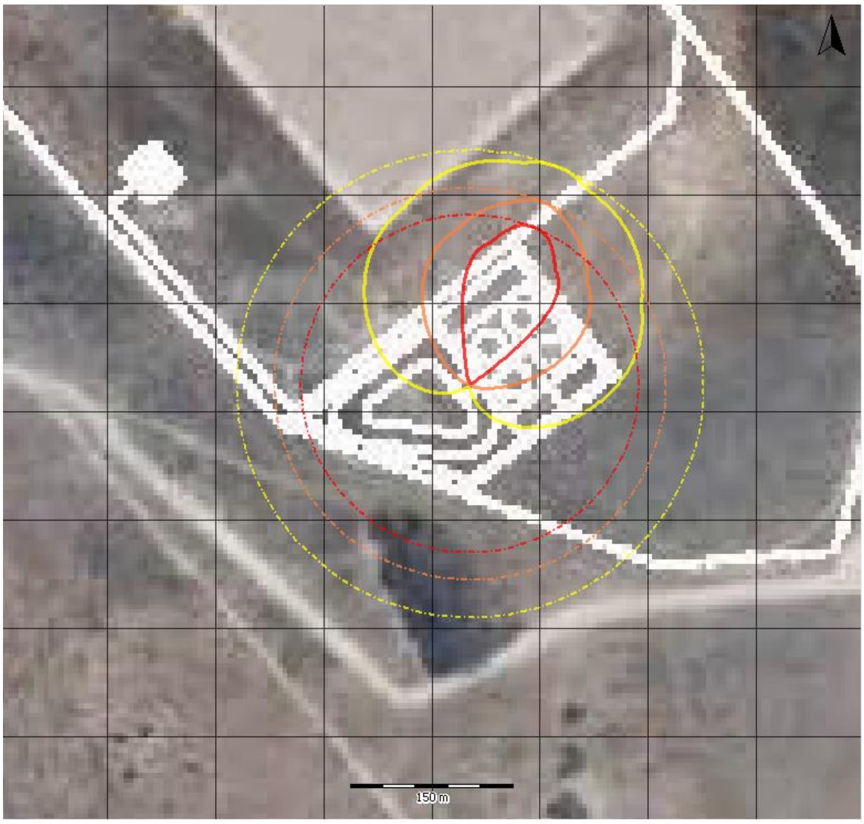
10mm Jet Fire Pipe Leak



Jet Fire from a Pipe Rupture



Jet Fire from a Loading Arm Leak



Jet Fire from a Loading Arm Rupture

c) Flash Fires

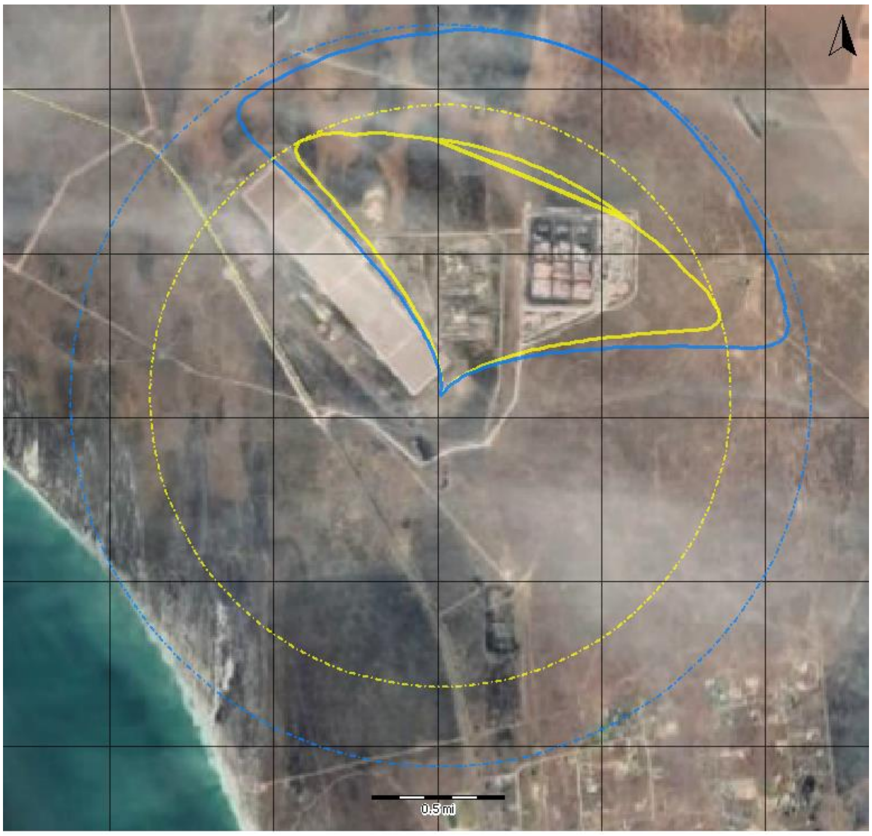
Flash fires as a result of catastrophic leaks and vessel failures were modelled with a 1.5m/s wind and the results are as follows:

Scenario	Max LFL Distance	Max ½LFL Distance
Road Tanker Loading Arm Rupture	220m	277m
LPG Vessel Catastrophic Failure	1934m	2947m
LPG Vessel Empty in 10 min	1515m	2256m
Road Tanker Catastrophic Failure	189m	275m

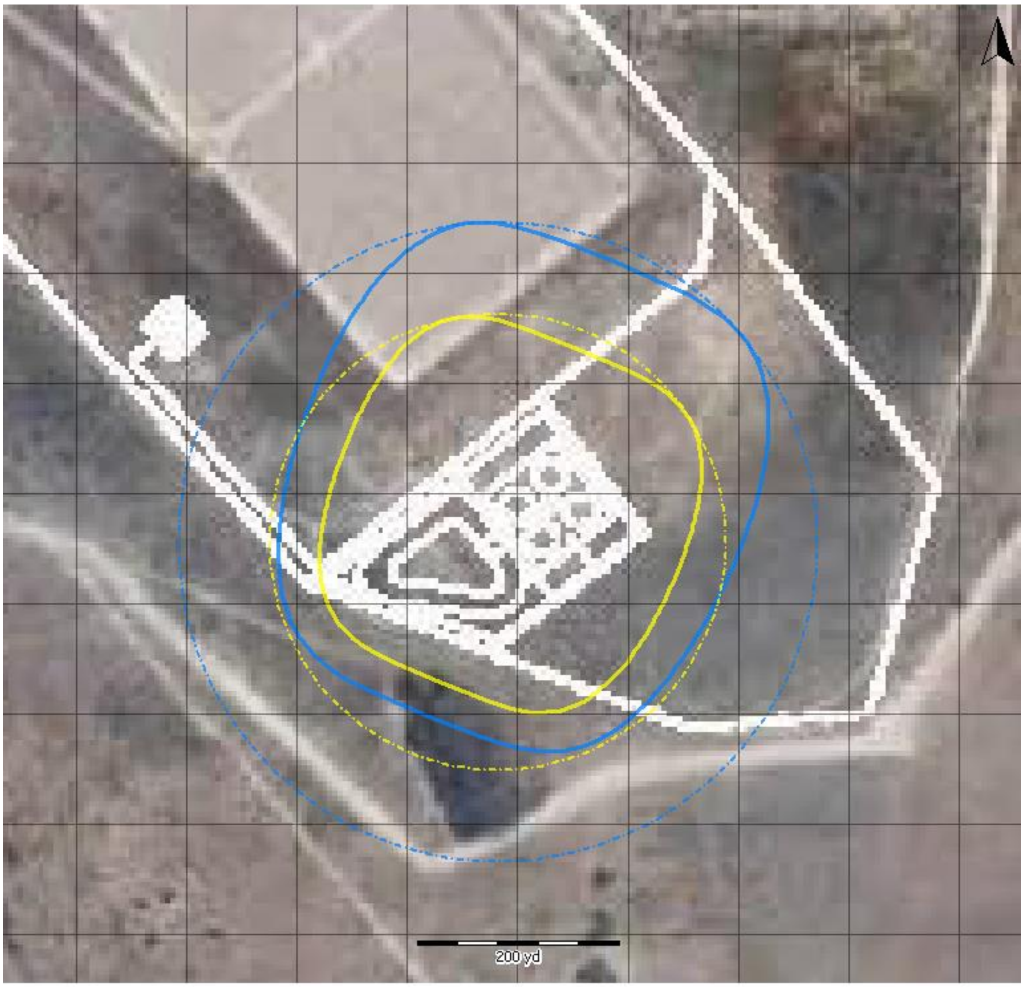
For flash fires the LFL (yellow contour) and ½LFL (blue contour) are shown as follows:

Flash Fire as the result of an LPG Loading Arm Rupture

Flash Fire as the result of an LPG Vessel Catastrophic Failure



Flash Fire as the result of an LPG Vessel Emptying in 10 minutes



Flash Fire as the result of an LPG Road Tanker Failure

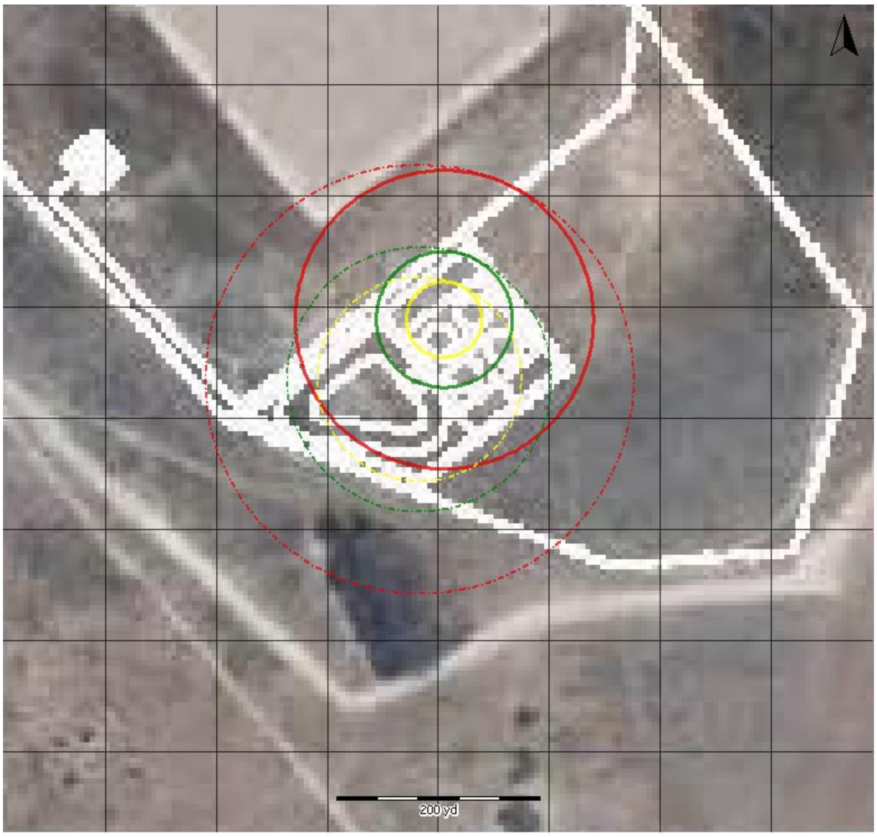
d) Vapour Cloud Explosions

Explosions at the crude oil tank farm were modelled with a 1.5m/s wind and the results are as follows:

