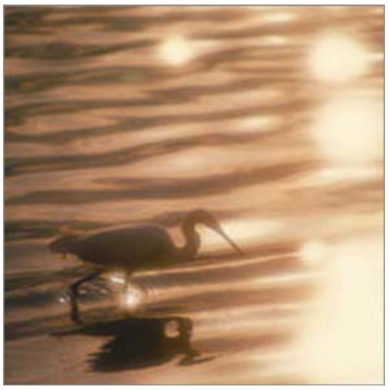
Prepared for

BlueScope Steel Limited

Illawarra Cogeneration Plant Project

Preliminary Hazard Analysis



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Executive Summary

A Preliminary Hazard Analysis (PHA) has been performed for the BlueScope Steel Illawarra Cogeneration Plant Project, to be located at Port Kembla, NSW, Australia. The PHA comprised:

- A hazard identification of incidents that may pose a significant risk to the public from accidental / atypical events;
- An outline of proposed operational and organisational safety management controls;
- A consequence and frequency analysis to determine the magnitude, impact and likelihood of major incidents; and
- An evaluation of the calculated hazard and risk levels with respect to the NSW Department of Planning (DoP) guidelines.

A range of incident scenarios has been considered in the analysis, including major gas leaks, explosions and fires from gas holder and pipe failures. The consequence analysis has shown that, except for major release from a catastrophic failure of the gas holder or gas supply lines during unfavourable wind/weather conditions, impacts are contained within site boundaries. However, the likelihood of such occurrences resulting in land use safety impact is very low, and has been estimated to be at less than one chance in ten million per year at the nearest residential area.

The risks imposed by the proposed works have been identified, estimated and compared with the criteria currently applied by the DoP for land use safety assessment. The land use safety planning risk levels for the proposed facilities satisfy the land use safety planning criteria.

The proposed safety systems (both hardware and software) are to be finalised for individual plant items. As the design is progressed further, the integrity, need and performance of such systems will be specifically considered through formal safety reviews such as Hazard and Operability (HAZOP) studies.

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1 INTRODUCTION

In May 2001, Duke Energy International (DEI) submitted a development application to Wollongong City Council (WCC), the consent authority, to build and operate a cogeneration plant at the Port Kembla Steelworks (PKSW). The project was called the "Illawarra Cogeneration Plant" (ICP). The development application (DA) was supported by an Environmental Impact Statement; *Illawarra Cogeneration Project Environmental Impact Statement Volume 1 and 2 – Appendices* (CH2M HILL, 2001), which included a Preliminary Hazard Analysis.

WCC granted consent for the construction and operation of the ICP in August 2002 (*Notice of Determination of Integrated Development Application No. D767/01*)(the Development Consent).

In October 2002, DEI ceased its involvement in the Project. BlueScope Steel is now sole developer of the Project. Since then, BlueScope Steel has had the opportunity to optimise and modify the design of the project.

The main components of the proposal now include the following:

- The cogeneration plant itself consisting of four boilers: three new 340-360 tonne steam/hr boilers and the use of existing upgraded No. 25 Boiler at 140 tonne steam/hr, a nominal 225 MW turbo alternator and auxiliary equipment required for the operation of the plant;
- A once-through saltwater cooling water system;
- Linz Donawitz Gas [Basic Oxygen Steelmaking Waste Gas] (LDG) collection system, including a 120,000m^3 nominal gross volume Wiggins Dry-Seal gas holder with dimensions approximately 65m high and 65m diameter and associated gantry and pipework; and
- Power cable infrastructure connections from the 18 Area HV Substation to the proposed cogeneration plant.

The proposed location of the cogeneration plant is south of the No. 2 Blower Station on land currently occupied by some minor production facilities and offices. The ICP will occupy an area of approximately 22,500 m^2, excluding the LDG collection and storage facility and the cooling water discharge infrastructure. The final location and area requirements for these facilities will be decided during detailed design. The proposed location for the LDG holder is to be north of Allan's Creek west of No.2 Electrical Construction Store.

A detailed environmental assessment has been prepared for the proposed modified facility (CH2M HILL, 2008), and provides more information for the basis and details of the proposed design. This revised Preliminary Hazard Analysis (PHA) accompanies this assessment as part of the document submission to Department of Planning (DoP) for approval under Part 75W of the *Environmental Planning and Assessment Act* 1979.

2 SCOPE AND METHODOLOGY

As part of the consent process for an industrial development, State Environmental Planning Policy No. 33 - Hazardous and Offensive Development (SEPP 33) may apply if the proposed development involves handling, storage or processing of a range of substances which in the absence of locational, technical or operational controls may create an off-site risk or offence to people, property or the environment. Such activities are defined as potentially hazardous or potentially offensive (DUAP, 1994), and may require the preparation of a PHA to determine the risk to people, property and the environment at the proposed location and in the presence of controls.

The PHA evaluates the acute land use safety planning risks that the proposed facility imposes on surrounding land uses. The NSW DoP Guidelines for the preparation of hazard analysis reports (Hazardous Industry Planning Advisory Planning Papers No.4 and No.6) were used as a basis for this report. The assessment considers abnormal (accidental) events, and not normal (or licensed) site emissions. It encompasses both qualitative and quantitative assessment. As outlined by the DUAP (1999), the level and extent of the analysis reflects the nature, scale and location of the proposed development.

The overall methodology involved hazard identification, consequence analysis, frequency analysis and risk assessment.

Hazard identification involves the qualitative definition of possible initiating events that could lead to hazardous incidents and assessment of their possible implications. It determines which events and incidents should be considered in more detail. An analysis of initiating events, and the consequences and effectiveness of the accident prevention/protection system was conducted.

Consequence analysis is the estimation and examination of consequences from identified events in terms of the physical effects such as heat flux, overpressure and toxic impact, on plant, people and the environment. The consequence analysis was carried out to determine the effective distances of specific incident scenarios defined in the hazard identification. This was achieved, where appropriate, by using various mathematical models and appropriate assumptions.

Frequency analysis involves the estimation and examination of the likelihood of each identified event and the probability of impacts from such events.

Risk assessment involves the estimation and examination of the frequency of target levels of impact to various land uses around a potentially hazardous installation. The risk levels are calculated for each event by combining the results of the consequence analysis and the frequency analysis to produce a cumulative risk result. The quantified risk result is used for comparative purposes against risk criteria recommended by the DoP (**Appendix A**).

3 SITE DESCRIPTION

3.1 Location and Surrounding Environment

Port Kembla is located approximately 80 km south of Sydney, within the City of Wollongong Local Government Area in the Illawarra region of New South Wales. The BlueScope Steel PKSW occupies about 742 hectares of the industrial area and is located mainly around the western and southern side of Port Kembla's Inner Harbour.

The site selected for the ICP is south of the existing No. 2 Blower Station. The area is currently occupied by a material testing laboratory, the torpedo ladle rebricking facilities and offices. The site selected for the gas holder is located North of Allan's Creek, west of No.2 Electrical Construction Store. The nearest public access area to the gas holders is the railway line and Springhill Road (approximately 250m to the west), with the nearest residences to the gas holders located about 1000m to the south-west (i.e. Cringila residential area), as shown in **Figure 1**.

3.2 Existing Operations

The PKSW is a fully integrated iron and steel making plant. It employs approximately 6,000 people and is the largest single employer in the Illawarra Region. The plant specialises in slabs for a plate mill and hot strip mill (at Port Kembla), feedstock for a hot strip mill at Western Port (Victoria), and for direct sale to international markets. In addition, it supplies hot rolled coil, plate and cold-rolled coil to domestic and international markets. Facilities on site include two blast furnaces, three BOS furnaces and three slabcaster machines. The steelworks site comprises manufacturing areas, storage areas for raw materials and final products, internal private roads, office buildings, internal railway, port berths and associated facilities.

3.3 Replaced Operations

The proposed development will result in the decommissioning of No.1 Power House located towards the nearest residential area at the south-west side of the steelworks site. Boilers in No.2 Blower Station will also be decommissioned but these are located away from the boundary of the steelworks.

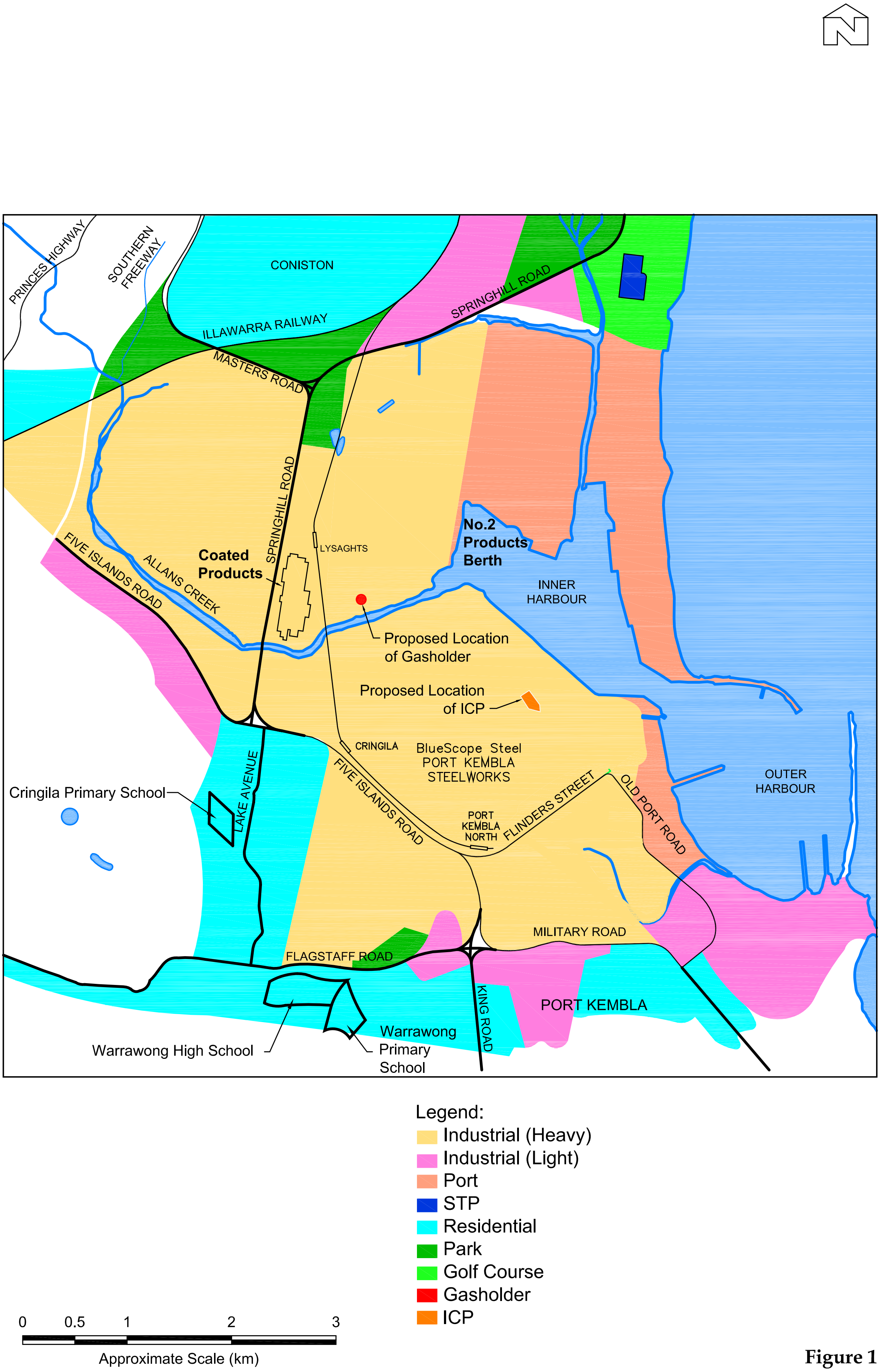


Figure 1
Land Use Map

4 PROPOSED FACILITY

The main components of the proposal are:

- The cogeneration plant itself consisting of four boilers: three new 360 tonne/hr boilers and the use of existing No. 25 Boiler at 140 tonne/hr, a nominal 225 MW turbo alternator and auxiliary equipment required for the operation of the plant;
- A once-through saltwater cooling water system;
- Basic oxygen steelmaking LDG collection system, including a 120,000m^3 gross volume Wiggins Dry-Seal gas holder with dimensions approximately 65m high and 65m diameter and associated gantry and pipework; and
- Power cable infrastructure connections from the existing 18 Area HV substation to the proposed cogeneration plant.

A comprehensive description of the cogeneration plant was contained in the EIS (CH2M HILL, 2001). The basic design philosophy of the cogeneration plant is not expected to significantly change.

The key change from the 2001 design has been a reduction from four new boilers to three new boilers and retaining and utilising existing boiler No. 25 as well as the relocation of the gas holder to a less congested (in terms of onsite staff and equipment) part of the site, as well as an increase in the gas holder size, due to a need to be less reliant on Coke Ovens and Blast Furnace Gases.

The following information outlines details of the proposal to assist with understanding land use safety planning considerations. The proposed location of the facilities on the BlueScope Steel site is shown in **Figure 2**.

The design as proposed is preliminary. In this regard, vessel sizes, pipe diameters, operating conditions, type, location and number of valves are not yet confirmed. Therefore, assumptions have been made based on the probable conservative expectations by BlueScope Steel.

4.1 Cogeneration Plant

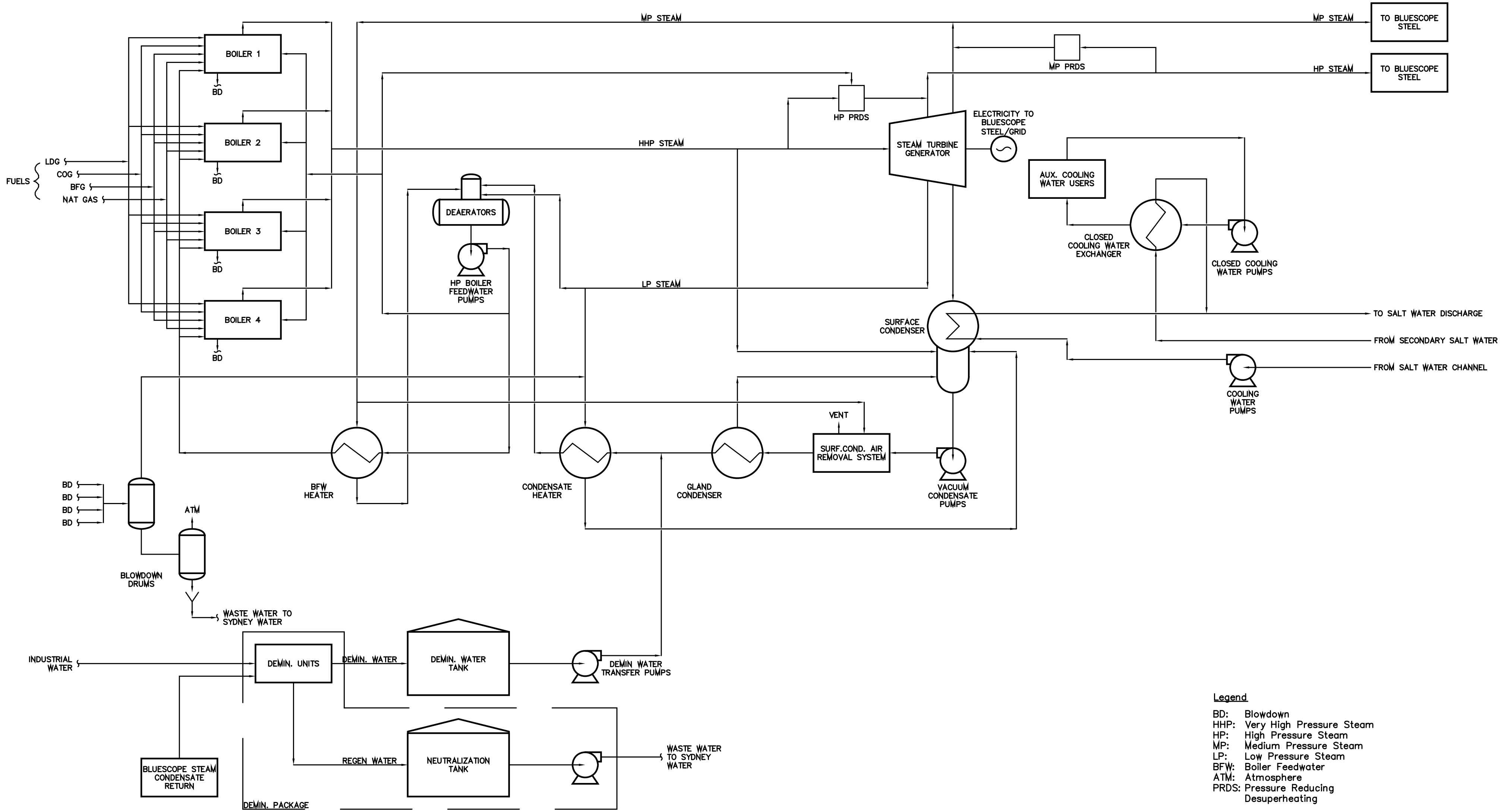
The cogeneration plant will utilise waste gases that are a by-product of BlueScope Steel iron and steelmaking processes. These by-product gases consist of blast furnace gas (BFG), coke ovens gas (COG) and Basic Oxygen Steelmaking (BOS) LDG. The LDG is named after Linz-Donawitz, the locations of steelworks where the basic oxygen process was first commercially used. It is referred to onsite as "BOS off-gas" before collection and LDG once the gas has been collected. The term LDG will be used in this report.