LPG Installation	10kPa		30kPa		70kPa	
	Max Distance	Min Distance	Max Distance	Min Distance	Max Distance	Min Distance
LPG Vessel Catastrophic Failure	1581m	-83m	1128m	370m	960m	538m
LPG Road Tanker Catastrophic Failure	193m	-76m	119m	-3m	92m	24m

For vapour cloud explosions the 10kPa (red contour), 30kPa (green contour) and 70kPa (yellow contour) are shown as follows:

VCE as the result of an LPG Vessel Catastrophic Failure



VCE as the result of an LPG Road Tanker Failure

e) **Boiling Liquid Expanding Vapour Explosion (BLEVE)**

The major consequences of a BLEVE are the intense thermal radiation from the fireball, a blast wave and fragments from the shattered vessel. These fragments may be projected to considerable distances. A blast wave from a BLEVE is fairly localised, but can cause significant damage to immediate equipment. In the case of a BLEVE formed from the LPG road tanker, the radius of the fireball is estimated to be 157,12m. The vessel at this facility has a deluge system that will reduce the risk considerably. The bulk spheres will be mounded so a BLEVE was not modelled.

The following figure shows the fireball diameter of a BLEVE in red.

LPG Road Tanker BLEVE

8.11. Potential Onsite and Offsite Domino Effects

Fires and explosions from the road tankers could impact on the bulk vessels. Fires from the bulk vessels would cause a domino effect on the road tankers. The LPG installations could impact on the crude oil tanks.

9. FREQUENCY ANALYSIS

9.6. Site Specific (Final) Frequencies

The frequencies indicated below are generic frequencies as specified in *BEVI*. Site specific frequencies are calculated utilising these generic frequencies as a base. The final frequency calculations are included in the appendices.

9.7. Generic Equipment Failure Scenarios

The main hazard when storing toxics and flammables is the loss of containment, which when/if ignited may result in a fire or an explosion. In the absence of ignition, the flammable vapour cloud would move with the wind until the effects of dispersion dilute the vapours below the flammable concentration. A loss of containment of flammables may occur during delivery, storage or distribution. The possible hazards are to be identified, together with the failure modes and the possible initiating events that may cause such a failure. Failure rates were obtained from *'RIVM - Reference Manual Bevi Risk Assessments'.*

9.8. Buncefield Scenarios

Buncefield scenarios refer to tank overfilling and a substantial transfer rate. According to the UK HSE the Buncefield scenarios must be included for all tanks higher than 5m and with a transfer rate in excess of 100m^3 per hour.

The Buncefield scenario assumes that all alarms and emergency shut-off fail and that the product is pumped out through the vents located at the top of the tank. The product spills over the edge which causes a vapour cloud that could lead to a vapour cloud explosion.

The consequences from an overfill are as follows:

- Pool Fire;
- Vapour Cloud Explosion;
- Flash Fire.

The failure frequency is calculated as follows:

Overfill Frequency = BF x MSQ x LG x ASD x ASTFO
where:
BF = Base Frequency (1.0 x 10E-4)
MSQ = Management System Quality
LG = Level Gauging
ASD = Automatic Shut Down
ASTFO = Aboveground Storage Tank Fill Operations

9.9. Blocking Systems

Blocking systems are used to limit the released quantity following a loss of containment. A blocking system consists of a detection system, for example gas detection, combined with shut-off valves. The shut-off valves can be closed automatically or manually. The effectiveness of a blocking system is determined by various factors, such as the position of gas detection monitors and their distribution throughout the various wind directions. Furthermore, the detection limit and the response time of the system as well as the operator's intervention time are also relevant.

The following conditions must be met to include the operation of a blocking system in the risk analysis:

- An automatic detection system must be present that results in signalling within the control room or automatic control of the blocking valves. An example of this is a gas detection system with sufficiently sensitive monitors and adequate detection points. In the case of signalling in the control room this room must be continuously staffed.
- The detection system and the shut-off valves must be regularly tested.

The default values specified here for three representative systems were used as a guideline:

1. Automatic blocking system

An automatic blocking system is a system in which the detection of the leak and the closing of the blocking valves take place automatically. Action by an operator is not necessary. The introduction of a Safety Integrity Level system (SIL) would be seen as an automatic blocking system.

2. Semi-automatic blocking system

A semi-automatic blocking system is a system in which the detection of the leak takes place automatically and leads to an alarm signal in a continuously staffed control room. After validation of the signal the operator closes the blocking valves by actuating a switch in the control room. The probability of failure per operation is equal to 0.01, the time required for closing the blocking valves is equal to 10 minutes.

3. Non-automated blocking system

A non-automated blocking system is a system in which the detection of the leak takes place automatically and leads to an alarm signal in a continuously staffed control room. The operator does not have the facilities to shut off the blocking valves by actuating a switch in the control room but must take action outside the control room. For such a system the time required to effectively perform the required actions is so long that there is no effect on the QRA, given the maximum duration of an outflow of 30 minutes that is generally applied.

The crude oil import facility including the jetty, pipelines and tank farm operates on a SIL 1 rated system. The LPG import facility including the jetty, pipeline and the terminal will operate on a SIL 2 rated system.

9.10. Pressure Vessels

The scenarios considered under this category are partial failure and catastrophic failure. Factors that have been identified as influencing the integrity of cylinders are related to design, inspection, maintenance, and corrosion.

- A pressure cylinder is a storage vessel that contains a fluid under a design pressure equal to, or greater than 50kPa.

The failure frequencies are as follows:

	Frequency (per annum)
Instantaneous release of entire contents	5 x 1.0e-7
Release of entire contents in 10 minutes in a continuous and constant stream	5 x 1.0e-7
Continuous release of contents from a hole with an effective diameter of 10mm	1x 1.0e-5

9.11. Atmospheric Tanks

The scenarios considered under this category are partial failure and catastrophic failure. Factors that have been identified as having an effect on the integrity of tanks are related to design, inspection, maintenance, and corrosion.

- An atmospheric vessel is a storage vessel that contains a fluid under a design pressure equal to or less than 50kPa.

The scenarios and failure frequencies for an atmospheric storage tank apply to tanks with welded stumps, mounting plates, pipe connections up to the first flange and instrumentation pipes.

The failure frequencies for above-ground atmospheric tanks are as follows:

	Frequency (per annum)
Instantaneous release of entire contents	5 x 1.0e-6
Release of entire contents in 10 minutes in a continuous and constant stream	5 x 1.0e-6
Continuous release of contents from a hole with an effective diameter of 10mm	1 x 1.0e-4

9.12. Valves

The failure frequency of valves is dependent on the valve and the leak size. The ratio of the leak size *(d)* to the valve size *(D)* should firstly be determined in order to determine the valve failure frequency per year, for example:

d/D	**Leak Frequency (per valve per year)**
0.1	1.4 x 1.0e-4
0.2	1.9 x 1.0e-4
0.5	2.5 x 1.0e-4
1.0	3.0 x 1.0e-4

9.13. Flanges

Pressure surge or significant deviations of pressure or temperature may cause a flanged joint to be over stressed, resulting in a small leak. Larger holes through to complete line fracture may conceivably result from mechanical impact or pressure surge. These events are likely to be detected more rapidly, resulting in a quicker isolation of the leak.

The flange failures per year vary greatly with the flange and gasket quality. A reasonable average based on current practices is summarised below:

Pipe Diameter (mm)	Equivalent Hole Size (mm)	Leak Frequency (per item per year)
100	5	1 x 1.0e-5
> 100	25	1 x 1.0e-6

9.14. Ignition Probability of Flammable Gases

9.14.4. Direct Ignition

The probability of direct ignition depends on the type of installation (stationary installation or transport unit), the substance category and the outflow quantity.

- Values for stationary installations are given in the table below;
- Values for transport units are given in the next table;
- Definition of the substance category is given in the third table.

Substance Category	Source Term Continuous	Source Term Instantaneous	Probability of Direct Ignition
Category 0 Average/High reactivity	<10 kg/s 10 – 100 kg/s >100 kg/s	<1000 kg 1000 – 10000 kg >10000 kg	0.2 0.5 0.7
Category 0 Low reactivity	<10 kg/s 10 – 100 kg/s >100 kg/s	<1000 kg 1000 – 10000 kg >10000 kg	0.2 0.4 0.9
Category 1	All flow rates	All quantities	0.065
Category 2	All flow rates	All quantities	0.01
Category 3, 4	All flow rates	All quantities	0

Substance Category	Transport Unit	Scenario	Probability of Direct Ignition
Category 0	Road tanker	Continuous	0.1
	Road tanker	Instantaneous	0.4
	Tank wagon	Continuous	0.1
	Tank wagon	Instantaneous	0.8
	Ships – gas tankers	Continuous, 180m^3	0.7
	Ships – gas tankers	Continuous, 90m^3	0.5
	Ships – semi gas tankers	Continuous	0.7
Category 1	Road tanker, tank Ships	Continuous, instantaneous	0.065
Category 2	Road tanker, tank ships	Continuous, instantaneous	0.01
Category 3, 4	Road tanker, tank ships	Continuous, instantaneous	0

Category	WMS Category	Limits
Category 0	Extremely flammable	Liquid substances and preparations with a flash point lower than 0°C and a boiling point (or the start of a boiling range) less than or equal to 35°C. Gaseous substances and preparations which may ignite at normal temperature and pressure when exposed to air.
Category 1	Highly flammable	Liquid substances and preparations with a flash point below 21°C, which are not however, extremely flammable
Category 2	Flammable	Liquid substances and preparations with a flash point greater than or equal to 21°C and less than or equal to 55°C.
Category 3	Flammable	Liquid substances and preparations with a flash point greater than 55°C and less than or equal to 100°C.
Category 4	Flammable	Liquid substances and preparations with a flash point greater than 100°C.