In addition, natural gas (imported fuel) will also be used as a supplementary source of fuel for pilots and any incremental peak period capacity. A preliminary Process Flow Diagram for the cogeneration plant is shown in **Figure 3**.

The works fuel balance may be significantly affected during plant outages such as Blast Furnace shutdowns, BOS Furnace stops and gas holder outages. Boiler firing will be adjusted to cope with variations in availability of by-product fuels while still maintaining critical steam supply. The cogeneration plant will have sufficient capacity and controls to accommodate significant variations in the volumes supplied, in order to consume most of the surplus by-product fuels.

Source: Gregory's Maxi 2003 Street Directory 15th Edition

Figure 2
Site Layout



Legend

- BD: Blowdown
- HHP: Very High Pressure Steam
- HP: High Pressure Steam
- MP: Medium Pressure Steam
- LP: Low Pressure Steam
- BFW: Boiler Feedwater
- ATM: Atmosphere
- PRDS: Pressure Reducing Desuperheating

Figure 3
ICP Process Diagram

The three new boilers will raise steam at 105 bar(g) and 520°C, existing No. 25 Boiler will raise steam at 42bar(g) and 454°C. The steam raised will be directed to the steam distribution network supplying the steelworks and drive a generator, which will produce electricity.

The cogeneration plant will be designed to have a maximum combined steam export rate from the three new boilers of 600 tonnes/hr which will be exported as 500 tonne/hr high pressure (HP) (42 bar(g)/454°C) and 100 tonne/hr medium pressure (MP) (18 bar(g)/320°C) steam. The average steam export for supply to the BlueScope Steel PKSW (from three new boilers and the use of existing No. 25 Boiler) will be approximately 352 tonnes/hr of HP steam and 10 tonnes/hr of MP steam.

Ancillary equipment to be used for the cogeneration plant includes:

- A once through saltwater cooling system and discharge device;
- A substation and electrical connection (132 kV and 33 kV powerlines) from the cogeneration plant to the substation;
- Steam export pipelines from the plant to the existing distribution system;
- Wastewater collection and discharge; and
- Water treatment (demineralisation plant).

4.2 LDG Recovery System

4.2.1 Supply of LDG

LDG will be supplied via a recovery system. A diagram representation of a possible arrangement for the proposed system (with gas holder) is shown in **Figure 4**.

LDG is produced during the steelmaking process. In the current process, all of this gas is disposed of by combustion in flares (a total of three, one per vessel). It is proposed that acceptable portions of the LDG be collected and used as fuel in the ICP boilers.

The BOS process involves blowing oxygen into a vessel containing liquid iron. The oxygen reacts with the carbon content of the iron to form carbon monoxide (CO), leaving a low carbon steel. Other impurities are removed by slag reactions. The CO is extracted from the furnace by an Off Gas (OG) system.

BOS OG systems at BlueScope Steel are a “suppressed combustion” type, which utilise a moveable skirt above the furnace to limit the amount of air drawn into the gas ductwork at the furnace mouth and ensure that no more than 10% of the CO generated is burned at the furnace mouth. This ensures a high “quality” gas is available for collection.

Steel is manufactured in 300 tonne “heats”. The batch nature of this process means that the existing OG system alternates between being an air filled system and a gas filled system. The OG process employed at BlueScope Steel ensures safe transitions at the start and finish of each heat. This is done by controlling the skirt position and furnace mouth suction at each transition, to completely combust the LDG at the furnace mouth for a short period of time. This generates a plug of inert gas that provides a separation buffer between the air filled and gas filled conditions. Consequently, the inert plug period is composed mainly of CO_2 and N_2 and contains only small concentrations of CO or Oxygen. At start of the blow, once the inert plug period has elapsed, the skirt is lowered and the quantity and quality of the gas quickly increases. At approximately one or two minutes after the start of blowing, the gas recovering conditions are satisfied, and gas recovery commences automatically.

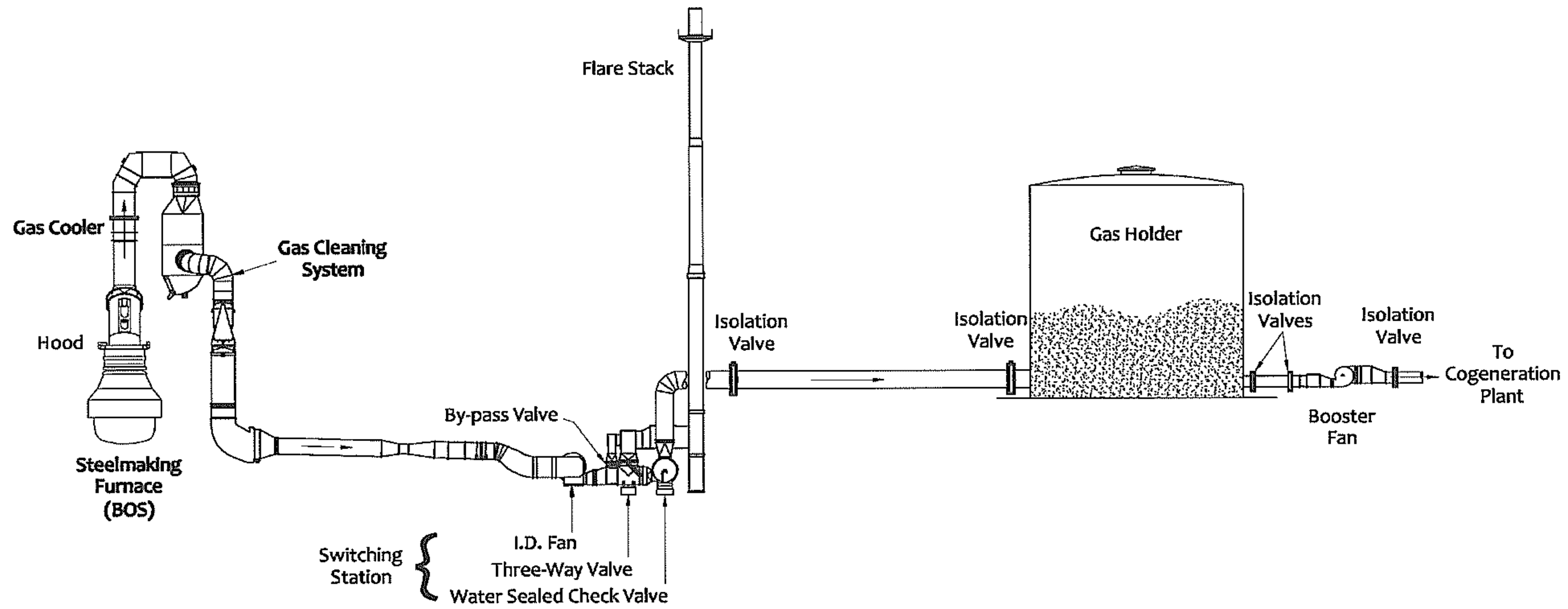


Figure 4
LDG Recovery System

When directed to collection mode, the LDG is transported via a 2400mm nominal diameter gas duct to a gas holder.

The blowing period for each heat is approximately 15 – 16 minutes. Towards the end of the blowing period as the reaction in the vessel nears completion, the CO content of the gas deteriorates until it is no longer acceptable for collection. Gas flow is automatically diverted back to the flare stack for the remainder of the blowing period.

Just prior to stop blow there is another period of inert plug generation and after stop blow air is drawn through the OG ductwork to the stack.

LDG will be collected during the part of the blowing cycle when the gas is high in CO (approximately 30 - 70% on a dry basis) and low in O_2, and has a high calorific value (CV). Collection periods during each heat are typically 10-12 minutes.

All changeover sequences are done automatically.

4.2.2 Switching Station

A new switching station arrangement will be installed for each BOS furnace to allow the changeover to occur. These switching stations are located adjacent to the existing flare stacks.

Each switching stations includes a:

- **Three Way Changeover Valve**

This consists of two mechanically linked butterfly valves, which allow a flow path to either the flare stack or the collection system. The valves are hydraulically actuated and have a fail safe mode which will direct gas to flare stack in the event of failure/ emergency.

- **Water-seal Check Valve**

This is a rotary drum device which provides isolation of the furnace OG system from the collection network via a water-seal during non- collection periods. The water seal check valve is hydraulically actuated and has a fail safe condition which shuts off gas flow to the collection system.

- **Bypass Valve**

The bypass valve operates in parallel to the flare stack leg of the three way valve. It ensures there will be a flow path for discharge of gas in the event of a failure of the three way valve. The bypass valve is hydraulically operated and has a fail safe mode which allows discharge of gas to the flare stack.

- **Nitrogen Purge System**

Three new automatic nitrogen purge systems are incorporated to ensure mixing of air and gas does not occur during the changeover sequences, specifically:

- B1 Purge; which provides a nitrogen purge to the bypass duct.
- B2 Purge; which provides nitrogen purges to the main discharge duct and stack.
- D Purge; which provides a nitrogen purge to the section of duct between the 3 way valve and the water seal check valve.

- **Goggle Valve**

A fully enclosed goggle valve is provided at the outlet of the switching station. This is used to provide positive isolation of an individual switching station from the collection system during maintenance. It may also be used in the event of failure of the water seal check valve. The goggle valve is hydraulically actuated but will have provision for manual operation in the event of power failure.

- **Hydraulic System**

The hydraulic system to actuate all the above valves is a fail safe design to ensure valve actuation actually occurs during power outages and other emergencies. This is achieved through stored accumulator capacity sufficient for at least two valve operations after a power failure.

4.2.3 LDG Collection

The gas generated from the commencement of blowing to the start of LDG recovery is discharged by an existing Induced Draft (ID) fan to the flare stack via the three-way valve. At the same time the collect leg is nitrogen purged (D purge) in preparation for recovery. When all conditions for gas recovery are established, the water-sealed check valve opens. The gas flow is then switched over from the flare stack side to the recovery side by the three-way valve, and sent to the gas holder. The flare stack leg is then nitrogen purged to remove flammable gas (B1 and B2 Purge).

During collection, the ID fan will boost the pressure for delivery of gas to the gas holder. Discharge pressure from the fan will be approx 6 kPa, while the holder pressure will be maintained at approx 2 kPa.

When the LDG quality drops below acceptable limits or any one of the other gas recovering conditions are not satisfied, or an emergency such as power failure occurs during gas recovery, the gas recovery operation is automatically stopped, and the gas flow is changed over from the recovery side to the flare stack side by the three-way valve. The water-seal check valve is subsequently closed to prevent reverse flow from the gas holder and a nitrogen purge of the collection leg is carried out (B1 purge).

Collected LDG will be delivered to the gas holder area by a 2.4m diameter collection main. As the gas travels along the main, some cooling will occur and condensate will drop out in the main. A number of condensate collection points will be provided along the gas main to allow continuous removal of condensate.

4.2.4 Gas Holder

The batch nature of the steel making process and the variability in the CV of the LDG means that there needs to be a Gas Holder to act as a buffer in the LDG Collection System to regulate the LDG supply to the power plant.

Computer modelling of the process indicates that a gross volume of 240,000 m^3 would be required as a buffer to eliminate interruption of LDG to the boilers. However, by optimising the use of LDG in the cogeneration plant, it is possible to reduce the volume of the holder to 120,000 m^3 whilst maintaining the capability to collect a high proportion of usable LDG.

The selection of the type of gas holder for use in the LDG Collection System is based on its material compatibility with LDG and the process requirements of the collection system. Factors that were considered in the selection of a “Wiggins” type dry seal gas holder for the collection system are:

- The compatibility of the seal with LDG;
- Suitability for use with wet gas;
- Suitability for use with dirty gas;
- Maximum piston operating speed;
- Pressure profile; and
- Seal temperature limit.

The capacity of the gas holder is determined based on the fluctuation in gas production at the BOS and the pattern of gas consumption at the power plant.

Whilst not finalised at this stage, the expected dimensions of the gas holder are approximately 65m high and 65m diameter.

The Wiggins type gas holder is the gas holder of choice for use in LDG Collection Systems. There are in excess of 40 Wiggins type gas holders operating in LDG Collection Systems throughout the world, and many more being utilised with other gases such as Blast Furnace Gas and Coke Ovens Gas.

The bottom two thirds of the holder is utilised for gas storage. The proposed holder will be capable of storing 120,000m^3. The seal attaches to the holder wall approximately one third up the tank wall and forms a gas tight gas space with the piston. The space above the piston is known as the air space and contains apertures for access and ventilation. As gas enters beneath the piston, the space fills, the seal unfolds and lifts the piston. The piston will continue to rise as long as more gas enters the holder than is being removed by the gas consumer. When the holder reaches its top limit, a signal is sent to the changeover valves to close. This prevents the holder from becoming overfull and activating the relief valves. Similarly, if more gas is being consumed than produced and the gas holder level nears empty, a signal is sent to the power plant to reduce its gas consumption to avoid emptying the holder. If the gas holder is emptied the seal acts like a shut off valve on the exit of the holder.

Access to the holder is provided via an external staircase, and shell access doors. Shell vents in the roof of the gas holder allow air to be displaced from the inside of the gas holder as the piston rises.

The seal is designed so that it pulls over the outlet main and acts as a shut off valve when the piston reaches low/empty position. This prevents any further discharge of gas, and as a result avoids creating a vacuum in the holder.

Volume relief pipes are provided to protect the gas holder from over pressurisation. The volume relief valves are actuated by the piston fender (at high level the fender pushes open the “lid” of the relief pipe) and a limit switch is actuated. Relief valve position is indicated in the control room and alarmed in the event of any vents not being closed.

A mechanical counter balance keeps piston moments in equilibrium. Level weights run up and down tracks on the gas holder shell and actuate limit switches when gas holder volume reaches pre-defined settings. The limit switches send signals to activate operation of gas import and export valves. Each weight exerts an action on two diametrically opposite points on the piston, through a pair of cables. Any attempt by the piston to tilt out of level, results in one of the pair of cables taking the entire load of the counterweight and the other taking nothing. Under tilt condition, the cable taking the entire load will reduce the out-of-level of the piston until the loads are once again evenly shared.

The ropes used in the level weight system are designed to give improved wear and fatigue resistance.

The most common form of problems with the holder structure itself is corrosion and mechanical damage of the fender and its support structure. BlueScope Steel is investigating the use of a single seal holder, which eliminates the coupling and uncoupling issues problems, which are common in the double seal, older design Wiggins type gas holders. The single seal design has less chance of radial movement and out of balance. The holder is also painted internally with a coating specified to be compatible with the gas, and the condensate the holder will contain. Compared to the older two stage Wiggins design, the shell material is also thicker, which means that external and internal corrosion is a less critical issue.

The most likely mode of failure is damage affecting the seal. This failure may be due to thermal degradation, or to the piston coming into contact with the shell resulting in the seal being trapped, as a result from incorrect operation, incorrect maintenance or tools being dropped into the gap between the shell and the seal.

No particular locations of the seal are more vulnerable than any others. As a general rule, joints may be more vulnerable, and butt-lapped seams provide greater protection than others. Operating at higher than recommended temperatures will cause embrittlement of the seal material. This can lead to cracking, but not catastrophic failure.

During the commissioning of the gas holder, the seal is aligned and set-up to the optimum operating position.

4.2.5 Booster Fans

LDG is drawn from the steelmaking process via an ID fan and is delivered to the gas holder. The gas drawn from the gas holder is boosted by the gas booster fan for delivery to the boilers. Two LDG booster fans will be provided (with one in operation and one on standby). A pressure control mechanism will ensure constant delivery pressure to the boilers. The booster fans will be located at the LDG gas holder site.

4.2.6 Gas Main Purging

It is anticipated that the LDG Collection ducting between the BOS switching stations and gas holder will need to be taken off line on a regular basis (approx 1 per year) for cleaning out accumulated dust. The LDG Supply gas main between the holder and the cogeneration plant will eventually need to be isolated for repairs and/or cleaning.

Provision must be made to safely isolate and purge these mains. They both represent a significant volume of high CO gas that must be vented during this operation.

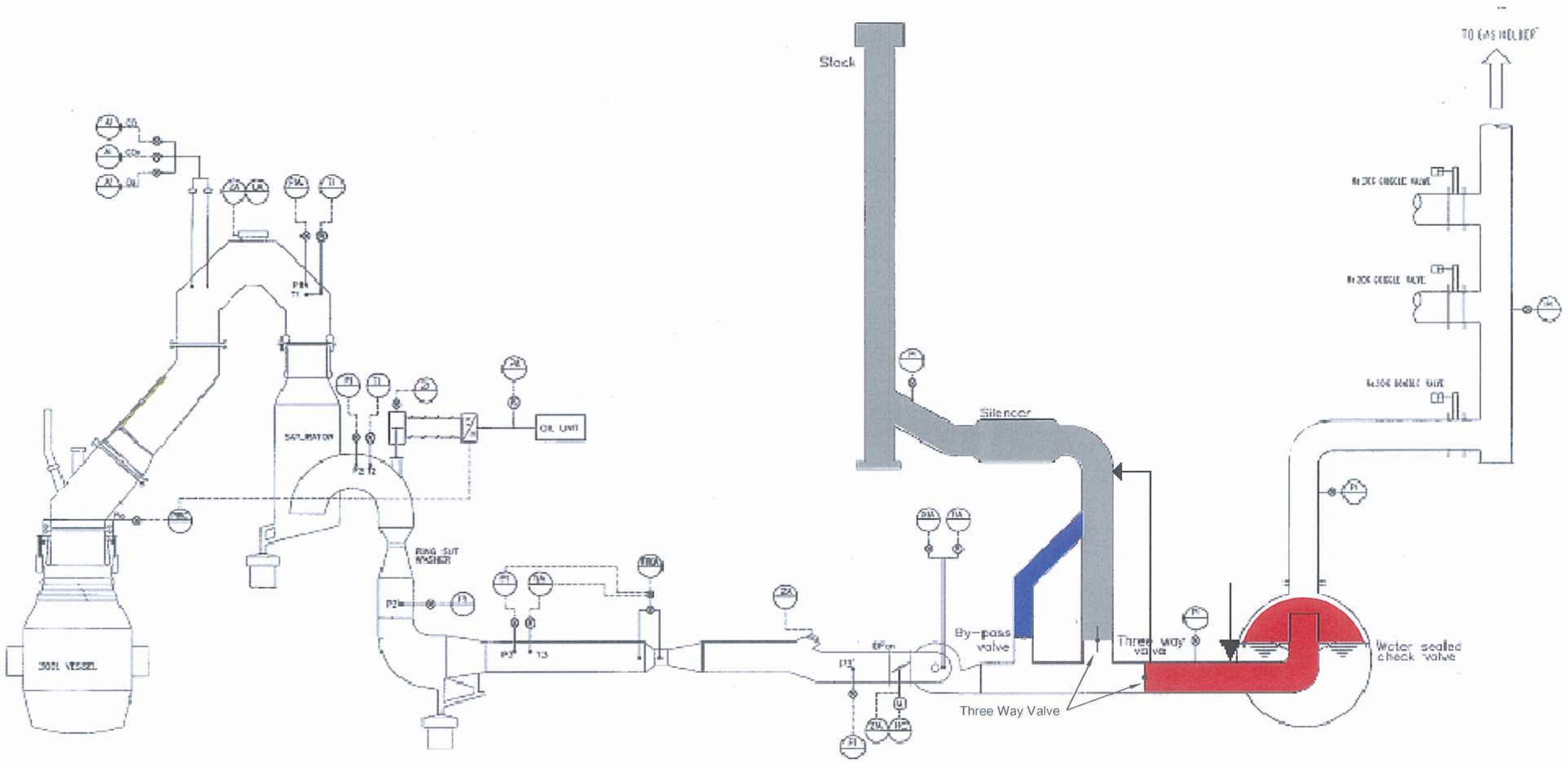
The BlueScope Steel preference would be to minimise the release of gas in an unburnt state to the atmosphere.

The collection main and holder will require purging when maintenance is required. Procedures will nominate purge rates and gas monitoring requirements to ensure excessive CO levels are not created. It is anticipated that LDG consumption will continue so as to reduce the holder level to approx 10% prior to start of purging.

When the supply main is to be purged, the connection to the gas bleeder will be opened and purge gas will be flared.

At this stage, the design is still considering options for bleeders. One approach is to allow the contents of the LDG supply main to be flared during the purging process via the No 6 BFG excess gas bleeder. A connection to this existing bleeder system would be installed to enable this. Alternately, a small maintenance bleeder (dedicated permanent or site mobile) could be provided.

The proposed nitrogen purge system is shown in **Figure 5**.



Key:

B1 Purge

B2 Purge

D Purge

Figure 5
Nitrogen Purge System

4.3 Supply of Coke Ovens, Blast Furnace and Natural Gas

At present, BFG and COG are collected and are partly used in the existing steam and electricity generation on-site, with excess gas being flared. The LDG is not collected, but is also flared. The production of by-product gases is variable due to the inconsistent rates of steel and iron production and the batch nature of production. In the cogeneration plant, variability in by-product gas production will be offset by the gas holders in the existing BFG and COG reticulation system, the new LDG gas holder and use of natural gas (if necessary to maintain boiler operation).

BFG is produced by the two blast furnaces at BlueScope Steel. Currently some BFG is used in the steelworks and the excess is flared. BFG will be consumed by the cogeneration plant as made available by the steelworks. The BFG will be supplied to the cogeneration plant by joining into a major BFG line that is located near the proposed plant.

COG is produced by the Coke Oven Batteries at BlueScope Steel and it is currently used by many processes at the steelworks. Any excess COG is currently flared. The COG will be supplied to the cogeneration plant by joining into an adjacent COG line that is located near the proposed plant.

Natural Gas will be supplied from the existing site main, located near the proposed cogeneration plant that provides natural gas for other uses at the site.

Table 1 provides the characteristic composition of the fuels to be used in the cogeneration plant. These figures were supplied by BlueScope Steel and assume that BFG, COG and LDG are saturated with water.

Table 1 Characteristic Composition of Cogeneration Plant Fuels

Component	Volume (%)			
	BFG	COG	LDG (BOS off-gas)	Natural Gas
Water	5.20	6.02	29.56	
Carbon Dioxide	19.91	2.43	12.22	0.66
Hydrogen	4.97	52.88		
Ethane		0.39		3.76
Oxygen		0.29	0.17	
Ethylene		1.94		
Propane				0.40
Nitrogen	48.02	5.82	10.34	3.46
Methane		24.26		91.55
Carbon Monoxide	21.90	5.82	47.71	
Benzene		0.03		
Butane plus		0.00		Balance
Ammonia		0.00		
Hydrogen Sulphide		0.07		
Hydrogen Cyanide		0.05		
Selected Properties				
Specific Gravity (Air=1)	1.04	0.38	Typical values - varies with process cycle	0.62
Calorific Value (MJ/m3)	3.2	19.8		39.3
LEL(%)	35	5.3		5
UEL(%)	74	34		15

4.4 Other Ancillary Chemicals / Equipment

A range of other chemicals (turbine oil, etc) will also be involved, but at such quantities and distance from site boundaries, that they clearly do not exceed the quantity and transportation threshold criteria of DoP (DUAP, 1997) or are not considered potentially hazardous in terms of land use safety planning.

The proposed development will contain process streams of water, nitrogen, and chemical mixtures. For the purposes of this study, incidents involving these materials were not considered to have the potential to result in significant land use safety planning impacts.

A range of chemicals (in relatively minor quantities) will be used for water treatment, cleaning, etc. These are still to be finalised, but the expected chemical types and quantities are summarised in **Table 2**. Most of the materials will be of the type defined as Dangerous Goods Class 8. Their storage and handling have limited potential for off-site harm, provided appropriate technical and management controls are observed (DUAP, 1999), and hence are not considered to present a land use safety planning concern.

Table 2 Expected Ancillary Chemical Types and Quantities for ICP

Plant Location	Material	Dangerous Good Class	Quantity of Site	Storage Mode
Demineralisation Plant	Hydrochloric Acid	8	22,750L	Liquid – Tanks
	Sodium Hydroxide	8	22,750L	Liquid – Tanks
	Sodium Bisulfite	8	1500L	Liquid – Tanks
Cooling Water Chemical Injection	Sulphuric Acid	8	10,000L	Liquid – Tanks
	Sodium Hypochlorite	8	10,000L	Liquid – Tanks
	Specialist Treatment Chemicals	8	4,500L	Liquid – Tanks
Boiler Feedwater	Hydrazine	3	2,500L	Liquid – Tanks
	Amine	8	2,500L	Liquid – Tanks
Hydrogen cooled generator	Hydrogen (compressed)	2.1	2 *15 * 6m3	Compressed Gas - Cylinders

Electrical generating equipment may typically have a hydrogen cooling system for the generator. Hydrogen conducts heat much better than air, making it a useful coolant. It also minimises windage losses given its low density. Cooling systems circulate the hydrogen through the generator shell for heat dissipation.

It is proposed to store one set of 15-off "G" size cylinders attached to the generator system to supplement hydrogen. As it is a pressurised system, hydrogen leaks will occur at the bearings and shaft seals, in the seal oil supply system, and from the H2 supply piping, mechanical and analytical equipment and hence the need for the supplement hydrogen supply. A second unconnected spare set will be used as a backup (for when the first set is almost empty and needs to be replaced).

Ambient gas detection will be used to detect hydrogen leaks.

The hydrogen usage is not considered to be potentially hazardous in terms of land use safety planning as it involves essential cylinder (rather than bulk) storage, and the distance to any sensitive land uses is quite significant (ie. the nearest residential location is approximately 1.5 km away, the harbour is approximately 400m).

Regardless, the Hydrogen System will be part of the HAZOP assessment to ensure the design is appropriate in terms of workplace safety.

5 HAZARD IDENTIFICATION

5.1 Introduction

Hazard identification involves examining possible initiating events, which can lead to hazardous incidents, and result in fatality. Hazard identification results in the definition of discrete scenarios, which represent the range of possible incidents and highlights the causes of such events.

5.2 Properties of Hazardous Materials

In general, material can be hazardous because it is flammable and/or toxic. While definitions of flammability (eg. lower flammability limit, etc) and impacts in terms of fire (heatflux) and explosion (overpressure) are relatively straightforward, toxicity levels require further definition.

Workplace exposure limits are sometimes referred to as Threshold Limit Values (TLV). TLV-TWA (Time Weighted Average) is defined as the time weighted average concentration limit for a normal 8-hour workday and 40 hours per week, to which nearly all workers may be repeatedly exposed, day after day, without adverse effect. Workplace levels do not provide sufficiently useful values for QRA as they relate to long term rather than atypical exposures.

Emergency Response Planning Guidelines (ERPGs) were developed under the guidance of a committee within the American Industrial Hygiene Association, and consist of three different toxicity levels for an assumed 1-hour duration. The values are specifically derived for planning and emergency response guidelines (Craig et. al., 2000)

The National Institute for Occupational Safety and Health (NIOSH) publish "Immediately Dangerous to Life and Health" (IDLH) values for various chemicals. **IDLH** is defined to mean conditions that pose an *immediate threat to life or health* or conditions that pose an immediate threat or severe exposure to contaminants which are likely to have an adverse cumulative or delayed effect on health. The IDLH concentration represents the *maximum concentration of a substance in air from which healthy male workers can escape without loss of life or irreversible health effects under conditions of a maximum thirty-minute exposure time.*

Other short-term exposure measures also exist for selected chemicals. One such approach is that used by the UK Health and Safety Executive (HSE, 2006[1]). The HSE use a level of toxicity in relation to the provision of land use planning (LUP) in terms of Specified Level of Toxicity (SLOT). HSE has defined the LUP SLOT as:

- Severe distress to almost everyone in the area.
- Substantial fraction of exposed population requiring medical attention.
- Some people seriously injured, requiring prolonged treatment.
- Highly susceptible people possibly being killed.

Fatality concentrations for other percentage levels can, in some cases, also be estimated from probit equations (where available). Probit equations provide an

[1] *http://www.hse.gov.uk/hid/haztox.htm*

estimate of the population fatality % for a given concentration and exposure duration to a chemical.

5.2.1 Blast Furnace Gas (BFG)

BFG is a by-product of the iron making process and is used as a fuel gas for some combustion systems on the site. It is an odourless, colourless and toxic gas. Its toxic properties are due to the presence of carbon monoxide (CO) (typically 22.3% v/v) in the gas.

BFG is a very low heating value fuel, containing inerts of approximately 58% nitrogen and 17% carbon dioxide. Therefore, the gas is only likely to support stable combustion at elevated temperatures, or with a permanent pilot flame. Site experience suggests that BFG may be ignited by a high ignition source such as a welding torch. However, the resulting combustion is slow and lazy. BFG is used as a fuel by enriching the gas with COG and/or natural gas to improve its burning characteristics.

BFG is not typically considered an explosion hazard for the following reasons:

- Very high ignition energies are required to initiate BFG combustion;
- High concentration (approximately 75% v/v) of inerts in the gas; and
- Very low combustion energy (3.2 MJ/m^3).

5.2.2 Coke Ovens Gas (COG)

COG is a toxic and flammable gas and has a very strong odour. Its toxic properties are due to the presence of CO (typically 9% v/v) in the gas. COG has a specific gravity of 0.38 and, therefore, is a very buoyant gas, which tends to disperse rapidly when released to the atmosphere.

The high concentration of hydrogen and methane in COG suggests that the gas can be ignited by a low ignition energy (e.g. static). Therefore, the probability of ignition of COG leaks is likely to be high relative to other flammable gases.

COG is a corrosive gas due to the presence of hydrogen and sulphides. This has significant implications for the maintainability of COG systems. COG pipework frequently develops small corrosion holes. Site experience indicates that one to three small (10mm) corrosion holes can occur at affected areas (ACARRE/BHP, 1991).

5.2.3 LDG

LDG is extremely toxic due to its high concentration of Carbon Monoxide (up to 90% dry basis) and its lack of detectable odour. The proportion of Carbon Monoxide (CO) in LDG varies dependant on the stage of the BOS blowing period. The specific gravity of dry LDG is approximately 1.1 and so may accumulate in pits or spaces. LDG contains CO, CO_2, N_2 and is saturated with water vapour.

LDG is flammable given its high level of CO. Flammability of LDG increases with oxygen enrichment and temperature increase. Ignition of a release of LDG may result in a flash fire. Auto ignition temperature of LDG is approximately 600°C.

As LDG is presently sent straight to flare from the process and not piped to other users, there is limited site experience with piping LDG at positive pressure to other consumers. However there is significant experience and familiarity with the hazards associated with LDG within the site given the experience with operating the BOS plant.

5.2.3 Carbon Monoxide

For BFG, COG and LDG, the principal toxicity is attributed to CO. CO is a colourless, odourless gas, which is also flammable (limits 12% to 75%). It has an auto-ignition temperature of 610°C. It is classified as a Class 2.3 (poisonous gas) 'Dangerous Good' as per the classification system presented in the 'Australian Code for the Transport of Dangerous Goods by Road and Rail'.

The health effects of CO are largely the result of the formation of carboxyhemoglobin (COHb), which impairs the oxygen carrying capacity of the blood. Resumption of the normal oxygen supply process takes place once an individual is removed from the contaminated atmosphere. However, any damage due to the prolonged loss of oxygen supply to the brain may not be reversible.

Table 3 (WHO, 1999) outlines the symptom-associated carboxyhemoglobin levels and the corresponding CO concentrations in the atmosphere. At a concentration of 1.3% it causes unconsciousness after a few breaths (Lees, 1996).

Table 3 Symptoms associated with varying levels of CO poisoning (WHO, 1999, adapted from Winter & Miller, 1976)

CO in the atmosphere			COHB in blood (%)	Physiological and subjective symptoms
%	mg/m^3	ppm		
0.007	80	70	10	No appreciable effect, except shortness of breath on vigorous exertion; possible tightness across the forehead; dilation of cutaneous blood vessel.
0.01	140	120	20	Shortness of breath on moderate exertion; occasional headache with throbbing in temples.
0.02	250	220	30	Decided headache; irritable; easily fatigued; judgement disturbed; possible dizziness; dimness of vision.
0.035-0.052	400-600	350-520	40-50	Headache, confusion; collapse; fainting on exertion.
0.080-0.122	900-1400	800-1220	60-70	Unconsciousness; intermittent convulsion; respiratory failure, death if exposure is long continued.
0.195	2200	1950	80	Rapidly fatal.

The TLV[2] for CO is 30 ppm.

ERPG-2 is the maximum airborne concentration below which it is believed that nearly all individuals could be exposed for up to one hour without experiencing or developing irreversible or other serious health effects or symptoms which could impair an individual's ability to take protective action. The ERPG-2 value for CO is 350ppm.

The IDLH for CO is 1,200ppm (NIOSH, 2006[3]).