9.14.5. Delayed Ignition

The probability of delayed ignition depends on the end of the calculation. In the calculation of the location-specific risk only ignition sources on the site of the establishment are considered. Ignition sources outside the establishment are ignored: it is assumed that if the cloud does not ignite on site and a flammable cloud forms outside the establishment, ignition always occurs at the biggest cloud size. In the calculation of societal risk, all ignition sources are considered, including population. If ignition sources are absent, it is possible in the societal risk calculation that the flammable cloud does not ignite (see the table below).

Substance Category	Probability of Delayed Ignition for the Biggest Cloud Size, PRm	Probability of Delayed Ignition, GR
Category 0	$1 - P_{direct\ ignition}$	Ignition sources
Category 1	$1 - P_{direct\ ignition}$	Ignition sources
Category 2	0	0
Category 3	0	0
Category 4	0	0

10. RISK CALCULATIONS

Consequence analysis has been the focus of the report up to now while the consideration of probability has not been discussed. Risk is defined as consequence times probability.

Probability is defined as the risk of an event happening and impacting on the individual and society at large.

10.6. Location Specific Individual Risk Levels

The likelihood that a person in some fixed relation to a hazard (e.g. at a location, level of vulnerability, protection and escape) might sustain a specific level of harm.

The frequency at which an individual may be expected to sustain a given level of harm from the realisation of specified hazards. For example, there may be an individual risk of one-in-a-million that a person would be killed by an explosion at a major hazard near their home for every year that a person lives at that address. *[HSE Societal Risk: Initial briefing to Societal Risk Technical Advisory Group: p60].*

10.7. Employee Risk

Scenarios considered regarding risk to employees are toxic vapour clouds from Ammonia and chlorine plant failures, vapour cloud explosions and BLEVEs from gas vessel failures, and pool fires from fuel installations. Employees and the public are indoors and outdoors during the day and major events associated with these installations would occur outside of the building near the installation areas. When exposed to hazards such as toxic clouds, people who are indoors (sheltered) will generally be less vulnerable than those outdoors (unsheltered). The risks should not be more than one-in-a- thousand (1.0e-3 per year).

10.8. Individual Risk

This Assessment modelled the risks of the existing crude oil installations and proposed LPG installations in the port, the pipeline corridor and the SFF bulk terminal facility. The total individual risks are as follows:

Jetty at the Port of Saldanha

- The total individual risk involving the current crude oil jetty loading/offloading operations in the port is acceptable, with the 1.0e-6 (one-in-a-million) contour extending 12m beyond the Saldanha side of the jetty edge. The 3.0e-7 (one-in-thirty-million) contour extends 21m beyond the Saldanha side of the jetty edge and does not reach any sensitive areas or MHIs.
- The total individual risk involving the proposed LPG jetty loading/offloading operations in the port is acceptable, with the 1.0e-6 (one-in-a-million) contour extending 34m beyond the Langebaan side of the jetty edge. The 3.0e-7 (one-in-thirty-million) contour extends 81m beyond the Langebaan side of the jetty edge and does not reach any sensitive areas or MHIs.
- The total individual risk involving the current crude oil and the proposed LPG jetty loading/offloading operations in the port is acceptable, with the 1.0e-6 (one-in-a-million) contour extending 35m beyond the jetty edge. The 3.0e-7 (one-in-thirty-million) contour extends 85m beyond the jetty edge and does not reach any sensitive areas or MHIs.

Pipeline to the SFF Import Terminal Facility

- The total individual risks involving the current crude oil pipelines are acceptable, with the 1.0e-6 (one-in-a-million) contour extending 25m beyond pipelines. The 3.0e-7 (one-in-thirty-million) contour extends 45m beyond the pipelines and does not reach any sensitive areas or MHIs.
- The total individual risks involving the proposed LPG pipeline are acceptable, with the 1.0e-6 (one-in-a-million) contour extending 25m beyond the. The 3.0e-7 (one-in-thirty-million) contour extends 60m beyond the pipeline and does not reach any sensitive areas or MHIs.
- The total individual risks involving the proposed crude oil and LPG pipelines are acceptable, with the 1.0e-6 (one-in-a-million) contour extending 46m beyond the. The 3.0e-7 (one-in-thirty-million) contour extends 98m beyond the pipelines and does not reach any sensitive areas or MHIs.

SFF Import Terminal Facility

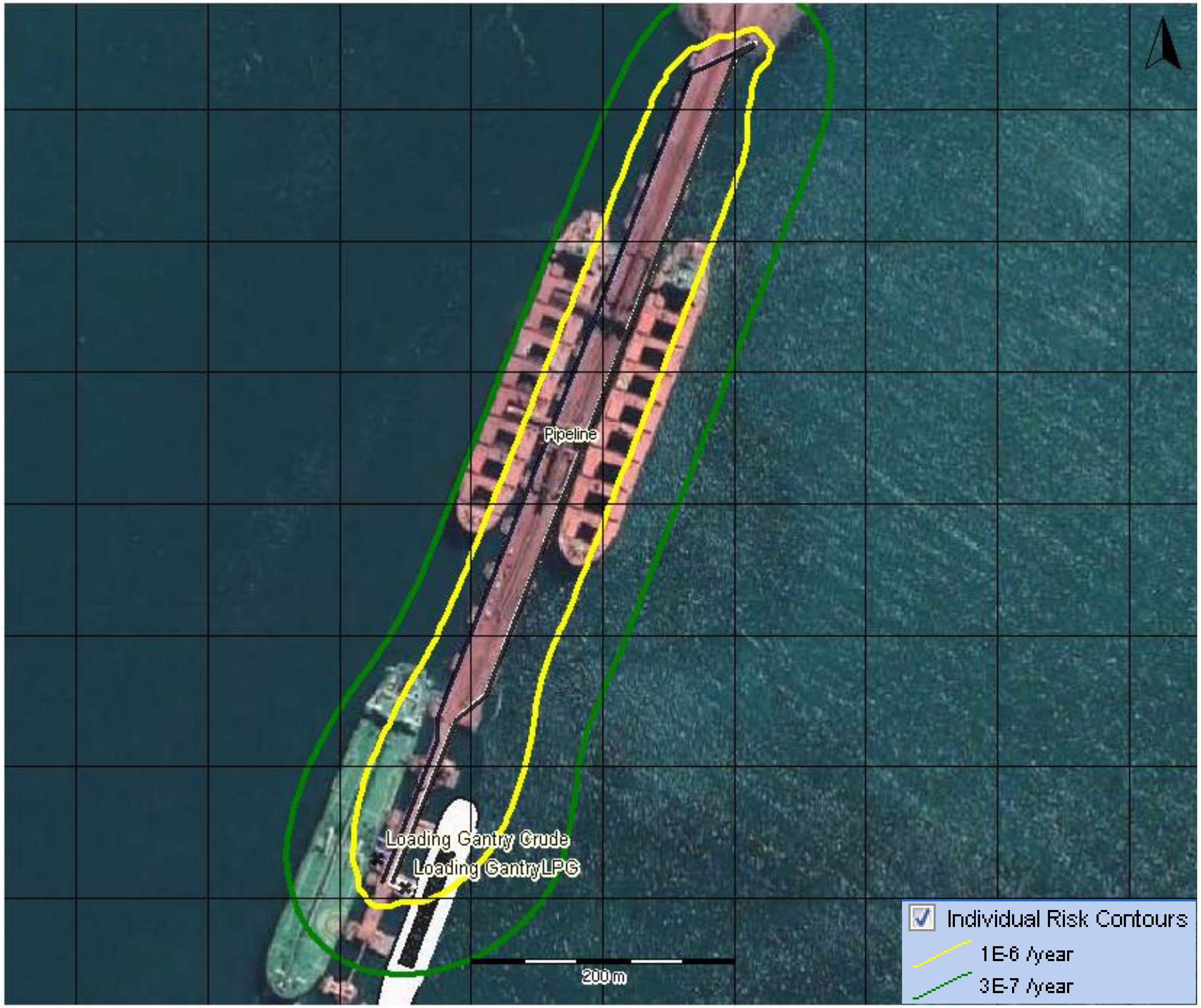
- The total individual risks involving the current crude oil installations are acceptable, with the 1.0e-5 (one-in-a-hundred thousand) reaching the edge of the bulk tanks but not reaching the property boundaries. The 1.0e-6 (one-in-a-million) contour extends beyond the property boundaries and just reaches the Moggs boundary to the east and across the main Saldanha/Langebaan road to the west. The 3.0e-7 (one-in-thirty-million) contour extends 570m beyond the boundaries, just reaching the Moggs tanks and does not reach any sensitive areas.
- The total individual risks involving the proposed LPG installations are acceptable, with the 1.0e-6 (one-in-a-million) contour extends 136m beyond the property

boundaries but does not reach the Moggs boundary to the east and does not reach the main Saldanha/Langebaan road to the west. The 3.0e-7 (one-in-thirty-million) contour extends 706m beyond the boundaries, just reaching the Moggs tanks and does not reach any sensitive areas.