The IDLH value was based on information that a 1-hour exposure to 1000 to 2000 ppm would cause unpleasant but no dangerous symptoms, but that 1,500 to 2,000

[2] *http://www.nohsc.gov.au/OHSInformation/Databases/ExposureStandards/az/Carbon_monoxide.htm*
[3] *http://www.cdc.gov/niosh/idlh/intridl4.html*

ppm might be a dangerous concentration after one hour. In general, a COHb level of 10-20% will only cause slight headaches and a COHb of 11-13% will have no effect on hand and foot reaction time, hand steadiness or coordination. At a COHb of 35%, manual dexterity is impaired. At 40% COHb, mental confusion, added to increasing lack of coordination, precludes driving an automobile. A 30-minute exposure to 1,200ppm will produce a COHb of 10-13% (NIOSH, 2000).

The SLOT value for carbon monoxide is:

(Concentration) x (time) = 40125 ppm.min

Two probit relationships are also reported for carbon monoxide.

Pr = -37.98 + 3.7ln(Ct) (CCPS, 2000)

where:

Pr = probit value (a normalised probability unit "probit", see **Appendix B**)

C = concentration (in ppm)

t = exposure time (in minutes)

and

Pr = 7.4 + ln(Ct) (TNO, 1999)

where:

Pr = probit value (a normalised probability unit "probit", see **Appendix B**)
C = concentration (in mg/m^3)[4]
t = exposure time (in minutes)

For example, using the CCPS probit equation, the 20% fatality concentration for 10, 20, 30 and 60 minutes exposure is 8836ppm, 4418ppm, 2945ppm and 1472ppm respectively, whilst the 10% fatality concentration is 7845ppm, 3923ppm, 2615ppm and 1307ppm respectively.

For the above probit toxic relationships, the following relationship applies (assuming a 20 minutes exposure, taken to be the upper limit for exposure time).

Probit Percentage	CCPS Concentration (ppm)	TNO Concentration (ppm)
10	2615	1956
20	2945	3038
30	3211	4183
40	3454	5481
50	3696	7037

Mannan (2005) notes that *"it is desirable to allow for the fact that data for probabilities in the middle of the probability range are likely to be more accurate than those for probabilities at the extremes of the range"*. Therefore, the probit relationship which was more conservative around the 50% range (ie. the CCPS probit) was used.

5.2.4 Carbon Dioxide

Carbon dioxide is a simple asphyxiant. At a concentration of 10%, it can cause unconsciousness within one minute (Lees, 1983). It is classified as a Class 2.2 (non-

[4] *Conversion factor for CO for mg/m^3 to ppm is 1.15*

flammable, non-toxic gas) 'Dangerous Good' as per the classification system presented in the 'Australian Code for the Transport of Dangerous Goods by Road and Rail'.

The TLV for carbon dioxide is 5,000ppm. The IDLH for carbon dioxide is 40,000ppm.

5.2.5 Natural Gas

Natural gas consists of 89% v/v methane, 8% v/v ethane and a balance of inerts (nitrogen and carbon dioxide). Therefore, it essentially exhibits the same physical and hazardous properties as methane.

Methane is a gas at normal temperature and pressure, but can be handled industrially in the liquid phase by the use of low temperatures, high pressures or both. It is a colourless, odourless, non-poisonous, flammable gas which burns with a pale, faintly luminous flame and can give rise to the hazards of fire and explosion (Lees, 1983; Merck, 1989). It has wide flammability limits of 5-15% v/v. Methane gas clouds are considered to be a low explosion hazard and only produce damaging overpressures if located in significant congested and/or confined locations. It has an auto-ignition temperature of 540 C. It is classified as a Class 2.1 (flammable gas) 'Dangerous Good' as per the classification system presented in the 'Australian Code for the Transport of Dangerous Goods by Road and Rail'.

5.2.6 Hydrogen

A relatively small quantity of pure hydrogen will be used for cooling the turbine generator.

Hydrogen is a highly flammable gas, which has a non-luminous flame. It is a colourless, odourless, tasteless gas which is flammable or explosive when mixed with air, oxygen, chlorine, etc (Merck, 1989). Hydrogen has wide flammability limits of 4%-75%, a low minimum ignition energy and a high burning velocity. It is therefore easily ignited and burns rapidly. As hydrogen flames are non-luminous, an operator may walk unaware into a hydrogen flame. As far as the hazard of an open-air explosion is concerned, hydrogen gas has a low density and it tends to rise and dissipate rapidly unless it is very cold. Nevertheless, vapour cloud explosions of hydrogen have occurred. Further, because of the uncertainty of the properties of hydrogen mixed with hydrocarbons, it is prudent to assume that a hydrogen-rich stream has the same hazard potential as one of pure hydrogen (Lees, 1996).

Hydrogen is classified as a Class 2.1 (flammable gas) 'Dangerous Good' as per the classification system presented in the 'Australian Code for the Transport of Dangerous Goods by Road and Rail'.

5.3 Process Hazards

The preliminary design was reviewed to identify potential major incidents that can result in fires, explosions and toxic releases. Hazardous incidents include:

- Jet fires from ignited gas leaks;
- Flash fires or vapour cloud explosions from delayed ignition of gas; and
- Toxic release of gas, which involves a percentage of CO.

5.2.4 Fire Hazards

The proposed development will handle a range of flammable gases with the potential to produce significant fires.

Releases of methane (and hydrogen, COG, LDG and possibly BFG) can result in fire upon ignition. When under higher pressures, ignition can result in severe gas fires (known as jet fires), which produce intense flames with high heat radiation intensities. Jet fires involve the ignition of high velocity gas or liquid flammable release, which are characterised by high momentum and good combustion condition. Jet fires are not only a major hazard for personnel, but can result in rapid escalation and domino incidents as a result of flame impingement on unprotected steel structures and equipment.

A flammable release of gas that does not ignite at the leak source, or has a delayed ignition, can produce a large vapour cloud, which covers a significant area if the gas is not dispersed. In the absence of significant confinement or obstruction, ignition of the cloud results in a low velocity flame front with minimal overpressure effects, known as a flash fire, and typically results (initially) only in impacts within the flammable cloud. After the gas cloud has burnt, a jet fire or more diffuse flame may remain.

5.2.5 Explosion Hazards

A gas explosion can occur where combustion of a premixed gas cloud (eg fuel-air) causes a rapid increase in pressure. If the gas cloud formed is outside the flammable concentration range (ie. outside the Lower Flammable Limit (LFL) and Upper Flammable Limit (UFL)), or an ignition source is lacking, no combustion will occur. The pressure generated by the combustion wave will depend on how fast the flame propagates and how the pressure can expand away from the gas cloud (governed by confinement).

In an accidental gas explosion of a hydrocarbon-air cloud (ignited by a weak source – a spark), the flame will normally start out as a slow laminar flame with low velocity. If the cloud is truly unconfined and unobstructed (ie no equipment or other structures are engulfed by the cloud), the flame is not likely to accelerate to higher velocities and the overpressure will be negligible if the cloud is not confined (Bjerketvedt, et. al., 1997)

However, flame speed is strongly modified by the presence of obstacles, congestion or partial confinement. Therefore, significant overpressures can begin to build up (provided flammable vapour remains) as the flame front continuously expands over and around obstacles. Flammable gas releases in confined and/or congested regions can result in explosions upon ignition.

5.2.6 Toxic Hazards

Toxic injury may arise from the effects of exposure to gases, vapours, dust, or liquids with the effects ranging from irritation through to fatality. The extent of impact from toxic exposure is dependent on the extent of spreading of the substance, the concentration at which the substance is still considered 'toxic' and the duration of exposure. For this assessment, release of CO can result in non-localised toxic impacts.

5.4 History of Major Incidents Worldwide

Difficulty exists in sourcing documentation on incidents relating to gas holders. Information from BlueScope Steel indicates the awareness of the following gas holder incidents:

- A major gas holder explosion occurring in Germany in 1923. This explosion resulted in significant damage and fatalities in surrounding areas. The cause of the explosion was due to incorrect maintenance work on the valving to the gas holder.
- A major holder fire occurred at British Steels South Teesside Works (UK) in 1971. The fire was caused by lightning striking a COG holder. The lightning apparently flashed across the shell to the piston, resulting in simultaneous seal rupture (grease seal), gas release and ignition. The piston was apparently not earthed.
- A major gas holder explosion at the Raurkela Steel Plant India in 1997, as a result of unauthorised testing of BOS fan during a maintenance outage. Air was introduced into the system and entered the gas holder, forming an explosive mixture that was ignited by the downstream electrostatic precipitator. The explosion blew off the roof of the gas holder and destroyed much of the gas recovery system.
- A major fire involving a Cylindrical Oil Seal Gas holder in Japan in 2003, due to incorrect protocols being followed on work being carried out on an adjacent piece of equipment causing a fire which spread to the gas holder.

The root causes of these scenarios will be addressed as part of the design process. For major maintenance work involving the gas holder, the inventory of flammable gas will be removed or reduced to as low as reasonably practicable and purged with nitrogen to reduce the flammability of the gas. The design of the gas holder will account for lightning strikes and ensure that appropriate earthing mechanisms exist and are maintained.

5.5 Potential Hazards

Hazard identification involves examining possible initiating events, which can lead to hazardous incidents, and result in significant consequences in terms of land use safety planning. Hazard identification results in defining discrete scenarios to represent the range of possible incidents, and highlight the causes of such events. A hazard identification table has been prepared (**Table 4**). This outlines potentially hazardous events associated with the existing operations and the proposed facility. The table does not mean that the hazards identified will occur, rather the potential exists for them to occur, and hence the need for mitigation measures.

The hazard identification table identifies the range of hazards identified in the risk assessment for which risk reduction strategies may need to be developed. These identified hazards are used to identify scenarios for which consequence modelling is performed and each scenario has a frequency associated with it. The consequence and frequency information is combined to give the risk result.

As only preliminary design information is available, the scenarios identified are intended to be representative rather than a complete list, with the focus being on mechanical rather than process failures. Additional safety studies that should be performed when the design has further progressed (eg. HAZOP, etc.), would be expected to identify in greater detail hazard scenarios (especially process failures) specific to individual plant items and locations.

Therefore, the scenarios identified in **Table 4** are not intended to include a complete list of all potential causes. However, the major scenarios identified will be those of most importance from a land use safety planning perspective.

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
A. COGENERATION PLANT				
A1	Boiler Explosion (explosion in combustion chamber)	• error during startup/shutdown • component/instrument failure • maloperation	• Internal explosion - - a major explosion could result in severe damage to boilers and surrounding plant from overpressure and boiler fragments - - given distance to site boundaries, not considered a land use planning safety issue	• Burner Management System (BMS) to contain a range of safeguards and redundancies to elimination, control and monitor this hazard • BMS to be designed to the appropriate Australian Standards and approved by NSW WorkCover • DCS (Distributed Control System) • safe shutdown systems as part of plant design philosophy
A2	Failure of high pressure steam lines	• corrosion • impact • overpressure / maloperation	• High pressure steam release – potential for injury in localised area from scalding (but not in terms of land use safety planning)	• water treated to reduce corrosion • pressure relief valves in system • Quality management systems (qualified welders, weld checks, etc) • operational and maintenance strategies to ensure safe operations • safe shutdown systems as part of plant design philosophy
A3	Steam Generator component failure	• loose turbine blade, overspeed, corrosion, erosion, fatigue/creep, impact, bearing failure • Water carry over	• Physical damage from objects such as rotor escaping containment – potential for injury in localised area (but not in terms of land use safety planning)	• automatic control and trip systems to be in place • operational and maintenance strategies to ensure safe operations • safe shutdown systems as part of plant design philosophy
A4	Hydrogen used in Generator for cooling	• Failure of pipework/equipment (generator is a pressure vessel) • failure of pipework/ equipment due to corrosion, impact, overpressure, human error, etc	• Release of high pressure hydrogen gas - potential for explosion / fire, but relatively small quantity and impact would be localised at generation location (not a land use safety planning issue)	• hydrogen to be stored in a well ventilated area to assist dispersion of any releases • hydrogen stored in standard cylinders rather than single vessel • ignition sources near hydrogen to be minimised. • Gas detection to be provided as appropriate

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
A5	Failure of cooling /demineralised water system, tanks, pipes, water side of heaters, condensers, deaerators, pumps etc	• impact, corrosion, fabrication, maintenance error, etc	• Release of water - - possible issue with environmental impacts of treatment chemicals, but quantity, concentrations and dilution by time offsite would be sufficient to ensure not a significant environmental impact in terms of land use safety planning • Failure of operations due to lack of water	• intend to have monitoring of water levels • safe shutdown systems as part of plant design philosophy
A6	Transformer failure resulting in a release of oil	• external flange / component failure due to corrosion, impact, etc • operational issue (eg. overheating)	• Release of oil, possible ignition and fire. Bunded area designed to contain and spills so that fire would be localised (not a land use safety planning issue)	• termination box location and explosion protection (eg blast wall, segregation, orientation) to be considered as part of the design • bunded area to contain release • deluge system to be installed
A7	High Voltage switchboard failure	• equipment or component failure • lightning strike / storm damage	• Arcing or local explosion (not a land use safety planning issue)	• fire detection to be installed in switchroom, segregation in switchrooms to be carefully considered in design
B. OG RECOVERY SYSTEM (LDG incidents)				
B1.	Major release of gas due to pipeline failure	• corrosion, impact • overpressure, human error (working on wrong equipment)	• Release of gas - - given gas characteristics, potential fire risk, unlikely to be a explosion given need significant level of confinement • Possible toxic impacts as due to CO content	• fixed CO gas monitoring to detect leaks • operating and maintenance strategies • Pressure monitoring system upstream/downstream of gas holder • Inject nitrogen for maintaining positive pressure • safe shutdown systems as part of plant design philosophy

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Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
B2.	Minor release of gas due to seal pot blowout	• overpressure, human error	• Release of gas - - not considered to be a significant land use release scenario • Possible localised toxic impacts due to CO content	• fixed CO gas monitoring to detect leaks • operating and maintenance strategies • seal pot design to try and prevent blowout • Pressure monitoring system upstream/downstream of gas holder • Inject nitrogen for maintaining positive pressure
B3.	Failure of LDG line to / from gas holder	• corrosion, impact • overpressure, human error (working on wrong equipment) • large vehicle impact	• Release of gas - - given gas characteristics, potential fire risk, unlikely to be a explosion given need significant level of confinement • Possible toxic impacts due to CO content	• pipe and valve identification to be used • trip condition on BOS furnace switching stations (send to flare stack) • protection barriers at vulnerable locations
B4.	Failure of LDG line to / from gas holder and air ingress into line	• Incorrect operational procedures • Overriding control systems • Vehicle impact with pipe supports and/or mechanical damage (ie tray of truck left in raised position)	• Release of gas - - given gas characteristics, potential fire risk, unlikely to be a explosion given need significant level of confinement • Possible toxic impacts due to CO content	• Training / staff skill such that overriding control systems is not considered credible • Proposed control system will provide levels of protection prior to ability to override controls • Ability to send LDG to flare and isolate flare from downstream system during emergency • protection barriers at vulnerable locations
B5	Failure of ID Fan, By-pass valve or three-way valve	• corrosion, impact, mechanical failure • operational error, control system failure	• Reverse flow, ingress of air, leakage of CO - possible fire / explosion in pipe, release of unburnt CO could provide a toxic hazard	• interlock ensures that only one three-way valve is operating at a time, other two valves are set to flare, trip on BOS to stop blow. • Goggle valve can be tripped • Position indication on 3 way valve • Automatic integrity checks • safe shutdown systems as part of plant design philosophy

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
B6	Failure of flare stack	• loss of pilot flame • corrosion, impact • operational error	• Release of unburnt CO	• flare monitoring devices • ability to stop BOS blow if necessary (LDG generation is a controllable intermediate process)
B7	Failure of flare stack	• blockage in the silencer	• increase in pressure	• detection in pressure control system
B8	Power failure	• site or localised power failure	• All valve controls fail • Loss of switch rooms • Possible loss of gas control • dead legs result in residual flammable gas still available in dead legs	• Water seal valve closes (driving force on valve is hydraulics, sufficient hydraulic storage to shut valve without power exists) • Manually/automatically close of goggle valves • Nitrogen purge system • Emergency restart procedures will be developed • Operation can go to flare if concerns over downstream system • safe shutdown systems as part of plant design philosophy
B9	Change in Flare operation from continuous to sequence process associated with LDG capture process	• Can't re-ignite gas at top of flare	• Cloud of low CO % released from stack for short period	• Flare monitoring system • Can't re-ignite because CO levels already low (hence less of toxic concern), and release time relatively short • Flare pilots / ignition system
B10	Failure of Water Seal check valve	• corrosion, impact • operational error • high pressure slug • Stuck valve	• Reverse flow release of stored LDG • Stuck valve (detected in control sequence) • Stops collecting	• range of devices (measuring, detection) are used to ensure water seal is maintained • considering system of automatically operated nitrogen purge and close goggle valve • position detector on water seal valve
B11	Failure of Booster Fan	• corrosion, impact • operational error	• Gas holder surge capacity exceeded	• Design will involve consideration of linked permissive on three way valve to regulate excess flow • duty / stand-by for fans • gas holder has pressure relief system
C. GAS HOLDER (LDG incidents)				

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
C1	Gas holder - Damage / deterioration of rubber seal	• natural degradation over time from wear, etc • temperature of gas – with no gas cooler temperature of gas is higher and closer to seal design limit	• Release of LDG through seal failure • Puncture of seal from seals, construction of seal suggests expect small rips in seams are possible	• formalised operating and maintenance strategies • CO monitoring stations in and around gas holder • routine checks on hardness of seal • seals design involves lamination, long rips unlikely due to manufacture process • Gas Detection at top of gas holder • Monitoring seal condition
C2.	Gas holder – high gas holder piston level	• LDG production rates exceed consumption rate	• Over pressurisation of gas holder – potential of rupture or holder pushing through roof	• high piston level limit switch • a mechanically operated pallet type vent valve is opened by the piston when it reaches the top position thereby venting gas into the atmosphere • inlet to gas holder will have isolation valve (interlocked with piston High-High level alarm) • pressure is monitored • excess gas can be flared • high design pressure for gas holder • when piston reaches high-high limit gas collection will cease so no more gas enters holder
C3.	Gas holder – low gas holder position level	• LDG consumption rate exceeds production rate • major leak or valve failure in LDG network	• Air ingress to gas holder with potential of confined explosion	• water seals can be initiated on the gas holder outlet lines • when the piston reaches the low or empty position any further discharge of gas will pull a grid seal over the inlet and outlet connection and act as a shut-off valve

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
C4.	Piston fall/rise rate exceeds design capacity	• Consumption rate greatly exceeds production rate • Major leak in LDG network	• Possible damage to gas holder, and damage to seal	• limit speeds to be set on gas holder piston • shorter, thicker gas holder means expected piston rate significant less than design maximum • maximum piston speed 4 – 6 m/s • boiler control system manages piston speed
C5.	Piston movement is hindered by sticking/tilting	• Deteriorating seal • Seal jamming • Incorrect initial set-up of alignment during commissioning	• Piston may tilt resulting in damage to seal with the potential for air ingress and confined explosion	• weigh system used to stabilise piston (ie levelling system) • piston weigh such that highly unlikely that holder pressure can fall below atmospheric • guides are not required for the Wiggins Dry Seal Gas holder. The need for close tolerances and constant inspection to prevent jamming or binding of the structure is therefore eliminated • excess tilt initiates safety shutdown
C6.	Catastrophic failure of gas holder	• air ingress leading to catastrophic failure by explosion • incorrect procedure (incorrect purge) • rubber seal failure • Gas Holder base corrosion • vehicular collision	• Release of gas-given gas characteristics, potential fire risk, unlikely to be a explosion (if confined) unless at elevated temperature or a high energy ignition source is present. • Possible toxic impacts due to CO content	• gas holder inventory can be isolated by shutoff valve • CO gas detection • gas holder and seal maintenance program • expect control system to have a range of detectors/interlocks to ensure safe operation • proposed gas holder location is remote from normal vehicle traffic as well as from steelmaking and slag handling operations • design to be such that vehicles not permitted near holder
C7.	Gas leak from gas holder vessel (non-seal portion)	• corrosion, impact	• Toxic vapour release	• CO monitors to be installed at perimeter of gas holder • Thick shell design for support rather than internal pressure (which is relatively low)

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
C8.	Formation of explosive space above gas holder system	• Leak in seal	• Flash fire/explosion	• Gas monitoring system above the piston • Area is not fully enclosed
D. Supply of Blast Furnace Gas (BFG)				
D1.	Release of gas due to pipeline or seal pot failure/blowout	• corrosion, impact • overpressure, human error (working on wrong equipment)	• Release of gas-given gas characteristics, unlikely to be a fire or explosion (if confined) unless at elevated temperature or a high energy ignition source is present. • Possible toxic impacts as 22% CO	• fix gas monitoring to detect leaks • operating and maintenance strategies • water seal design
D2.	Failure of line and air ingress into line	• corrosion, impact • overpressure, human error (working on wrong equipment)	• Given gas characteristics, unlikely to be a fire or explosion (if confined) unless at elevated temperature or a high energy ignition source is present	• Isolation procedures maintain positive system pressure (prevent air ingress)
E. Supply of Coke Ovens Gas (COG)				
E1.	Release of gas due to pipeline or seal pot failure/blowout	• corrosion, impact • overpressure, human error (working on wrong equipment)	• Release of gas-given gas characteristics (50% H2, 25% Methane), likely to be a fire or explosion (if confined).	• fix gas monitoring to detect leaks • operating and maintenance strategies • water seal
E2	Failure of line and air ingress into line	• corrosion, impact • overpressure, human error (working on wrong equipment)	• Given gas characteristics, likely to be a fire or explosion (if confined)	• Isolation procedures maintain positive system pressure (prevent air ingress)

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
F. Supply of Natural Gas				
F1.	Minor releases	• Leak or failure of pipework or associated equipment (eg wear at valve gland) due to corrosion, fatigue, impact, fabrication error)	• Flammable gas release, possible fire, little potential for explosion unless release into confined/congested area	• design to consider plant layout and pipe runs to be such to minimise confinement / congestion of any releases • consideration to be given to fixed gas monitoring
F2.	Major release: Pipe failure	• Leak or failure of pipework or associated equipment (eg wear at valve gland) due to corrosion, fatigue, impact, fabrication error) or maloperation issues (eg: overpressurisation)	• Flammable gas release, ignition and fire (jet fire if immediate ignition, flash fore if delayed ignition, explosion if release occurs in confined/congested area)	• design to consider plant layout and pipe runs to be such to minimise confinement / congestion of any releases • consideration to be given to fixed gas monitoring, emergency response shutdown and isolation of gas supply
F3.	Working on wrong equipment	• insufficient signage • piping/equipment congestion • human error	• Flammable gas release	• Procedures, signage, lockout isolation requirements, colour coding of pipework
F4.	Explosion in combustion chamber	• Burner management system failure (valve passing)	• Explosion	• burner management system
G. Boiler Treatment and other speciality chemicals				
G1.	Tank Leak	• Tank failure from overfilling, corrosion, impact	• Possible release to soil and stormwater system so eventually could lead off-site. Impacts varying depending on chemical toxicity (eg: sodium hypochlorite is a biocide, acid or caustic soda change pH conditions, etc)	• bunded area • operator training / audits

Table 4 Land Use Safety Planning Hazard Identification

No.	Functional/Operational Area – Hazard Event	Possible Initiating Events (Causes)	Possible Consequences	Prevention/Protection Measures
G2.	Tank delivery	• incorrect material loaded to tank • tank over-fill and overflow	• Toxic gas release • Release to bund	• Operator training and clear signage will be installed
H. Other System Support Failures				
H1	Power Failure	• Various	• Equipment cannot operate, possible hazard scenario	• system will be designed such that it fails safe when a power failure occurs
H2	Instrument Air Failure	• power failure, interworks air compressor failure	• No instrument air supply to actuate valves etc	• instrument failure scenario to be part of design considerations
H3	High Voltage switchboard failure	• Equipment or component failure	• Arcing or local explosion	• fault condition circuit response (ie. breakers) • fire detection in switchroom • only HV qualified personnel permitted in HV areas
H4	Chemical Tanker / Delivery truck drive-away incident	• Truck driven away while still connected	• Loss of containment	• appropriate signage and training to be provided • Design to consider appropriateness of bunding and camlock fittings or other drive-away isolation techniques

6 CONSEQUENCE ANALYSIS

Consequence analysis involves the estimation of the potential effects of accident events by the use of knowledge, experience, theoretical models, logic models or a combination of these methods. The analysis involves three distinct steps:

- Identification of the types of consequences that may occur for each accident event by examining the different sequences of events that can occur.
- Estimation of the physical effects of each event in terms of thermal radiation, explosion overpressures, toxic dose or physical impact.
- Translation of these physical effects into fatality estimates taking into account factors such as personnel exposure and distribution, potential escalation and evacuation success.

Consequence analysis also assists in identifying potential land use safety hazards which do not have impacts outside site boundaries (and hence do not represent a land use safety planning hazard regardless of their frequency). Where hazard scenarios have been identified to have potential impact outside site boundaries, these have been further addressed in the Frequency Analysis Section.

The release of a gas into the plant environment can result in a range of possible outcomes. The outcomes and associated consequences of such a release is dependent on a number of factors, including:

- If and when the release ignites;
- Properties of the gas;
- Location of the release; and
- Magnitude and duration of the release.

Immediate ignition of a high-pressure flammable gas release may result in a jet fire. If the gas leak does not ignite for a period of time (known as delayed ignition) a flammable vapour cloud may develop. The size of this cloud would be dependent on the leak magnitude and duration, gas properties and leak location environment. The ignition of the flammable cloud has the potential to result in an explosion and/or a flash fire.

A gas explosion is characterised by a very rapid combustion (within milliseconds) of a flammable vapour cloud producing damaging overpressures. The potential of a vapour cloud to explode upon ignition is dependent on the properties of the gas and the degree of congestion and confinement of the cloud location. If the ignited cloud does not explode, a flash fire will still occur. A flash fire is the name given to the rapid combustion (tenths of seconds) of a flammable vapour cloud. Delayed ignition of a flammable vapour cloud, whether it explodes or not, will also result in a jet fire as the flame will flash back to the leak source.

Unignited gas releases can still pose a hazard to personnel if they contain toxic components.

Based on the hazard identification step, discrete representative fire, explosion, and toxic hazard scenarios were selected for further analysis of their potential impacts.

As noted by CCPS (2000), consequence models have uncertainties, which can arise from issues including an incomplete understanding of the release (i.e. hole size) and poor understanding of the release process. Uncertainties are addressed by assigning conservative values to the unknowns, resulting in an upper case estimate of the consequence envelope. Unfortunately, as CCPS note:

"This procedure can result in a consequence that is many times larger than the actual, leading to a potential overdesign of the mitigation procedures or safety systems. This occurs, in particular, if several decisions are made during the analysis, with each decision producing a maximum result. For this reason, consequence analysis should be approached with intelligence tempered with a good dose of reality and common sense".

For this analysis, only major process hazards from the major process flows were considered. It is acknowledged that smaller process lines, because of their location, could result in significant local damage, but not to the level of those incidents required to be a possible land use safety planning concern (i.e. impact outside site boundaries).

The mathematical models used to determine the consequence distances for each hazard scenario are outlined in **Appendix B**. It should be noted that many of the hazard scenarios postulated have yet to occur in chemical plant history, and some many also have not been experimentally performed. Therefore, the generalised consequence models used have not been fully validated for most of the release conditions of interest. Whilst the effect distances are calculated from models considered to be appropriate, they should be taken only as providing indicative distances to impact levels (ie. heatflux, overpressure, toxic load) of concern.

The main inputs for each mathematical model, as well as the criteria for plant damage, fatality and injury, are shown in **Table 5**. For toxicity, the two injury levels exist as required by the DoP (see **Appendix A**), these being 'seriously injurious' and 'irritation'.

For fatality and injury to plant personnel, the values provided are typically well established for hazard analysis by regulatory authorities. A representative value was selected to assist in determining the extent of material damage based on research by TNO (1992).

Table 5 Framework for Consequence Modelling

Hazard Type	Failure type where applicable	Model inputs	Consequence levels used for this study		
			Damage	Fatality	Injury
Heat radiation from fires	failure mechanism results in release of high velocity gas release	• hole diameter • discharge rate	**23 kW/m²** • unprotected steel will reach thermal stress temperatures which can cause failures (DoP, 1992a) • damage to steel such as discolouration, peeling off of paint, and/or appreciable deformations of structural elements (TNO, 1992)	**12.5 kW/m²** • Significant chance of fatality for extended exposure. High chance of injury (DoP, 1992a) • chance of fatality for unprotected person exposed for 30 seconds (Pieterson, 1990)	**4.7 kW/m²** • will cause pain in 15-20 seconds and injury after 30 seconds exposure (at least second degree burns will result) (DoP, 1992a)
Heat radiation from flash fires	failure mechanism results in release of gaseous flammable flow	• source term parameter • atmospheric conditions	Combustible materials damage within cloud will be practically complete. Material damage outside cloud is negligible (TNO,1992)	Fatality within the LFL (lower flammability limit) is 100%. Outside the LFL, fatality is 0%	Assumed that injury within the LFL (lower flammability limit) is 100%. Outside the LFL, injury is 0%
Overpressure from vapour cloud explosions	failure mechanism results in release of gaseous flammable flow	• flammable cloud mass • congestion estimate • gas cloud properties	**14kPa** • Regulatory accident propagation criteria (DoP, 1992a) • Agreement with TNO (1992) generalised comparisons:(8kPa: - minor damage to steel frames, 20kPa: - collapse of steel frames, slight pipebridge deformation, cracking in empty oil-storage tanks)	**21kPa** • 20% chance of fatality to a person in a building (DoP, 1992a) • 25% chance of fatality and injury (IChemE, 1989)	**7kPa** • probability of injury is 10% (no fatality) (DoP, 1992a)
Exposure from release of toxic material	vessel with gaseous inventory pipe with gaseous toxic flow	• source term parameters • atmospheric conditions	Not applicable for atypical process safety incident. For this assessment, environmental risks from atypical release of CO not expected to have any impacts	Maximum time of exposure taken as 20 minutes, based on engineering judgment – it is considered most unlikely that a continuous release is not isolated or controlled, or potentially affected people are not evacuated and have full exposure greater than this timeframe. Probit equation (for 100% CO) gives 5550ppm, 4800ppm, and 3900ppm for fatality probability of 0.5, 0.3 and 0.1	CO 'Seriously injurious' **1200ppm**: IDLH value (DoP, 1992a) CO 'Irritation' **350ppm** (ERPG-2 value, WHO reported effects)

7 SAFETY MANAGEMENT

A PHA may involve use of quantitative risk assessment (QRA) techniques to assess the risks to the public from a potentially hazardous facility. QRA's have traditionally employed generic failure rate data to estimate the likelihood of hazardous incidents. In using generic data, it is assumed (or implied) that equipment and systems are operated, maintained, and managed at standards equivalent to the industry average. However, it is generally accepted that the magnitude and frequency of hazardous incidents is dependent on a range of safety management factors, which in most cases, cannot be adequately quantified within a QRA framework.

The DoP require that a PHA provide an outline of organisational safety controls (ie. safety management systems). In some cases, implementation of a safety management system may be a formal requirement of the consent for development approval. The DoP has released guidelines (DoP, 1995) which describe safety management principles and their implementation in a formal Safety Management System.

BlueScope Steel has established, implements and maintains an Occupational Health and Safety (OH&S) Management System for the site (BlueScope, Feb 2005).

In addition to its Health, Safety, Environment and Community Policy, BlueScope has 14 Safety Management Standards, which contain the guiding principles in terms of the intent and key performance requirements. The standards apply for:

-
- Accountability
- Legal Compliance
- Risk Management
- Fit for Work
- Training and Competency
- Engagement, Consultation and Communication
- Document Control, Records Management
- Materials and Contractor Management
- Project Management
- Process, Plant and Equipment Integrity
- Emergency Preparedness and response
- Incident Management
- Preventative and Corrective Action
- Measurement and Verification

In addition, given the pending implementation by NSW WorkCover of Major Hazard Facilities (MHF) legislation, BlueScope has developed standard management requirements for High Risk Facilities (HRF's). A HRF is any facility that has the potential for a sudden process related event leading to any Level 5 Consequence on the BlueScope Steel Consequence Matrix. The 14 Safety Management Standards provide the basis for managing an HRF, and additional requirements are detailed in the HRF Audit Checklist. The proposed LDG gas holder and power plant will form part of existing HRF's and will therefore comply with all BSL's HRF requirements.

8 FREQUENCY ANALYSIS

Frequency analysis involves the estimation of the likelihood of hazardous events and the likelihood of impacts from such events. This may involve estimating failure rates or failure probabilities of equipment items, operators and protection systems. These frequencies are obtained from site historical data, published failure rate data and engineering judgement.

A major issue with obtaining relevant failure rate or historical data is its applicability to the system under analysis. Although plant-specific failure rate data for the specific equipment under investigation is preferred, this information is often not available even for mature existing plants.

The failure of pipes, vessels, and other plant equipment can be due to a range of factors such as corrosion, poor maintenance, maloperation, etc. The potential for such factors to cause equipment failures can be eliminated or mitigated by effective safety management systems.

Generally, it is not possible to adequately quantify the effects of safety management systems on the failure frequency of equipment. Therefore, risk analyses rely on generic failure rate data (some of which is available in the public domain) to estimate the likelihood of hazardous events.

The type of weather condition has a significant effect on the material dispersion resulting from a gas release. The probability of the direction of the wind and the type of weather conditions (i.e. wind speed and weather stability category) play an important role in determining the overall risk value. Appropriate wind speed and weather stability data is required for the dispersion modelling. Determining the frequencies of various wind speed/weather stability category combinations for each direction is necessary for estimating the toxic effect risks at points in each direction.