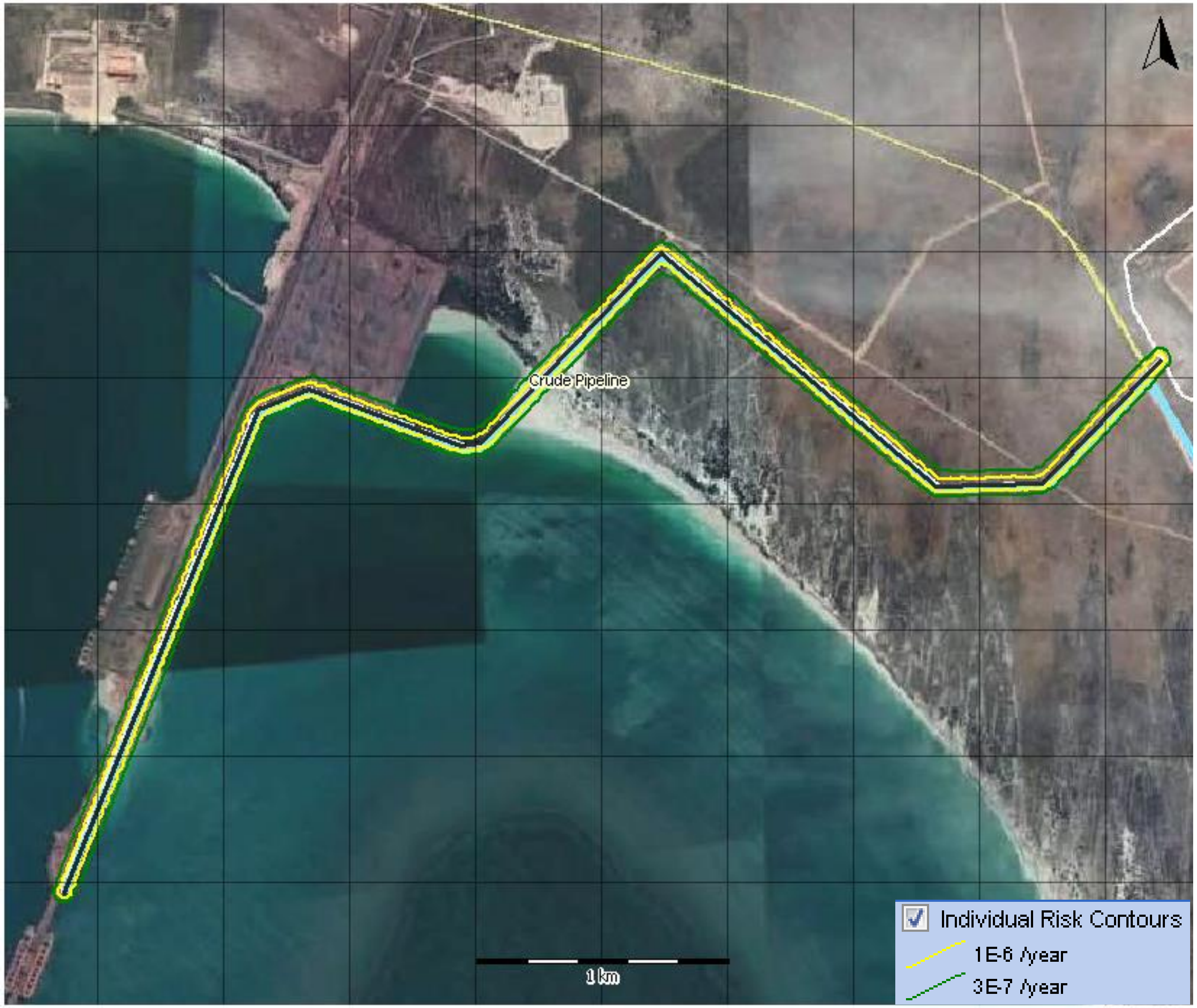
- The total individual risks involving the crude oil and LPG installations are acceptable, with the 1.0e-5 (one-in-a-hundred thousand) reaching the edge of the bulk tanks but not reaching the property boundaries. The 1.0e-6 (one-in-a-million) contour extends beyond the property boundaries and just reaches the Moggs boundary to the east and across the main Saldanha/Langebaan road to the west. The 3.0e-7 (one-in-thirty-million) contour extends 706m beyond the boundaries, just reaching the Moggs tanks and does not reach any sensitive areas.

Individual Risk Crude Oil Jetty

Individual Risk LPG Jetty

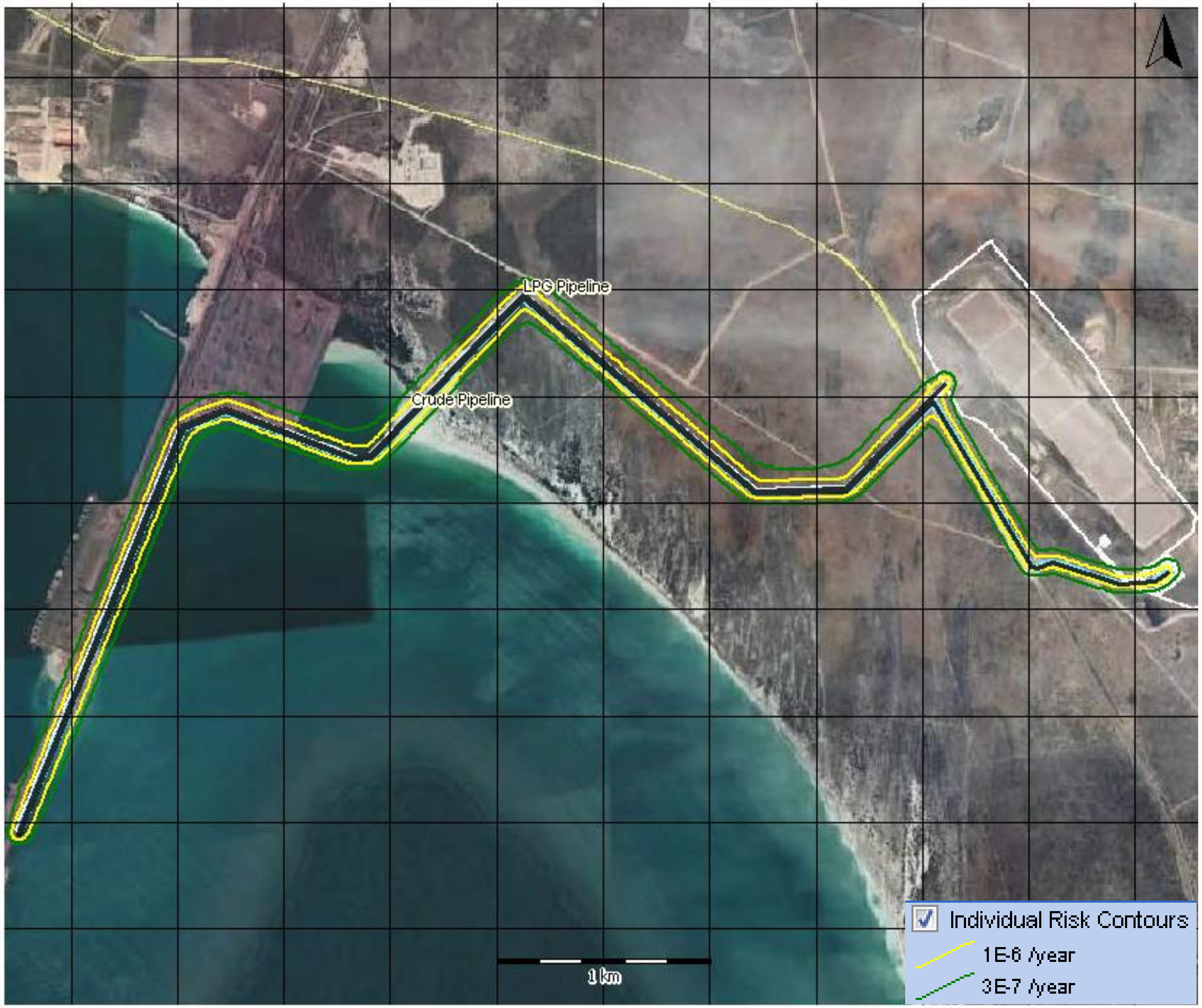


Individual Risk Crude Oil and LPG Jetty

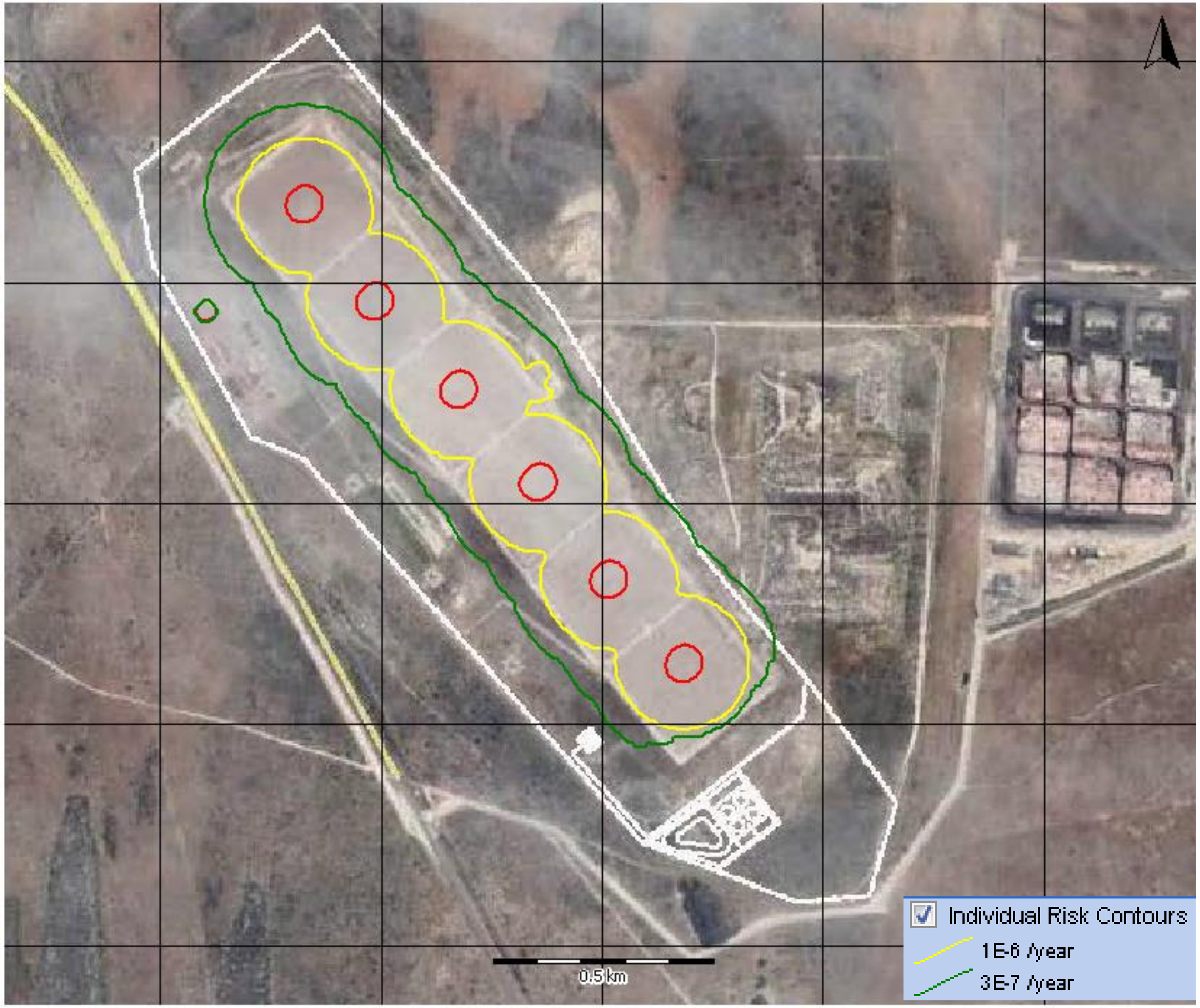


Individual Risk Crude Oil Pipeline

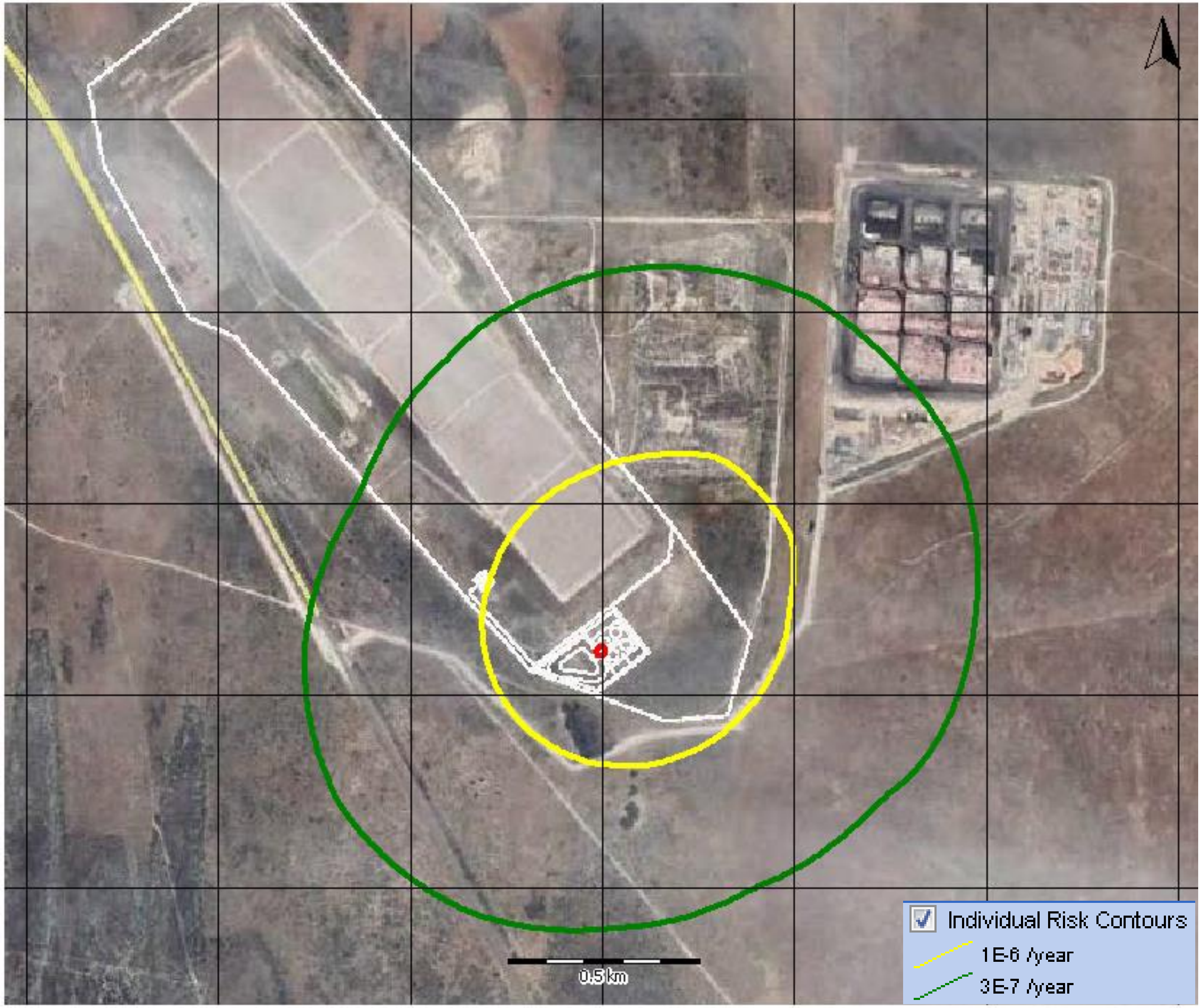
Individual Risk LPG Pipeline



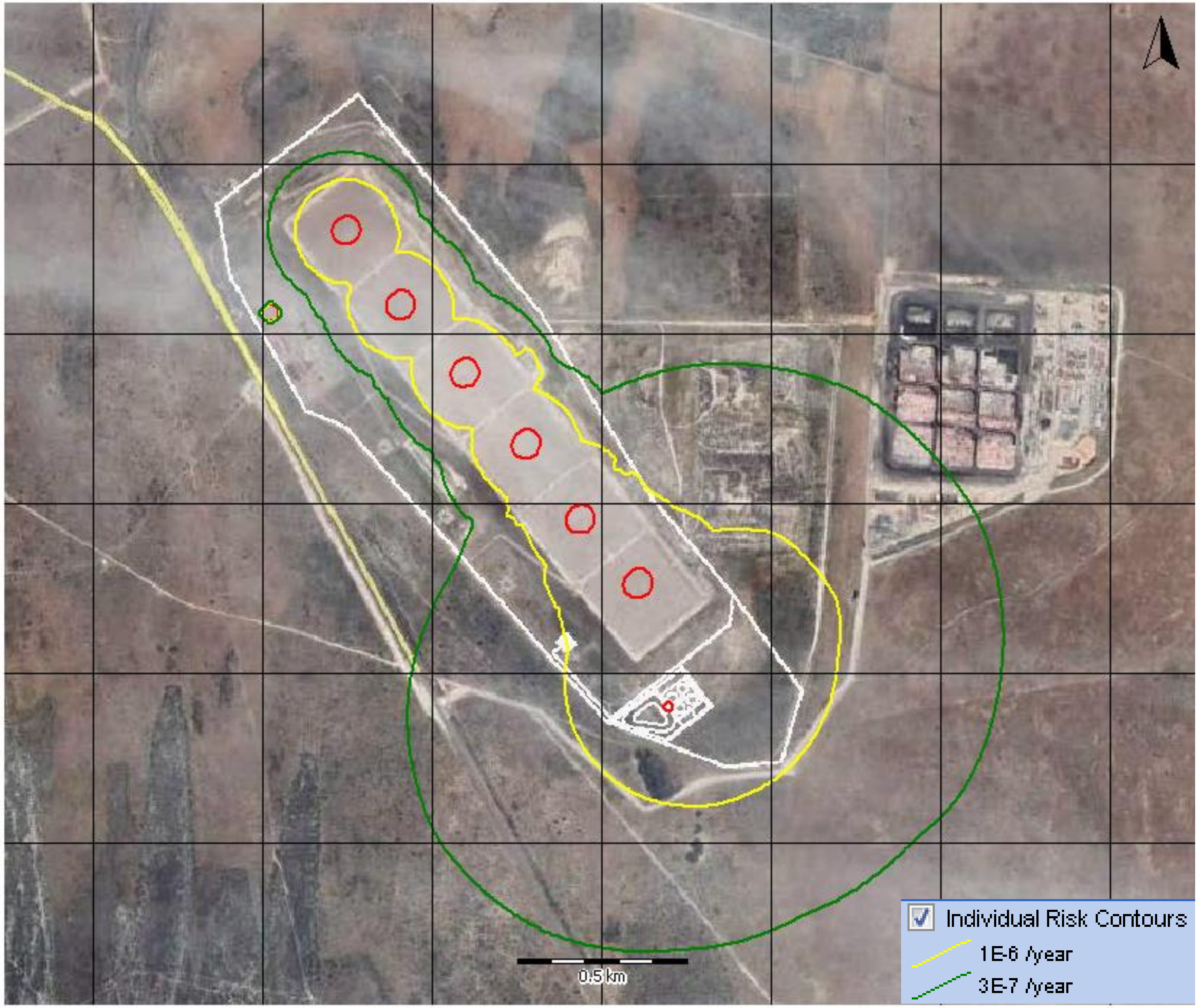
Individual Risk Crude Oil and LPG Pipeline



Individual Risk Crude Oil Installation



Individual Risk LPG Installation



Individual Risk Crude Oil and LPG Installation

10.9. Risk Levels and Ranking

Individual risk levels at several important points around the Jetty:

At the Ore Terminal

Scenario	Contribution %	Risk Value
LPG Pipe Rupture	64.2	1.13E-06
Crude Oil Pipe Leak	28.8	5.08E-07
Crude Oil Pipe Rupture	7	1.23E-07

Individual risk levels at several important points around the Pipelines:

25m Away from the Pipelines

Scenario	Contribution %	Risk Value
LPG Pipe Rupture	44.7	8.48E-07
Crude Pipe Rupture	38.3	7.46E-07
Crude Pipe Leak	16	3.03E-07

50m Away from the Pipelines

Scenario	Contribution %	Risk Value
LPG Pipe Rupture	40.2	1.54E-07
Crude Pipe Rupture	39.7	1.52E-07
Crude Pipe Leak	20.1	7.70E-08

Individual risk levels at several important points around the Import Terminal:

At Moggs Property

Scenario	Contribution %	Risk Value
Emptying in 10 Minutes Sphere 2	18.5	1.15E-07
Emptying in 10 Minutes Sphere 1	18.2	1.13E-07
Emptying in 10 Minutes Sphere 3	15.1	9.37E-08
Emptying in 10 Minutes Sphere 4	15	9.32E-08
Vessel Failure Sphere 4	8.36	5.18E-08
Vessel Failure Sphere 3	8.34	5.16E-08
Vessel Failure Sphere 2	8.21	5.08E-08
Vessel Failure Sphere 1	8.19	5.07E-08

At SFF Offices

Scenario	Contribution %	Risk Value
Vessel Failure Sphere 3	15.9	8.56E-09
Vessel Failure Sphere 1	15.5	8.36E-09
Vessel Failure Sphere 4	13.8	7.44E-09
Vessel Failure Sphere 2	13	7.03E-09
Emptying in 10 Minutes Sphere 3	10.8	5.82E-09
Emptying in 10 Minutes Sphere 1	10.6	5.74E-09
Emptying in 10 Minutes Sphere 4	10.3	5.57E-09
Emptying in 10 Minutes Sphere 2	10.1	5.47E-09

At Main Road

Scenario	Contribution %	Risk Value
Emptying in 10 Minutes Sphere 3	17.5	8.02E-08
Emptying in 10 Minutes Sphere 4	17.1	7.84E-08
Emptying in 10 Minutes Sphere 1	16.3	7.49E-08
Emptying in 10 Minutes Sphere 2	15.8	7.22E-08
Vessel Failure Sphere 4	8.58	3.93E-08
Vessel Failure Sphere 3	8.55	3.92E-08
Vessel Failure Sphere 1	8.36	3.83E-08
Vessel Failure Sphere 2	7.82	3.58E-08