The base failure rate and probability data used for the analysis is summarised in **Appendix D**.

9 RISK ASSESSMENT

Quantified Risk Assessment involved the estimation of the frequency of target levels of impact to various land uses. The risk levels are calculated for each event by combining the results of the consequence analysis and the frequency analysis. When the risk from all events identified are combined, a cumulative risk result is produced. The quantified risk result can be judged against adopted criteria for surrounding land uses of interest. The risk criteria adopted for this study are those recommended by the New South Wales DoP in "Hazardous Industry Planning Advisory Paper No.4 (DoP, 1992a)" and have been outlined in **Appendix A**.

With regard to risk assessment, the "Guidance Notes on Implementation" within "Hazardous Industry Planning Advisory Paper No.4 - Risk Criteria for Land Use Safety Planning" require that:

- The implementation of the criteria must acknowledge the limitations and in some cases the theoretical uncertainties with risk quantification.
- The criteria are best implemented when used as targets rather than absolute levels. Nevertheless, any substantial deviations from such targets should be fully justified.
- It is advisable that in all cases the assessment process emphasises the hazard identification and risk quantification process and procedures, rather than relying on absolute risk levels.
- Given the probabilistic nature of the assessment process, care must be exercised in interpreting/assessing compliance with a risk criteria in designating plants which exceed the suggested criteria as "unsafe". Nevertheless a higher resultant risk level relative to the suggested criteria indicates possible land use safety incompatibility and locational safety constraints.

The scope of this Preliminary Hazard Analysis is the calculation of risk associated with the proposed cogeneration project and not the cumulative risk associated with the BlueScope Steel site. If it can be shown that the risk is significantly lower than the risk criteria, then the contribution of the land use safety risk would not be a significant to any cumulative land use safety risk considerations.

It should be noted that part of this proposal will involve the closure of No.1 Power House located towards the nearest residential area at the south-west of the site. Therefore, the risk associated with this section of plant will be eliminated.

The calculated land use safety planning risk is a function of the severity (consequence) and the likelihood (frequency) of potential hazardous incidents occurring. The level of risk at any particular point is normally expressed as the sum of all risks from causes (scenarios) originating from a facility expressed over a one-year period. A summary of the information used to determine the risk contours is shown in **Appendix E**.

The contours for individual risk of fatality for risk levels of 1 in a million per year and 0.1 in a million per year are shown in **Figure 6.** These contours represent the land use safety planning risks associated with the proposed new facilities associated with the cogeneration plant and not the cumulative contour from the BlueScope Steel site.

In terms of individual fatality risk criteria (**Appendix A**), the residential criteria of 1 in a million per year is not exceeded. In fact, the individual risk fatality level at the nearest residential area is at least an order of magnitude less (ie: less than 1 in 10 million per year). With due consideration to the conservative nature of the assessment; and the technical, operational and locational controls which, although most difficult to quantify, would clearly reduced the actual risk levels, the actual risk levels would be significantly less than the calculated risk levels. Therefore, it is concluded that the proposed development does not significantly increase the land use safety planning risk outside the site boundary, and also satisfies the DoP's Individual Fatality Risk Criteria (**Appendix A**).

In terms of societal risk, it is noted that:

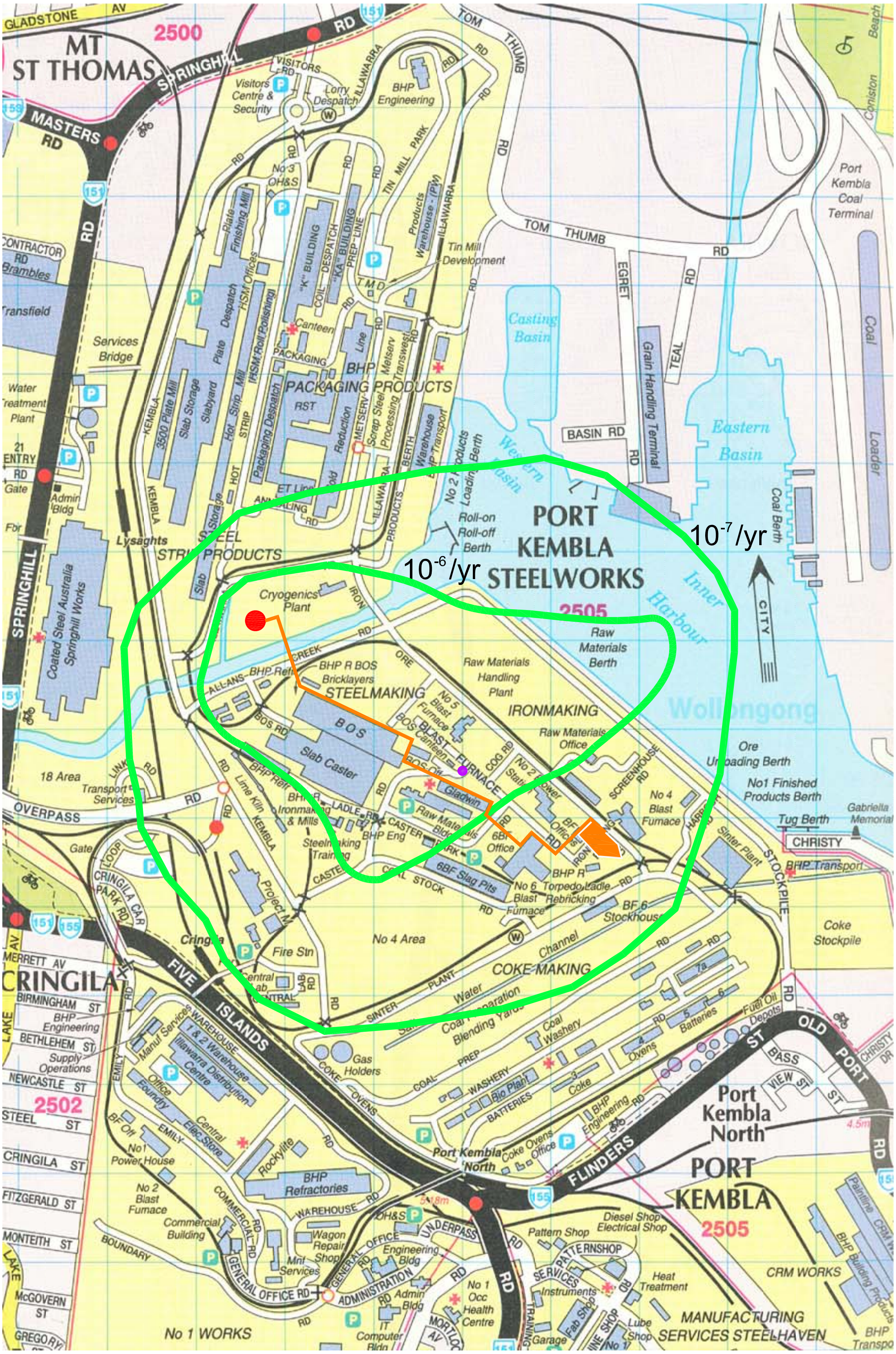
- The potentially hazardous components of the proposed facility is located well away from any major high density residential area or sensitive land uses (at least 1000m);
- The individual fatality risk levels at the nearest residential level are at least an order of magnitude below the associated individual risk criteria;
- Only under the extremely conservative modelling conditions of F1 weather conditions have fatality concentrations been calculated outside site boundaries; and
- The actual number of obstacles between the release location and nearest residential area are such that they would be expected to greatly interfere with the stability of any stable gas release, resulting in additional dispersion dynamics such that it would be expected the dispersion would be similar to less stable weather conditions which do not result in calculated fatalities outside site boundaries.

Therefore, it is concluded that the proposed facility will not be unacceptable on societal risk grounds.

The contours for injurious and irritant toxic levels are shown in **Figure 7** and **Figure 8** respectively. The respective criteria of 10 and 50 in a million per year are not exceeded. Therefore, it is concluded that the proposed development satisfies the DoP's Injury Risk Criteria (**Appendix A**)

As no hazard scenario results in heatflux or overpressure levels exceeding 4.7kW/m^2 and 7kPa at site boundaries, the Injury Risk Criteria (**Appendix A**) is satisfied. This also implies that the Risk Criteria for Property Damage and Accident Propagation (**Appendix A**) is satisfied.

As no accidental emission was identified which would threaten the long-term viability of any ecosystem or any species within it, it is considered that the intention of the DoP's risk criteria for the biophysical environment (**Appendix A**) is satisfied.



- Proposed Location of ICP

- Proposed Location of Gasholder

Figure 6
Individual Fatality Risk Contours

- Proposed Location of ICP

- Proposed Location of Gasholder

Source: Gregory's Maxi 2003 Street Directory 15th Edition

Figure 7
Injury Risk Contours

- Proposed Location of ICP

- Proposed Location of Gasholder

Figure 8
Irritation Risk Contours

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Appendix A
Risk Criteria

APPENDIX A – RISK CRITERIA

The DoP criteria (DoP, 1992a) for land use planning are:

Individual Fatality Risk Levels

- Hospitals, schools, child-care facilities and old age housing development should not be exposed to individual fatality risk levels in excess of half in one million per year (5×10^{-7} per year).

- Residential developments and places of continuous occupancy, such as hotels and tourist resorts, should not be exposed to individual fatality risk levels in excess of one in a million per year (1×10^{-6} per year).

- Commercial developments, including offices, retail centres, warehouses with showrooms, restaurants and entertainment centres, should not be exposed to individual fatality risk levels in excess of five million per year (5×10^{-6} per year).

- Sporting complexes and active open space areas should not be exposed to individual fatality risk levels in excess of ten in a million per year (1×10^{-5} per year).

- Individual fatality risk levels for industrial sites at levels of 50 in a million per year should, as a target, be contained within the boundaries of the site, where applicable.

Injury Risk Levels

- Incident heat flux radiation at residential areas should not exceed 4.7kW/m^2 at frequencies of more than 50 chances in a million per year.

- Incident explosion overpressure at residential areas should not exceed 7kPa at frequencies of more than 50 chances in a million per year.

- Toxic concentration in residential areas should not exceed a level which would be seriously injurious to sensitive members of the community following a relatively short period of exposure at a maximum frequency at 10 in a million per year.

- Toxic concentration in residential areas should not cause irritation to eyes or throat, coughing or other acute physiological responses in sensitive members of the community over a maximum frequency of 50 in a million per year.

Risk of Property Damage and Accident Propagation

- Incident heat flux radiation at neighbouring potentially hazardous installations or at land zoned to accommodate such installations should not exceed a risk of 50 in a million per year for the 23 kW/m^2 heat flux level.

- Incident explosion overpressure at neighbouring potentially hazardous installations, at land zoned to accommodate such installations or at the nearest public buildings should not exceed a risk of 50 in a million per year for the 14kPa explosion overpressure level.

Societal Risk Criteria

- Judgements on societal risk be made on the basis of a qualitative approach on the merit of each case, rather than on specified set numerical values.

Criteria for Risk Assessment to the Biophysical Environment

- Industrial developments should not be sited in proximity to sensitive natural environmental areas where the effects (consequences) of the more likely accidental emissions may threaten the long-term viability of the ecosystem or any species within it.

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

Appendix B
Consequence Methodology

APPENDIX B - CONSEQUENCE METHODOLOGY

B1. Gas Outflow

Discharges from vessels and pipes containing gas under pressure are normally readily calculated using standard equations for gas flow (World Bank, 1988).

The first step in the calculations is to determine whether the flow is critical, i.e. sonic or choked, or sub-critical. The distinction is made as follows assuming reversible adiabatic expansion and ideal gas behaviour:

Flow is critical if:

$$P_a < P_1 \left[\frac{2}{y+1}\right]^{\frac{y}{(y-1)}}$$

Flow is sub-critical if:

$$P_a > P_1 \left[\frac{2}{y+1}\right]^{\frac{y}{(y-1)}}$$

Where:
P_1 = upstream pressure (Pa)
P_a = atmospheric pressure Ps
γ = specific heat ratio

The second step is to calculate the discharge rate using the following formula:

$$Q = Y\ C_d\ A_r\ P_1 \sqrt{\left[\frac{M\gamma}{RT_1}\right]\left[\frac{2}{y+1}\right]^{\frac{(y+1)}{(y-1)}}}$$

Where:
Q = release rate (hg/s)
Ar = area of release
M = Molecular weight
R = Gas Constant
C_d = Discharge coefficient

For critical outflow:
Y = 1.0

For sub-critical outflow:

$$Y = \left[\frac{P_a}{P_1}\right]^{\frac{1}{y}} \sqrt{\left[1-\left\{\frac{P_a}{P_1}\right\}^{\frac{(y-1)}{y}}\right]\left\{\frac{2}{\gamma-1}\right\}\left\{\frac{\gamma+1}{2}\right\}^{\frac{\gamma+1}{\gamma-1}}}$$

The model employs an empirical discharge coefficient, Cd, to account for non-ideality (pipe friction, fittings, orifice shapes, etc). Several references provide formulae for predicting Cd using resistance coefficients (K), which are tabulated for a variety of fittings and pipe lengths. A perfect orifice has a value for Cd of 1.0, although in practice this is unattainable. For ideal well-rounded orifices Cd may be 0.97, whilst for sharp-edged orifices it is a function of Reynolds Number varying typically from 0.6 to 0.95. A good average value to employ for discharge calculations from vessels or short lengths of pipe (L/D <3) is Cd = 0.8. (Pitblado, 1996).

B2. Jet Fires

A number of correlations and techniques exist to estimate the dimensions and heat radiation intensities of jet fires. However, these methods have been developed specifically for estimating the characteristics of process plant flares. The experimental data used in developing and/or validating these models are derived from trials involving predominantly methane (natural gas) under flow conditions similar to those experienced in flare systems. Therefore, it is acknowledged that there is some uncertainty involved in applying these methods to conditions significantly different to those existing in the experimental basis.

The two most common models used for jet fire consequence analysis are those developed by Chamberlain (1987) and Cook (1987 a, b, c). The Chamberlain model has been based on the results of 31 large-scale methane flare trials and 98 laboratory scale trials. This model consists of a large set of complex correlations accounting for a range of factors including wind speed and direction and flame tilt and orientation.

The Cook model has been based on the results of 157 large-scale natural gas flare trials. This model is much simpler than that of Chamberlain and ignores the effects of wind speed and direction and flame orientation.

The flame lengths observed by Cook in his field scale experiments correlate reasonably well with the total heat release rate of the flares, despite the fact that such a correlation admits no effect of the wind on the flame length. Based on the Cook results the following correlation for flame lengths has been developed:

$$L_f = 0.062Q^{0.467}$$

Where: L_f = curvilinear flame length (m)

Q = energy release rate (kW)

The Cook correlation has been used to estimate the jet fire lengths for is project. Based on the data provided by Chamberlain (1987), the mean flame width is taken to be:

$$W_f = 0.2\,L_f$$

Where: Wf = mean flame width (m)

L_f = curvilinear flame length (m)

Estimation of the effects of jet fires on personnel and structures requires information on the magnitude of heat radiation intensities emitted from the flame surface (known as the flame surface emissive power, or SEP). It is the SEP (and not the convective heat) emitted from the flame that is the main radiative hazard to personnel and structures. Considerable research involving large-scale field trials have measured the SEP of jet fires to be in the order of 200 -

250 kW/m² (Bennett, 1990 and Cowley, 1990). All these trials have been based on methane and other light hydrocarbon jet fires.

The results of the trials conducted by Chamberlain (1987) and Cook (1987 a, b) have been used to estimate the SEP of the jet fires in this plant. These authors have measured the fraction of total heat radiated from the flame. The American Petroleum Institute (1992) also provides information on flame SEPs for flares. Based on these data it is assumed that the fraction of heat radiated from the flame is 0.2. Therefore, the flame SEP can be calculated from the following:

$$SEP = \frac{0.2xQ}{Flame\_Area}$$

SEP = flame surface emissive power (kW/m^2)

Q = total energy release rate (kW)

Flame area is derived from calculated flame length and width.

Exposure to radiation intensities from a large fire may result in either severe burns or fatality. The effect is a function of both the intensity of the radiation and the duration of exposure.

The model employed in estimating thermal radiation effects from jet flames is called the point source model. When the heat flux at the flame surface has been calculated, the heat incident on nearby objects is determined by assuming that all the heat is radiated from a small point at the centre of the flame, halfway along its length (as Cook (1987) reports that the intensity of the flame is noticeably greater at this point). The equation is:

$$q = \eta\, Q/4\pi R^2$$

where: q = incident radiation intensity at receiver (kW/m^2)

η = emissivity factor (0.2)

The quantification of the fatality potential (in terms of fatality probability and number of fatalities) for personnel exposed to jet fires is complex and difficult. The fatality potential is a function of a range of factors including:

- The location of personnel relative to the jet fire and the exposure levels of heat radiation;
- The availability of shielding provided by equipment and structures; and
- The ability of personnel to escape or seek refuge.

To ensure a conservative approach was adopted, it was assumed that all jet flames were horizontal and shielding effects were ignored.

B3. Explosions

The ignition of a flammable gas cloud will always produce a flash fire which is characterised by a rapid combustion process through the cloud producing an expanding volume of fire. If the combustion rate is very rapid, explosion overpressures can be produced with the potential to cause harm or damage outside of the final combustion volume. Hence, the ignition of a flammable gas cloud will result in either of the following outcomes:

(i) An explosion with an accompanying flash fire; and
(ii) A flash fire only.

Considerable advances have been made in recent years in the understanding of the fundamentals and mechanisms of gas explosions. The factors that influence the severity of explosions have been identified and have undergone significant theoretical and experimental research. These factors include fuel reactivity, congestion, confinement, gas cloud size and ignition source. However, there is no widely accepted method for quantitatively modelling explosions.

A number of models have been developed in recent years for modelling vapour cloud explosions. These correlations account for cloud size, congestion and reactivity. However, the results obtained by all these models are very dependent on the subjectivity of assumptions and judgements required in their methodology.

The TNT equivalence method has been widely used for gas explosions. The TNT equivalence method applies pressure-distance curves for TNT explosions to gas explosions and the equivalent TNT charge is estimated from energy content in the exploding gas cloud. In the original TNT equivalence method, the geometric conditions are not taken into account. Bjerketvedt et al (1997) state that the results from this type of analysis have hardly any relevance and should not in general be used.

A modified TNT equivalence method that takes geometrical effects into account has been proposed which has fairly good agreement between the predicted values and the experimental values, as long as the explosion pressure in the cloud is in a few atmospheres range. However, for explosion pressures below one atmosphere, this method overestimates the blast (Bjerketvedt *et. al.,* 1997).

Puttock (1995) outlines a model that has been based on both small and large scale experiments and trials. The model does not apply in areas which are more than 60% enclosed by a combination of walls or other obstructions.

The Puttock model relates the explosion overpressure P at any distance r from the edge of the vapour cloud by the following formula:

$$P = [R_0 / (r + R_0)] P_0$$

Where:

P = explosion overpressure at distance r from the cloud (kPa)

R_0 = radius of the vapour cloud (m)

r = distance to point of interest (m)

P_0 = overpressure at the cloud (also known as the source pressure) (kPa)

The source pressure P_0 is a product of a reference pressure P_{ref} and a fuel factor F. The reference pressure P_{ref} is the estimated maximum pressure that could be generated by ignition of a propane release in the same layout under consideration. A decision tree is provided in Puttock (1995) to assign a value to P_{ref}. The decision tree is difficult to use, but the following values are indicative:

P_{ref} = 0.1 kPa no congestion or confinement

P_{ref} = 0.2 kPa congestion present but no area exists where gas has to pass four or more obstacles to escape to open air

P_{ref} = 0.5 significant congestion requiring more than five obstacles to be passed for vapour to escape

P_{ref}= 1 major congestion

The influence of the reactivity of the fuel is evaluated by determining the fuel factor F for the fuel. Puttock provides values of F for a range of fuels, including methane (F = 0.6), propane & cyclohexane (F = 1.0) and ethylene (F = 3.0), which has the highest F value (Puttock 1995).

Low reactivity materials such as methane are not considered to be credible explosion scenarios when the physical environment is relatively uncongested and totally unconfined, reducing the potential of significant gas accumulation and high flame speeds. Lees (1996) provides the following conclusions:

"There is now considerable evidence that vapour clouds of methane at normal temperatures burn, but do not readily explode. Many experiments have been done in which attempts have been made to initiate explosions in methane clouds, but in which no explosion occurred. The occurrence of vapour cloud explosions involving methane has been reviewed by V C Marshall (1987). He cites expert opinion to the effect that there has been no case of an unconfined vapour cloud explosion with natural gas".

A value of 0.5 for Pressure reference, equivalent to relative congestion, was used for modelling purposes

The radius of the cloud is estimated from the volume, the volume of the cloud has been estimated from the mass in the cloud above the LFL. This has been estimated using a method from TNO (1992) where :

$M_{ex} = K * M_s * D_{LFL} / U_w$

where;

M_{ex} = amount of gas in explosive region (kg)

M_s = source strength (kg/s)

D_{LFL} = Distance to LFL

U_w = Windspeed (m/s)

K = constant based on stability class:

A stability = 0.638,	B stability = 0.632,	C stability = 0.629
D stability = 0.625,	E stability = 0.620,	F stability = 0.611

B4. Dispersion Modelling (for Toxic Gas and Flash Fires)

The dispersion of hazardous and pollutant materials in the atmosphere has been the subject of intense interest for some decades. This interest has resulted in the development of many different models for dispersion. The first models were developed in order to study the behaviour of pollutants discharged from vents and stacks. These pollutants usually form neutral plumes; i.e., plumes with densities similar to that of air; therefore the first models concentrated on neutral dispersion.

Hazardous releases involve atypical events, so modelling characteristics will generally be different to those used in pollution studies. In a hazard analysis the clouds which are denser than air are usually of most importance; clouds which are lighter than air will float upwards and are therefore more likely to disperse harmlessly.

Hazardous releases involve atypical events, so modelling characteristics will generally be different to those used in pollution studies. Hanna and Davies (1986) compared the characteristics of emissions of Routine versus Accidental Releases as:

Routine Release	**Accidental Release**
• Source well-defined (physical) and chemical parameters easily measured	• Source poorly defined (physical and chemical parameters not all measured)
• All gas release	• Gas and/or liquid release
• Continuous, steady-state	• Usually highly transient
• Little thermodynamic interaction with ground/water surface	• Often strong thermodynamic interaction with surface
• No phase changes	• Frequent phase changes
• Generally one hour averaging time	• Variety of averaging times, from one second to one hour or longer
• No dense gas effects	• Possible dense gas slumping, terrain-following

The rate at which gas concentration may be reduced depends greatly upon a property of the atmosphere known as its 'stability'. Pasquill defined a range of atmospheric stabilities, from A (hot sunny day with a high degree instability and vertical mixing) through to F (inversion, with stable air layering and little vertical mixing other than due to turbulence caused by the air movement with any breeze), and then developed graphs showing the extent of horizontal and vertical dispersion for different atmospheric stabilities. The Pasquill stability categories are typically characterised as follows in Table B3.

Table B.3 – Pasquill stability categories		
A	Unstable	Daytime – sunny, light winds
B	Unstable	Daytime - moderately sunny, light to moderate winds
C	Unstable	Daytime - moderate winds, overcast or windy and sunny
D	Neutral	Daytime - wind, overcast Nighttime – windy
E	Stable	Nighttime - moderate winds with little cloud or light winds with more clouds
F/G	Stable	Night time - light winds, little cloud

Factors affecting dispersion include:

- wind speed;
- roughness of the ground (measured by surface roughness), wakes caused by buildings or obstacles in the path of the dispersing gas or sloping of ground;
- momentum of gas released vertically of the escape point;
- air entrainment in the vicinity of the escape point;
- releases from the tops of buildings;
- density and buoyancy effects;
- atmospheric chemistry;
- variations in wind speed, wind direction and turbulence;
- treatment of the ground and the base of any elevated inversion as flat, perfectly reflecting surfaces; and
- averaging time.

Dispersion modelling was performed by using the SLAB (Ermak, 1990) computer model. SLAB can treat continuous, finite duration and instantaneous releases from four types of sources: a ground level evaporating pool, an elevated horizontal jet, a stack or elevated vertical jet and a ground-based instantaneous release. While the model is designed to treat denser-than-air releases, it will also simulate cloud dispersion of neutrally buoyant releases and includes lofting of the cloud if it becomes lighter-than-air. SLAB predictions have performed quite well when compared with a wide range of data obtained from both laboratory-scale and field scale heavy gas dispersion experiments.

For dispersion, the lowest wind speed used was 1m/s. This level is as recommended by Lindes *et al* (1997), as using dispersion models with very low wind speeds can lead to erroneously high concentration predictions. However, Lindes *et al* report that “for calm meteorological conditions, the speed of the upwind diffusion can exceed the windspeed, so that a well-defined plume may not actually form”. Further, “strong winds have a greater tendency to maintain their direction than light winds”.

As noted by CCPS (1996), if a dense gas release near the ground is instantaneous or of short duration (i.e., the time of release is less than the time of travel to the receptor), the worst-case conditions are likely to be associated with moderate wind speeds. Short-time releases of dense gases tend to spread out widely near the source when winds are light, thus delaying their advection downwind. Because the winds are light, by the time the cloud reaches a receptor at a distance of 500 m or 1000 m, it is relatively dilute. In contrast, when winds are higher, the dense cloud does not spread out widely near this source but is advanced downwind with less dilution.

Therefore, assessment of conservative ‘model’ results should be tempered by qualitative assessment of parameters that may lessen impacts to more realistic values.

Where release characteristics for postulated hazard scenarios were similar, these similar scenarios were grouped together (clustered) such that each cluster was represented by one Equivalent Discrete Failure (EDF) whose release rate was higher than each of the events in the cluster.

B.5 Toxic Impact Criteria

In general a ten minute exposure represents a period of exposure below which the effects relationship (eg probit) have been derived more by numerical extrapolation rather than actual animal experimental data due to the lack of existing experimental data for smaller exposure periods (Franklin 1992). Additionally, Withers and Lee (1985) state that the chlorine probit developed is strictly only applicable for the exposure range of ten to thirty minutes.

Limitations of exposure time are especially relevant to instantaneous releases. The distances thus estimated are the <u>maximum</u> for which the selected concentration occurs. Therefore, at these particular concentrations, the observer is exposed to this concentration for a very short time (the observer will for other times be exposed to concentrations below this value). It would seem that the most appropriate means may involve integration of a time of exposure/toxic concentration profile; however, limitations with animal experimental data (which generally involves a constant toxic gas exposure for exposure periods normally greater than 10 minutes) means that such an approach is not likely to be valid (Franklin 1992).

Therefore, an exposure time of ten minutes is used, even though for larger, transient releases (and obviously for release involving catastrophic), there would probably be insufficient inventory to provide a release time of ten minutes. Such an approach is considered conservative.

For calculation of individual fatality risk, the effect distances associated with toxic releases have a “greater spread” than those associated with fire scenarios. Therefore, the use of a single effect level (normally taken as 50% possibility of fatality) may not provide a true reflection of the risks ‘far-field’ from the hazardous source. For fatality risk, the different probabilities of fatality for various toxic chemical loads are reflected by the probit relationship. The probit relationship is represented by a range of discrete levels (10%, 30%, 50%, 90%) such that they are suitable (i.e. are cumulatively added) for input into the risk assessment model.

B.6 Source Term Parameter Considerations

In terms of land use safety planning, modelling of dry-seal gas holder failures using generic methods that might apply to pressure or atmospheric vessels may not be directly appropriate, given their unique design. A gas holder operates at low pressure (taken not to exceed 6kPa), and consists of two major components; the lower steel tank shell and the upper rubber seal.

The only known previous published risk assessment of large gas holders is that by Bernatik & Libisova (2004). This article summarises the results of quantitative risk assessment for the operation of six large gas holders in the Czech Republic. Unfortunately, none of the gas holders analysed were of the dry-seal type, nor is there sufficient information provided to fully recreate all the source term modelling used. The source term parameters for the worse case scenario would seem to be rupture of the gas holder inlet pipeline. It is noted that the paper states that "*in the worst case of gas holder accidents, the area of exposure to a cloud of toxic gases of the lethal concentration was determined to extend for about 150m from the gas holder*".

Many hazard scenarios are appropriately approximated by source term models which calculate the initial instantaneous flow rate based on the pressure and temperature existing in the source system or vessel when the release first occurs. In reality, the initial instantaneous flow rate from a leak in a pressurised gas system or vessel is much higher than the average flow rate during the overall release period, because the pressure and flow rate decrease with time as the system or vessel empties.

For this analysis, LDG, BFG and COG pipe release incidents involve relatively low pressure sources (ie. the flow will not reach sonic velocities) from failures associated with extremely large pipe diameters (eg up to 2.8m). This scenario therefore exhibits an unusual characteristic in that as the pressure is not significantly greater than ambient, the drive to release material will only continue until the pressure in the pipe equals the ambient pressure.

For example, consider a section of line between the BOS and the cogeneration plant. If this line is taken to be a closed system of 1000m of 2.8m diameter pipe at 1.15 bar and 20C, then assuming ideal gas conditions (and constant temperature as an open system), a catastrophic (100% diameter) failure of this line reaches atmospheric pressure (ie no flow conditions) within a few seconds. So, for the larger failure scenarios, the modelling is better approximated by a short term release of an average release rate rather than as a continuous release based on the initial rate.

However, the proposed design does involve a gas holder with a floating roof to maintain pressure. But it would be expected that this pressure makeup would have a lag time to respond, and may also be limited by a maximum allowable roof speed. In addition, on detection of any leak, it would be assumed that steps would be taken to isolate the gas holder so that its inventory is confined, or release rate reduced.

Therefore, to consider these aspects in the one scenario, it was assumed that a catastrophic failure of the system results in the assumed closed pipe system depressurising within 20 second (which corresponds to an average release rate of 52 kg/s). It was then assumed that this release is repeated again and again to account for any additional flows and inventory supply from the gas holder. Therefore, a "pseudo" diameter was calculated based on the average release rate for the case where the system is essentially at atmospheric pressure after 20 seconds. This diameter (680mm) was then used as the constant release rate for the dispersion modelling, as well as the scaling parameter for which to determine the size of the catastrophic, big and small leaks (given that the generic sizes of 1%, 5% and 20% of the x-sectional area would have been based on data from pipes mainly up to 100mm, and not for pipes of 2.8m diameter, where such data was not available in the literature).

To make the analysis consistent (as many of the characteristics of the releases are similar), the failure scenarios for the short COG and BFG interworks pipes to the cogeneration plant were also given a pseudo-diameter of 680mm.

For the gas holder itself, this is considered to consist of two types of failures. The lower steel shell, and failure of the rubber seal. Given the reinforcement and design parameters associated with the rubber seal, it is not expected to tear in long lengths, but instead for a small rip. Therefore, it is assumed it would be equivalent of a big leak (220mm).

B.7 Probit Relationship

The relation between probit (as used in the probit equation) and probability is shown in the following table.