Risk Ranking

Scenario	Contribution %	Risk Value
Pump Failure	23.8	2.42E-03
Vessel Failure (Sphere 1)	6.8	6.92E-04
Vessel Failure (Sphere 3)	6.75	6.87E-04
Vessel Failure (Sphere 4)	6.75	6.87E-04
Vessel Failure (Sphere 2)	6.75	6.87E-04
Emptying in 10 Minutes (Sphere 1)	5.29	5.39E-04
Emptying in 10 Minutes (Sphere 3)	5.27	5.37E-04
Emptying in 10 Minutes (Sphere 2)	5.27	5.37E-04
Emptying in 10 Minutes (Sphere 4)	5.27	5.36E-04
Loss of Containment (Crude Tank 5)	3.56	3.63E-04
Loss of Containment (Crude Tank 1)	3.56	3.63E-04
Loss of Containment (Crude Tank 3)	3.56	3.63E-04
Loss of Containment (Crude Tank 4)	3.56	3.63E-04
Loss of Containment (Crude Tank 6)	3.56	3.63E-04
Loss of Containment (Crude Tank 2)	3.56	3.63E-04
Tank Explosion (Crude Tank 2)	0.857	8.73E-05
Tank Explosion (Crude Tank 5)	0.857	8.73E-05
Tank Explosion (Crude Tank 6)	0.857	8.73E-05
Tank Explosion (Crude Tank 4)	0.857	8.73E-05
Tank Explosion (Crude Tank 3)	0.857	8.73E-05
Tank Explosion (Crude Tank 1)	0.857	8.73E-05
Pool fire Set (Crude Tank 1)	0.228	2.32E-05
Pool fire Set (Crude Tank 4)	0.228	2.32E-05
Pool fire Set (Crude Tank 3)	0.228	2.32E-05
Pool fire Set (Crude Tank 2)	0.228	2.32E-05
Pool fire Set (Crude Tank 5)	0.228	2.32E-05
Pool fire Set (Crude Tank 6)	0.228	2.32E-05
Vessel Failure (LPG Flare Vessel)	0.103	1.05E-05
Vessel Leak (G3) (LPG Flare Vessel)	0.0249	2.53E-06
Vessel Leak (G3) (Sphere 1)	0.0209	2.13E-06
Vessel Leak (G3) (Sphere 4)	0.0209	2.13E-06
Vessel Leak (G3) (Sphere 3)	0.0209	2.13E-06
Vessel Leak (G3) (Sphere 2)	0.0209	2.13E-06
Release in 10 min (G2) (LPG Flare Vessel)	0.0079	8.05E-07
Load Arm Leak (LPG Tanker2)	0.00118	1.20E-07
Load Arm Leak (LPG Tanker1)	0.00118	1.20E-07
Load Arm Rupture (LPG Flare Tanker)	0.00088	8.96E-08
Load Arm Rupture (LPG Tanker2)	0.000118	1.20E-08
Load Arm Rupture (LPG Tanker1)	0.000118	1.20E-08
Load Hose Leak (LPG Tanker)	0.000102	1.04E-08
BLEVE (LPG Tanker2)	4.94E-5	5.03E-09
BLEVE (LPG Tanker1)	4.94E-5	5.03E-09
BLEVE (LPG Tanker)	4.42E-6	4.51E-10

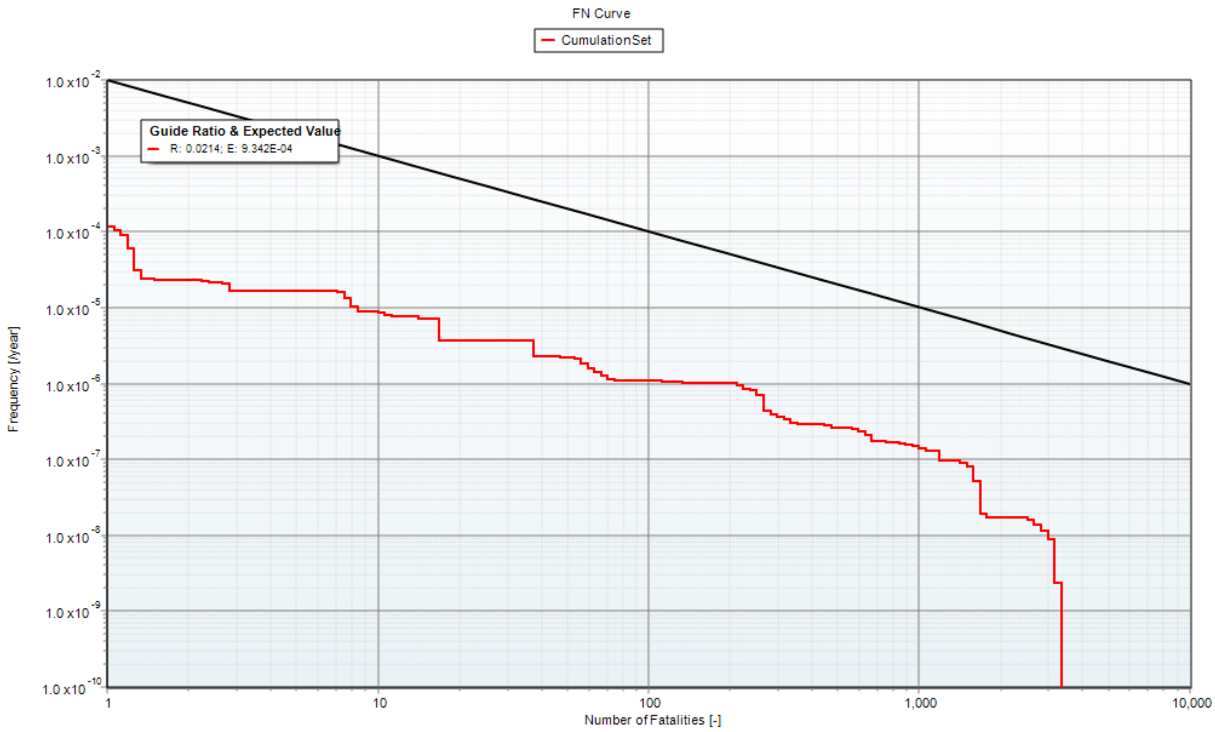
10.10. Societal Risk

Societal risk is defined as the relationship between frequency and the number of people suffering from a specified level of harm in each population from the realisation of specified hazards *[Jones, 1985].* Societal risk evaluation is concerned with estimation of the chances of more than one individual being harmed simultaneously by an incident. The likelihood of the primary event (an accident at a major hazard installation) is still a factor, but the consequences are assessed in terms of level of harm and the numbers affected (severity), to provide an idea of the scale of an accident in terms of numbers killed or harmed.

Societal risk is dependent on the risks from the substances and processes located on a major hazard installation. A key factor in estimating societal risk is the population around the site, its location and density. For example, the more (occupied) buildings in any area, the more people could be harmed by a flammable gas cloud passing through that area. For an installation with a population located in a specific compass direction, the chance of a flammable gas release would depend on the probability of drift in that direction.

Generally, scenarios to be included in a risk assessment can be characterised as having a frequency (F) and a consequence (N, number of casualties). *F* is used to denote the sum of the frequencies of all the individual events that could lead to *N* or more fatalities (hence the reference to *FN curves*).

Societal risk can be represented by FN curves, which are plots of the cumulative frequency (F) of various accident scenarios against the number (N) of casualties associated with the modelled incidents. The plot is cumulative in the sense that, for each frequency, N is the number of casualties that could be equalled or exceeded. Often 'casualties' are defined in a risk assessment as fatal injuries, in which case N is the number of people that could be killed by the incident.



FN Curves for Crude and LPG Installations

11. RISK JUDGEMENT

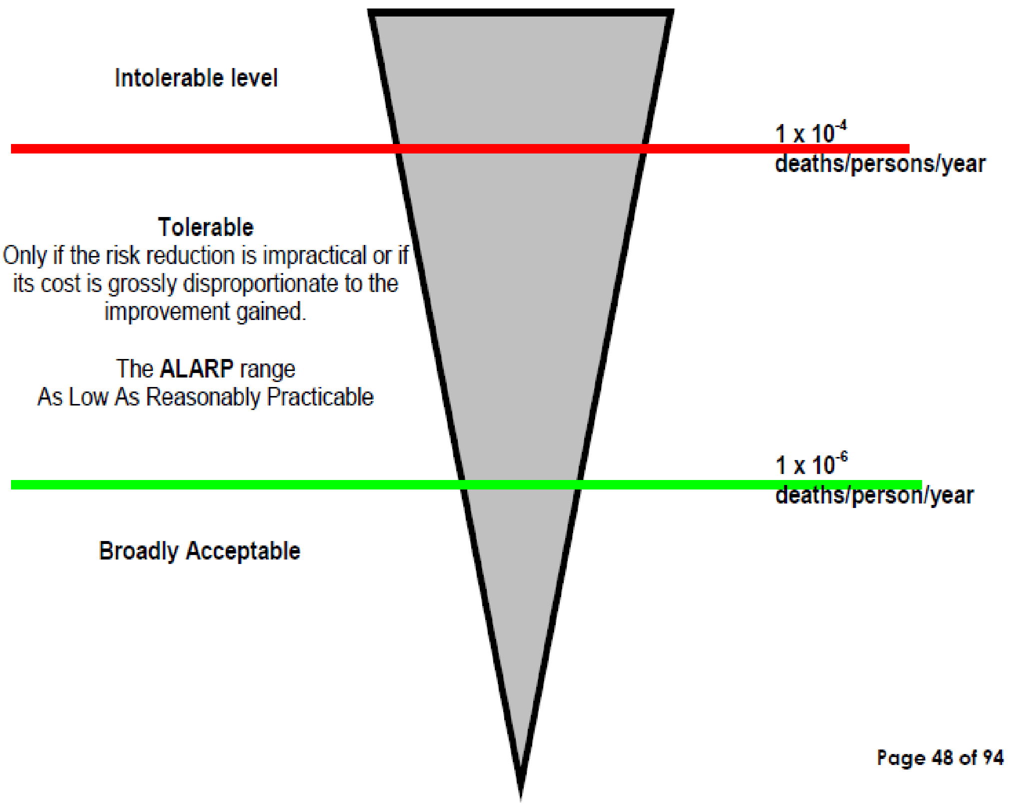
11.6. Risk Judgement Criteria

This Assessment indicates in a clear statement whether the risks or aspects of the risks are intolerably high, tolerable provided ALARP or broadly acceptable, both in terms of location specific individual risk and societal risk.

The risk evaluation criteria are set out as follows:

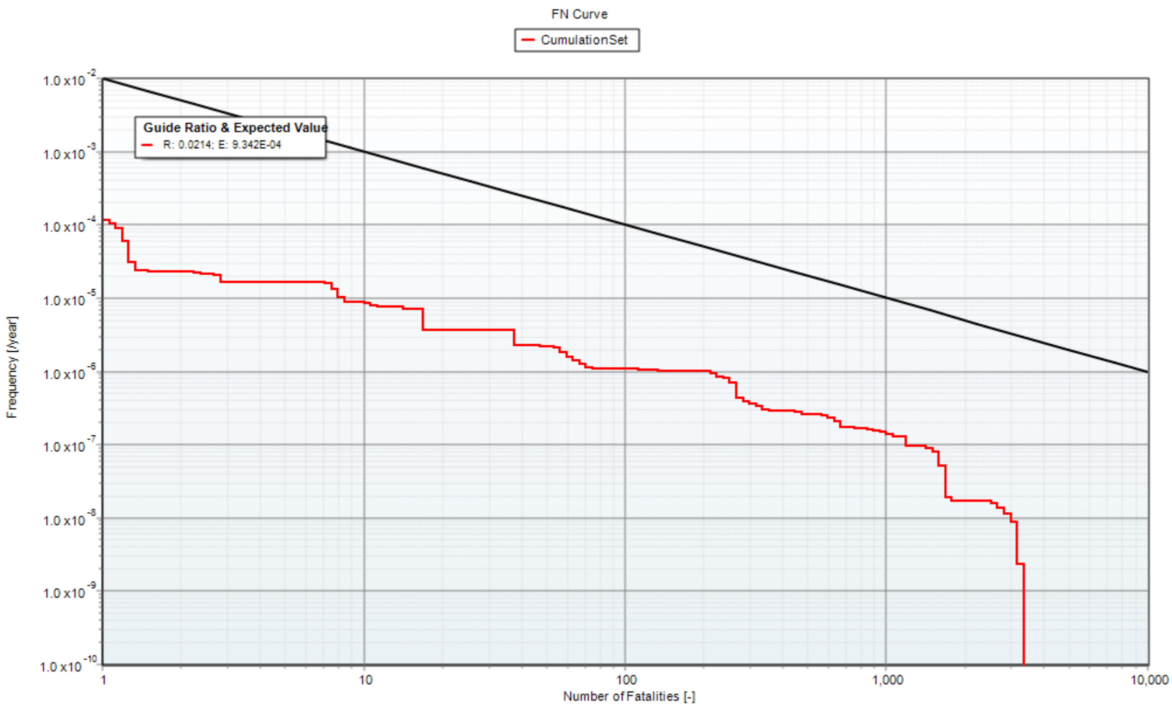
- A risk of death for members of the public greater than 1.0e-4 (one-in-ten-thousand) per year is considered intolerable.
- A risk of death below 1.0e-6 (one-in-a-million) per year for members of the public is considered broadly acceptable provided sensitive or vulnerable receptors in the vicinity have been considered.
- Risks between 1.0e-6 per year and 1.0e-4 per year for members of the public can be considered tolerable provided the risks have been be reduced so far as is reasonably practicable, i.e. this is referred to as the ALARP region.

Figure 1 - The public ALARP risk decision making framework



The individual risks on the SFF site are 'Tolerable' as it falls within the ALARP range. The risks off site are 'Broadly Acceptable'.

11.7. Societal Risks



FN Curves for Crude and LPG Installations

As can be seen on the graph above, the societal risk is less than 1x10e-4 of one fatality and is therefore acceptable.

12. RISK TREATMENT

9.1. Major Hazard Installation

At SFF the main risk contributing parts of the total risk at the facility are the LPG vessels, the crude oil installations.

The risks were found to be acceptable for the area in which the installations located.

9.2. Risk Reduction

The recommendations are as follows:

- Good housekeeping always needs to be observed on site;
- The Emergency Plan must be updated to include the scenarios described in this report;
- The Emergency Plan must comply with SANS 1514;
- The updated MHI report must be distributed to Local, Provincial and National Government as per MHI Regulations;
- Only suitably qualified people must be used for all work on the installations;
- Check alarms and emergency procedures regularly;
- Design and install proposed upgrades by suitably qualified people;
- Do a full pressure test to ensure that there are no leaks prior to commissioning the installations;
- Hazard Area Classification must be done as per SANS 10108;

9.3. ALARP Conclusions

If the installation is built as per plan with all the designed safety equipment in place, together with good maintenance and trained personnel, the risks imposed by the installation will always be acceptable.

10. LAND USE PLANNING

Where a site near to a major hazard chemical installation or pipeline is being developed, the City Council's Planning Authority has a statutory duty to refer to this Risk Assessment. This report will help the Planning Authority to 'Advise Against' or 'Don't Advise Against' the granting of planning permission on health and safety grounds that arise from the possible consequences of a major accident at the hazardous installation.

This report is designed to help planners, developers and others who want to work out for themselves about a planning proposal. In some cases, it may be that working through the report will allow one to modify the size, layout or location of a proposed development.

This report was compiled as per SANS 1461:2018 and land use planning is based on the United Kingdom's Health and Safety Executive *HSEs 'Planning advice for developments near hazardous installations (PADHI)'*.

10.1. The Principles Behind Land Use Planning Methodology

- The risk considered is the residual risk which remains after all reasonably practicable preventative measures have been taken to ensure compliance with the requirements of the Major Hazard Regulations.
- Advice takes account of risk as well as hazard, that is the likelihood of an accident as well as its consequences.
- Account is taken of the size and nature of the proposed development, the inherent vulnerability of the exposed population and the ease of evacuation or other emergency procedures for the type of development proposed. Some categories of development (e.g. schools and hospitals) are regarded as more sensitive than others (e.g. light industrial) and advice is weighted accordingly.
- Consideration of the risk of serious injury, including that of fatality, attaching weight to the risk where a proposed development might result in many casualties in the event of an accident.

10.2. Introduction to PADHI

The Risk Assessor sets a consultation distance (CD) around major hazard sites and pipelines after assessing the risks and likely effects of major accidents at the installation or pipeline.

Major hazards comprise a wide range of chemical process sites, fuel and chemical storage sites, and pipelines. The CDs are based on scientific knowledge using quantitative risk assessments.

PADHI uses two inputs to a decision matrix to generate the CDs or 'Restricted Development Distances':

- The zone in which the development is located of the three zones (that make up the CD);

The 'sensitivity level' of the proposed development (see 'Development Type Tables)

10.3. Zone Mapping

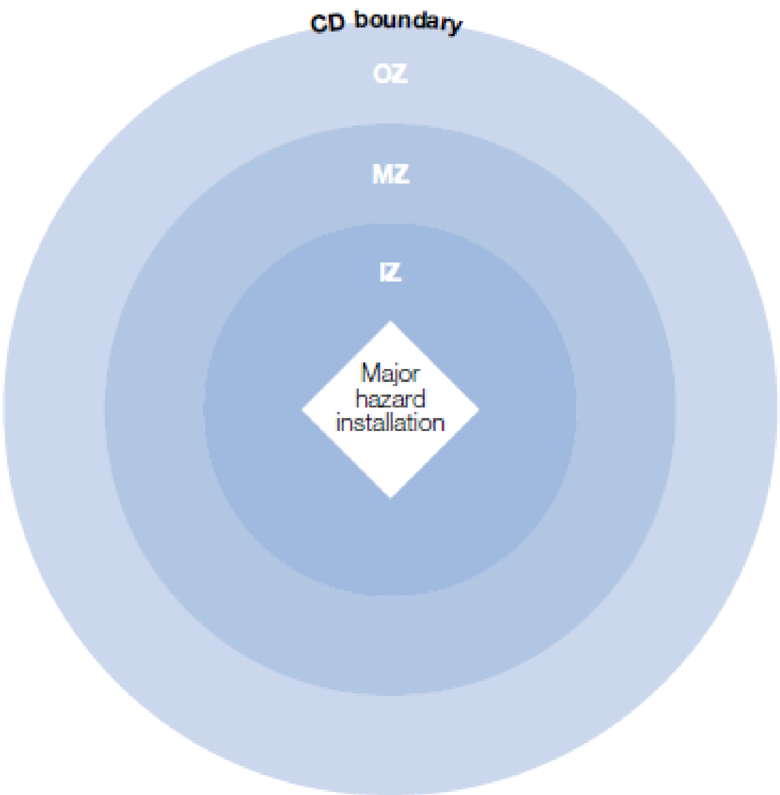
PADHI uses a 'three-zone' system. ('inner' (IZ), 'middle' (MZ) and 'outer' (OZ);) The zones are determined by a detailed assessment of the risks of the installation or pipeline which considers the following factors:

- The hazard ranges and consequences of the toxic and/or flammable substances present;
- The volume of those substances for which the site has consent;
- The method of storage. The risks and hazards from the major hazard are greatest in the inner zone, so the restrictions on development are strictest. The CD is all the land enclosed by all the zones and the installation itself.

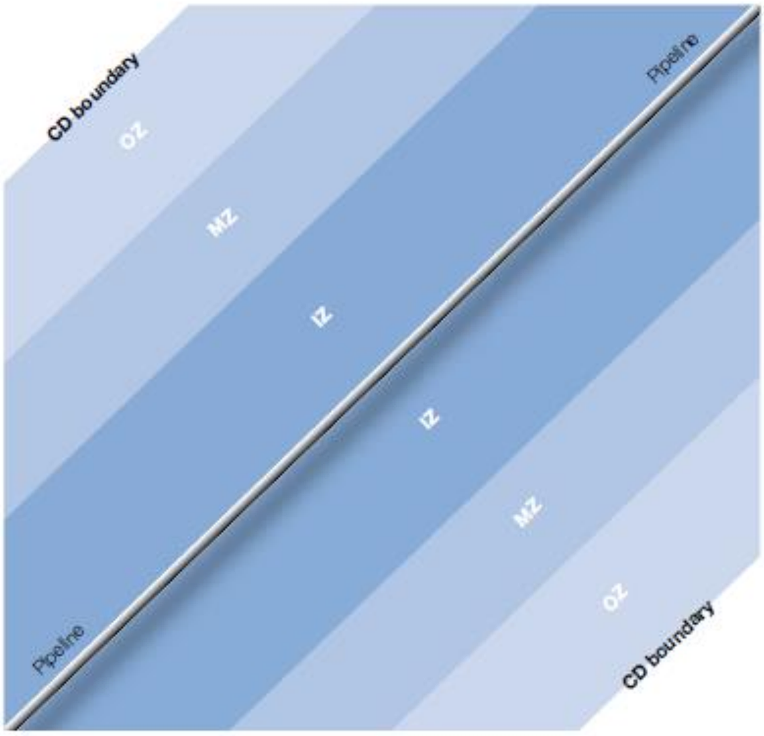
Inner zone includes all areas where risk is > 10 chances per million per annum.

Middle zone > 1 chance per million per annum.

Outer zone > 0,3 chances per million per annum.



Three Zone Map



Three Zone Pipeline Map

10.4. Development 'Sensitivity Levels'

The sensitivity levels are based on a clear rationale to allow progressively more severe restrictions to be imposed as the sensitivity of the proposed development increases. There are four sensitivity levels:

- Level 1 – Based on normal working population;
- Level 2 – Based on the general public – at home and involved in normal activities;
- Level 3 – Based on vulnerable members of the public (children, those with mobility difficulties or those unable to recognise physical danger);
- Level 4 – Large examples of Level 3 and very large outdoor examples of Level 2.

The tables in Appendix 15.5 expand on the four basic development types

10.5. Decision Matrix

Having determined which risk zone, the surrounding developments fall into and the sensitivity level of these developments, the matrix below can be utilised to decide whether one should advise for or against each specific development. Beyond the outer risk zone there are no specified restrictions on developments.

Level of Sensitivity	Development in inner zone	Development in middle zone	Development in outer zone
Level 1	Do not Advise Against (DAA)	Do not Advise Against (DAA)	Do not Advise Against (DAA)
Level 2	Advise Against (AA)	Do not Advise Against (DAA)	Do not Advise Against (DAA)
Level 3	Advise Against (AA)	Advise Against (AA)	Advise Against (AA)
Level 4	Advise Against (AA)	Advise Against (AA)	Advise Against (AA)

Decision Matrix

10.6. Site Specific Zoning

The area surrounding the site is already developed. All the existing developments are Level 1 developments. The existing zoning around the SFF site is correct.

10.7. Land Use Conflicts

There are no land use conflicts at the SFF site.

11. EMERGENCY RESPONSE DATA

11.1. Emergency Plan

Document Name	Emergency Plan SFF Saldanha Terminal
Date of Document	20 November 2018
Fire Fighting Addressed	Yes
Emergency Evacuation Addressed	Yes
Statutory Requirements	No not according to SANS 1514

12. CONCLUSION

12.1. Major Hazard Installation

This Assessment established that an incident involving the proposed upgrades and existing toxic and flammable installations on the premises of SFF could impact past the boundaries. The facility would not impact on any other MHIs. The risk associated with the MHI was found to be acceptable.

A site is only deemed to be an MHI if more than the prescribed quantity is stored as per the General Machinery Act or if a product is stored, handled or produced which has the potential to cause a major incident as per the Operational Health and Safety Act.

12.2. 1% Consequence Lethality Distances

Scenario LPG Installations	Maximum distance
Pipeline Leak Jet Fire	65m
Pipeline Leak VCE	62m
Flare Vessel Leak-Jet Fire	31m
Flare Vessel Catastrophic Failure-VCE	32m
Flare Vessel Catastrophic Failure-BLEVE	105m
Flare Vessel Catastrophic Leak-VCE	52m
Flare Vessel Catastrophic Leak-Jet Fire	29m
Sphere Vessel Leak-Jet Fire	32m
Sphere Vessel Catastrophic Failure-VCE	816m
Sphere Vessel Catastrophic Failure-BLEVE	1919m
Sphere Vessel Catastrophic Leak-VCE	19m
Sphere Vessel Catastrophic Leak-Jet Fire	484m
Loading Arm Leak-Jet Fire	11m
Loading Arm Rupture-Jet Fire	96m
Loading Arm Rupture-VCE	158m
Road Tanker Catastrophic Failure-BLEVE	202m

Scenario Crude Oil Installations	Maximum distance
30mm Loading Arm Leak - Pool Fire	40m
Loading Arm Rupture – Pool Fire	65m
Loading Arm Rupture - VCE	37m
Pump Failure – Pool Fire	43m
Fixed Roof Tank - Explosion	149m
Tank Top Fire - VCE	166m
Tank Top Fire – Pool Fire	55m
Pipeline Leak - Pool Fire	78m
Pipeline Leak - VCE	47m

12.3. Risk Level Posed to Various Populations

At Moggs Property

Scenario	Contribution %	Risk Value
Emptying in 10 Minutes Sphere 2	18.5	1.15E-07
Emptying in 10 Minutes Sphere 1	18.2	1.13E-07
Emptying in 10 Minutes Sphere 3	15.1	9.37E-08
Emptying in 10 Minutes Sphere 4	15	9.32E-08
Vessel Failure Sphere 4	8.36	5.18E-08
Vessel Failure Sphere 3	8.34	5.16E-08
Vessel Failure Sphere 2	8.21	5.08E-08
Vessel Failure Sphere 1	8.19	5.07E-08

At SFF Offices

Scenario	Contribution %	Risk Value
Vessel Failure Sphere 3	15.9	8.56E-09
Vessel Failure Sphere 1	15.5	8.36E-09
Vessel Failure Sphere 4	13.8	7.44E-09
Vessel Failure Sphere 2	13	7.03E-09
Emptying in 10 Minutes Sphere 3	10.8	5.82E-09
Emptying in 10 Minutes Sphere 1	10.6	5.74E-09
Emptying in 10 Minutes Sphere 4	10.3	5.57E-09
Emptying in 10 Minutes Sphere 2	10.1	5.47E-09

At Main Road

Scenario	Contribution %	Risk Value
Emptying in 10 Minutes Sphere 3	17.5	8.02E-08
Emptying in 10 Minutes Sphere 4	17.1	7.84E-08
Emptying in 10 Minutes Sphere 1	16.3	7.49E-08
Emptying in 10 Minutes Sphere 2	15.8	7.22E-08
Vessel Failure Sphere 4	8.58	3.93E-08
Vessel Failure Sphere 3	8.55	3.92E-08
Vessel Failure Sphere 1	8.36	3.83E-08
Vessel Failure Sphere 2	7.82	3.58E-08

12.4. Risk Reduction Recommendations

The following is recommended to reduce the risks associated with the installation on the site:

- Good housekeeping always needs to be observed on site;
- The Emergency Plan must be updated to include the scenarios described in this report;
- The Emergency Plan must comply with the MHI Regulations;
- The updated MHI report must be distributed to Local, Provincial and National Government as per MHI Regulations;
- Only suitably qualified people must be used for all work on the installations;
- Drawings must be done for all the flammable installations and submitted to council for approval;
- The flammable installations must comply to SANS 10131Codes of Practice;
- The Fire Department must witness a pressure test prior to issuing flammable substance certificate.

12.5. Emergency Plan

It is recommended that the Emergency Plan be updated to include the proposed flammable Installations. The Emergency Plan must include the risks identified in this report, as well as comply to SANS 1514.

12.6. Review of Risk Assessment

This Risk Assessment is valid for the duration of 5 years from the above date unless:

- Changes have been made to the plant that can alter the risks on the facility;
- The emergency plan was invoked or there was a near miss;
- The changing neighbourhood could result in offsite risks;
- There is reason to suspect that the current Assessment is no longer valid.

12.7. Risk Reduction Programmes

Risk reduction programmes should continually be investigated to reduce the impact from accidental fires and explosions on surrounding communities.

12.8. Surrounding Land Development

The development of land surrounding the site should be done with caution as not to pose unnecessary risks onto the surrounding communities. This caution is aimed at ensuring the adjacent developments are suitable for the risk imposed.

13. PROOF OF COMPETENCY

labour

Department:
Labour
REPUBLIC OF SOUTH AFRICA

National Department of Labour
Republic of South Africa

APPROVED INSPECTION AUTHORITY

Registered in accordance with the provisions of the Occupational Health and Safety Act, Act 85 of 1993, as amended and the Major Hazard Installation Regulations.

This is to certify that:

MAJOR HAZARD RISK CONSULTANTS CC

has been registered by the Department of Labour as an Approved Inspection Authority: Type A, to conduct Major Hazard Installation Risk Assessment, in terms of Regulation 5(5)(a), of the Major Hazard Installation Regulations.

CONDITIONS OF REGISTRATION:

- The AIA must at all time comply with the requirements of the Occupational Health and Safety Act, Act 85 of 1993, as amended.
- This registration certificate is not transferable.
- This registration will lapse if there is a name change of the AIA or change in ownership.

CHIEF INSPECTOR

Valid from: **21 January 2017**
Expires: **20 January 2021**
Certificate Number: **CI MHI 0007**

CERTIFICATE OF ACCREDITATION

In terms of section 22(2)(b) of the Accreditation for Conformity Assessment, Calibration and Good Laboratory Practice Act, 2006 (Act 19 of 2006), read with sections 23(1), (2) and (3) of the said Act, I hereby certify that:-

MAJOR HAZARD RISK CONSULTANTS CC
Co. Reg. No.: 2007/079078/23
CAPE TOWN

Facility Accreditation Number: **MHI0017**

is a South African National Accreditation System accredited Inspection Body to undertake **TYPE A** Inspection provided that all SANAS conditions and requirements are complied with

This certificate is valid as per the scope as stated in the accompanying schedule of accreditation, Annexure "A", bearing the above accreditation number for

THE ASSESSMENT OF RISK ON MAJOR HAZARD INSTALLATIONS

The facility is accredited in accordance with the recognised International Standard

ISO/IEC 17020:2012

The accreditation demonstrates technical competency for a defined scope and the operation of a management system

While this certificate remains valid, the Accredited Facility named above is authorised to use the relevant SANAS accreditation symbol to issue facility reports and/or certificates

Mr R Josias
Chief Executive Officer

Effective Date: 21 January 2017
Certificate Expires: 20 January 2021

This certificate does not on its own confer authority to act as an Approved Inspection Authority as contemplated in the Major Hazard Installation Regulations. Approval to inspect within the regulatory domain is granted by the Department of Labour.

14. REFERENCES

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SANS 1461:2018 *‘Major hazard installation — Risk assessments‘*

SANS 31000 *‘Risk management – Principles and guidelines’*

SANS 31010 *‘Risk management – Risk assessment techniques’*

SANS 10087:3-2015 *‘Liquefied petroleum gas installations involving storage vessels of individual water capacity exceeding 500 L’*

HSE "*Planning advice for developments near hazardous installations (PADHI)*"