Table B3 Relation between probits and probabilities %

%	0	1	2	3	4	5	6	7	8	9
0	-	2.67	2.95	3.12	3.25	3.36	3.45	3.52	3.59	3.66
10	3.72	3.77	3.82	3.90	3.92	3.96	4.01	4.05	4.08	4.12
20	4.16	4.19	4.23	4.26	4.29	4.33	4.36	4.39	4.42	4.45
30	4.48	4.50	4.53	4.56	4.59	4.61	4.64	4.67	4.69	4.72
40	4.75	4.77	4.80	4.82	4.85	4.87	4.90	4.92	4.95	4.97
50	5.00	5.03	5.05	5.08	5.10	5.13	5.15	5.18	5.20	5.23
60	5.25	5.28	5.31	5.33	5.36	5.39	5.41	5.44	5.47	5.50
70	5.52	5.55	5.58	5.61	5.64	5.67	5.71	5.74	5.77	5.81
80	5.84	5.88	5.92	5.95	5.99	6.04	6.08	6.13	6.18	6.23
90	6.28	6.34	6.41	6.48	6.55	6.64	6.75	6.88	7.05	7.33

Appendix C
Consequence Results

APPENDIX C – CONSEQUENCE TABLES FOR TOXIC FIRE AND EXPLOSION

C1. Consequence Tables for Toxic Fire and Explosion

Table C1 – Summary of Consequence Analysis – Jet Fires

Release Case	Description/Location	Material	Process Conditions		Release Type	Hole Diameter (mm)	Release Rate use (kg/s)	Length (m)	Radius Distance (m) around jetfire (located centre flame)		
			Pressure (bar-abs)	Temp (C)					23kW/m²	12.5kW/m²	4.7kW/m²
1.1	GAS HOLDER	BOS	1.07	20	Rupture	1000.00	77.93	25.39	16.54	22.44	36.59
1.2			1.07	20	Catastrophic Leak	447.21	15.59	11.97	7.40	10.03	16.36
1.3			1.07	20	Big Leaks and Seal Failure	223.61	3.90	6.27	3.70	5.02	8.18
1.4			1.07	20	Small Leaks	100.00	0.78	2.96	1.65	2.24	3.66
2.1	LDG releases from pipe running from Gas holder to Cogeneration Plant	BOS	1.15	20	Rupture (100% x-sectional area)	680.00	52.60	21.13	13.59	18.43	30.06
2.2			1.15	20	Catastrophic Leak (20% x-sectional area)	304.11	10.52	9.96	6.08	8.24	13.44
2.3			1.15	20	Big Leaks (5% x-sectional area)	152.05	2.63	5.22	3.04	4.12	6.72
2.4			1.15	20	Small Leaks (1% x-sectional area)	68.00	0.53	2.46	1.36	1.84	3.01
3.1	LDG releases from pipe running from LDG Switching Station to Gas holder	BOS	1.06	35	Rupture (100% x-sectional area)	680.00	32.55	16.89	10.69	14.50	23.65
3.2			1.06	35	Catastrophic Leak (20% x-sectional area)	304.11	6.51	7.96	4.78	6.48	10.58
3.3			1.06	35	Big Leaks (5% x-sectional area)	152.05	1.63	4.17	2.39	3.24	5.29
3.4			1.06	35	Small Leaks (1% x-sectional area)	68.00	0.33	1.97	1.07	1.45	2.36
4.1	COG Release from pipe running to Cogeneration Plant	COG	1.06	35	Rupture (100% x-sectional area)	680.00	20.78	33.39	22.18	30.09	49.06
4.2			1.06	35	Catastrophic Leak (20% x-sectional area)	304.11	4.16	15.75	9.92	13.45	21.94
4.3			1.06	35	Big Leaks (5% x-sectional area)	152.05	1.04	8.24	4.96	6.73	10.97
4.4			1.06	35	Small Leaks (1% x-sectional area)	68.00	0.21	3.89	2.22	3.01	4.91
5.1	BFG Release from pipe running to Cogeneration Plant	BFG	1.06	35	Rupture	680.00	34.02	12.64	7.84	10.64	17.35
5.2			1.06	35	Catastrophic Leak (20% x-sectional area)	304.11	6.80	5.96	3.51	4.76	7.76
5.3			1.06	35	Big Leaks (5% x-sectional area)	152.05	1.70	3.12	1.75	2.38	3.88
5.4			1.06	35	Small Leaks (1% x-sectional area)	68.00	0.34	1.47	0.78	1.06	1.73
6.1	CH4 Release from pipe running to Cogeneration Plant	Methane	11	20	Rupture (100% x-sectional area)	300.00	106.75	85.59	60.77	82.44	134.44
6.2			11	20	Catastrophic Leak (20% x-sectional area)	134.16	21.35	40.37	27.18	36.87	60.12
6.3			11	20	Big Leaks (5% x-sectional area)	67.08	5.34	21.13	13.59	18.43	30.06
6.4			11	20	Small Leaks (1% x-sectional area)	30.00	1.07	9.96	6.08	8.24	13.44

Nb(1): Final design may have temperature of BOS off-gas between 20C and 70C, value of 20C used as this results in a greater mass release rate
Nb(2): Report of results to two decimal points is strictly for calculation purposes and does not imply the accuracy of the models used

Table C2.1 – Summary of Consequence Analysis – Flash Fires and Explosions

Release Case	Description/Location	Material	Flash Fires – distance to LFL%							Puttock Explosion Method			Volume (m3) calculations						
			C4	D2	D4	D7	D10	E2	F1	Pref	Fuel Factor	Max Source P (kPa)	C4	D2	D4	D7	D10	E2	F1
1.1	Gas holder	BOS	36.00	38.00	39.00	39.00	38.00	41.00	42.00	0.50	0.60	30.00	391.67	821.59	421.61	240.92	164.32	879.37	1775.47
1.2			17.00	12.00	16.00	15.00	15.00	16.00	16.00	0.50	0.60	30.00	36.99	51.89	34.59	18.53	12.97	68.63	135.27
1.3			6.00	7.00	6.00	5.00	5.00	7.00	7.00	0.50	0.60	30.00	3.26	7.57	3.24	1.54	1.08	7.51	14.80
1.4			3.00	3.00	3.00	3.00	3.00	3.00	2.00	0.50	0.60	30.00	0.33	0.65	0.32	0.19	0.13	0.64	0.85
2.1	LDG releases from pipe running from Gas holder to Cogeneration Plant	BOS	24.47	25.06	25.85	26.31	26.34	26.27	26.52	0.50	0.60	30.00	176.28	358.82	185.04	107.62	75.43	373.14	742.22
2.2			9.93	10.22	10.32	10.28	10.03	10.39	10.39	0.50	0.60	30.00	14.30	29.26	14.77	8.41	5.74	29.50	58.14
2.3			3.80	3.69	3.80	3.89	3.94	3.51	3.36	0.50	0.60	30.00	1.37	2.64	1.36	0.79	0.56	2.49	4.71
2.4			2.21	2.17	2.20	2.22	2.23	2.08	1.98	0.50	0.60	30.00	0.16	0.31	0.16	0.09	0.06	0.30	0.55
3.1	LDG releases from pipe running from LDG Switching Station to Gas holder	BOS	22.40	24.13	24.70	24.58	23.71	25.56	25.65	0.50	0.60	30.00	107.28	229.61	117.54	66.84	45.14	241.35	477.26
3.2			8.81	9.37	9.27	8.86	8.22	9.50	9.48	0.50	0.60	30.00	8.44	17.83	8.82	4.82	3.13	17.94	35.28
3.3			3.59	3.49	3.59	3.65	3.62	3.37	3.14	0.50	0.60	30.00	0.86	1.66	0.85	0.50	0.34	1.59	2.92
3.4			2.08	2.06	2.09	2.10	2.09	2.00	1.89	0.50	0.60	30.00	0.10	0.20	0.10	0.06	0.04	0.19	0.35
4.1	COG Release from pipe running to Cogeneration Plant	COG	65.00	94.00	81.00	69.00	69.00	116.00	152.00	0.50	3.00	150.00	487.34	1400.58	603.44	293.74	205.62	1714.55	4428.09
4.2			29.00	40.00	35.00	31.00	26.00	45.00	49.00	0.50	3.00	150.00	43.49	119.20	52.15	26.39	15.50	133.03	285.50
4.3			13.00	18.00	17.00	14.00	12.00	20.00	21.00	0.50	3.00	150.00	4.87	13.41	6.33	2.98	1.79	14.78	30.59
4.4			6.00	7.00	7.00	5.00	5.00	8.00	8.00	0.50	3.00	150.00	0.45	1.04	0.52	0.21	0.15	1.18	2.33
5.1	BFG Release from pipe running to Cogeneration Plant	BFG	31.00	35.00	35.00	34.00	32.00	36.00	38.00	0.50	1.00	50.00	142.08	318.78	159.39	88.48	58.29	325.26	676.70
5.2			13.00	14.00	14.00	13.00	12.00	14.00	14.00	0.50	1.00	50.00	11.92	25.50	12.75	6.77	4.37	25.30	49.86
5.3			5.00	5.00	5.00	5.00	5.00	5.00	5.00	0.50	1.00	50.00	1.15	2.28	1.14	0.65	0.46	2.26	4.45
5.4			3.00	3.00	3.00	3.00	2.00	2.00	2.00	0.50	1.00	50.00	0.14	0.27	0.14	0.08	0.04	0.18	0.36
6.1	CH4 Release from pipe running to Cogeneration Plant	Methane	122.00	136.00	130.00	135.00	125.00	146.00	147.00	0.50	0.60	30.00	2441.03	5407.70	2584.56	1533.70	994.06	5758.88	11428.31
6.2			53.00	61.00	62.00	56.00	51.00	63.00	61.00	0.50	0.60	30.00	212.09	485.10	246.53	127.24	81.12	497.00	948.47
6.3			24.00	28.00	27.00	25.00	23.00	30.00	30.00	0.50	0.60	30.00	24.01	55.67	26.84	14.20	9.15	59.17	116.62
6.4			9.00	11.00	10.00	9.00	7.00	11.00	11.00	0.50	0.60	30.00	1.80	4.37	1.99	1.02	0.56	4.34	8.55

Table C2.2 – Summary of Consequence Analysis – Flash Fires and Explosions

Table D1 continued

Release Case	Description/Location	Material	Distance (m) to following overpressure levels																				
			21 kPa							14 kPa							7 kPa						
			C4	D2	D4	D7	D10	E2	F1	C4	D2	D4	D7	D10	E2	F1	C4	D2	D4	D7	D10	E2	F1
1.1	GAS HOLDER	BOS	18	22	18	15	13	23	29	26	34	27	22	20	34	44	53	67	54	45	39	69	87
1.2			8	9	8	6	6	10	12	12	13	12	10	8	15	18	24	27	23	19	17	29	37
1.3			4	5	4	3	2	5	6	5	7	5	4	4	7	9	11	14	11	8	7	14	18
1.4			2	2	2	1	1	2	2	2	3	2	2	2	3	3	5	6	5	4	4	6	7
2.1	LDG releases from pipe running from Gas holder to Cogeneration Plant	BOS	13	17	14	11	10	17	22	20	26	20	17	15	26	33	40	51	41	34	30	52	65
2.2			6	7	6	5	4	7	9	9	11	9	7	6	11	14	17	22	18	15	13	22	28
2.3			3	3	3	2	2	3	4	4	5	4	3	3	5	6	8	10	8	7	6	10	12
2.4			1	2	1	1	1	2	2	2	2	2	2	1	2	3	4	5	4	3	3	5	6
3.1	LDG releases from pipe running from LDG Switching Station to Gas holder	BOS	11	15	12	10	9	15	19	17	22	18	15	13	22	28	34	44	35	29	26	45	56
3.2			5	6	5	4	4	6	8	7	9	7	6	5	9	12	15	19	15	12	11	19	24
3.3			2	3	2	2	2	3	3	3	4	3	3	3	4	5	7	9	7	6	5	8	10
3.4			1	1	1	1	1	1	2	2	2	2	1	1	2	3	3	4	3	3	2	4	5
4.1	COG Release from pipe running to Cogeneration Plant	COG	94	134	101	80	71	143	197	141	201	152	119	106	215	295	283	402	304	239	212	430	590
4.2			42	59	45	36	30	61	79	63	88	67	53	45	92	118	126	177	134	107	90	183	237
4.3			20	28	22	17	15	29	37	30	43	33	26	22	44	56	61	85	66	52	44	88	112
4.4			9	12	10	7	6	13	16	14	18	14	11	10	19	24	28	36	29	21	19	38	48
5.1	BFG Release from pipe running to Cogeneration Plant	BFG	21	27	22	18	15	27	35	31	41	32	27	23	41	53	62	82	65	53	46	82	105
5.2			9	12	9	8	7	12	15	14	18	14	11	10	18	22	27	35	28	23	20	35	44
5.3			4	5	4	3	3	5	7	6	8	6	5	5	8	10	13	16	13	10	9	16	20
5.4			2	3	2	2	1	2	3	3	4	3	3	2	3	4	6	8	6	5	4	7	8
6.1	CH4 Release from pipe running to Cogeneration Plant	Methane	32	42	33	28	24	43	54	48	63	49	41	36	64	81	97	126	99	83	72	129	162
6.2			14	19	15	12	10	19	24	21	28	23	18	16	28	35	43	56	45	36	31	57	71
6.3			7	9	7	6	5	9	12	10	14	11	9	8	14	18	21	27	22	17	15	28	35
6.4			3	4	3	2	2	4	5	4	6	5	4	3	6	7	9	12	9	7	6	12	15

Table C3.1 – Summary of Consequence Analysis – Toxic Dispersion Distances														
Release Case	50% fat – CO 5,550ppm							30% fat – CO 4,800ppm						
	Dist (m)							Dist (m)						
	C4	D2	D4	D7	D10	E2	F1	C4	D2	D4	D7	D10	E2	F1
1.1	250.00	411.00	339.00	276.00	229.00	537.00	910.00	269.00	446.00	375.00	300.00	248.00	598.00	1101.00
1.2	106.00	185.00	149.00	115.00	101.00	236.00	323.00	115.00	199.00	165.00	124.00	104.00	257.00	380.00
1.3	51.00	94.00	70.00	55.00	46.00	114.00	145.00	55.00	104.00	77.00	60.00	50.00	123.00	169.00
1.4	22.00	38.00	30.00	23.00	19.00	50.00	55.00	24.00	43.00	32.00	25.00	21.00	56.00	64.00
2.1	191.17	344.92	262.80	208.42	177.05	489.24	801.20	206.72	380.58	290.85	227.03	193.28	554.50	986.62
2.2	82.08	142.99	112.39	89.45	76.64	193.16	278.19	89.75	159.63	121.93	97.18	82.38	218.38	338.97
2.3	39.84	67.05	53.86	43.35	37.04	87.08	114.18	43.67	75.18	58.95	46.82	40.06	99.25	139.33
2.4	17.24	28.04	22.47	17.89	14.74	34.65	41.59	18.70	31.47	25.08	19.72	16.36	39.65	50.44
3.1	150.83	220.58	193.37	169.49	145.78	318.19	566.24	162.92	225.21	206.96	181.39	159.07	359.77	679.67
3.2	58.34	104.68	88.17	73.45	61.64	132.18	192.97	74.03	114.59	94.86	79.68	66.57	146.44	228.35
3.3	32.28	48.94	44.21	34.67	29.33	63.23	81.21	35.37	55.95	48.05	37.70	31.68	69.62	96.94
3.4	13.41	21.93	17.68	13.05	9.98	24.02	30.59	14.76	24.48	19.64	14.69	11.26	27.44	36.10
4.1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
4.2	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
4.3	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
4.4	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
5.1	95.00	140.00	129.00	101.00	87.00	167.00	175.00	105.00	151.00	143.00	112.00	95.00	189.00	208.00
5.2	39.00	63.00	52.00	43.00	37.00	73.00	66.00	44.00	72.00	58.00	47.00	40.00	81.00	89.00
5.3	19.00	28.00	24.00	20.00	17.00	32.00	34.00	21.00	32.00	30.00	22.00	19.00	38.00	39.00
5.4	6.00	10.00	7.00	6.00	6.00	11.00	12.00	7.00	12.00	9.00	7.00	6.00	13.00	15.00
6.1	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3
6.2	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3
6.3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3
6.4	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3

Table C3.2 – Summary of Consequence Analysis – Toxic Dispersion Distances														
Release Case	**10% fat – CO** 3,900ppm							**Injurious – CO** 1,200ppm						
	Dist (m)							**Dist (m)**						
	C4	**D2**	**D4**	**D7**	**D10**	**E2**	**F1**	**C4**	**D2**	**D4**	**D7**	**D10**	**E2**	**F1**
1.1	304.00	507.00	419.00	340.00	280.00	697.00	1439.00	572.00	991.00	821.00	682.00	548.00	1678.00	2868.00
1.2	129.00	218.00	187.00	141.00	117.00	297.00	486.00	262.00	447.00	375.00	276.00	227.00	632.00	1741.00
1.3	62.00	119.00	88.00	67.00	56.00	142.00	212.00	117.00	236.00	176.00	131.00	109.00	303.00	672.00
1.4	27.00	50.00	37.00	29.00	24.00	66.00	87.00	51.00	107.00	75.00	56.00	47.00	138.00	259.00
2.1	233.58	446.18	332.34	256.57	215.58	669.66	1349.54	443.29	981.13	679.47	514.59	428.27	1686.06	3963.33
2.2	100.57	185.90	140.94	110.55	93.81	263.50	458.96	191.91	402.00	286.44	214.52	180.00	646.14	1860.69
2.3	49.32	88.22	67.63	53.54	45.43	120.00	187.13	94.52	191.33	138.02	104.27	87.41	295.15	774.38
2.4	21.42	37.28	29.07	22.48	18.75	48.39	67.62	41.64	81.76	59.29	45.28	37.76	120.65	284.23
3.1	178.46	297.75	231.08	203.77	177.94	426.58	886.07	320.48	611.82	428.90	374.71	350.89	1073.68	3294.88
3.2	82.60	130.43	105.27	90.39	75.35	172.50	294.22	157.23	254.49	213.77	176.12	142.69	391.59	1065.33
3.3	39.67	67.08	55.12	42.91	35.99	81.15	124.83	76.04	127.95	110.22	82.97	68.54	176.51	414.78
3.4	16.90	28.39	22.65	17.19	13.65	35.60	46.64	32.59	58.15	46.96	35.44	29.27	74.22	160.06
4.1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
4.2	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
4.3	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
4.4	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1	Note #1
5.1	119.00	180.00	166.00	128.00	108.00	225.00	276.00	227.00	392.00	330.00	260.00	214.00	556.00	1277.00
5.2	50.00	85.00	68.00	54.00	46.00	93.00	108.00	99.00	178.00	146.00	110.00	91.00	233.00	433.00
5.3	24.00	39.00	32.00	25.00	22.00	48.00	45.00	48.00	95.00	69.00	53.00	44.00	119.00	190.00
5.4	9.00	15.00	12.00	8.00	7.00	18.00	19.00	21.00	40.00	29.00	22.00	18.00	51.00	74.00
6.1	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3
6.2	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3
6.3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3
6.4	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3

Table C3.3 – Summary of Consequence Analysis – Toxic Dispersion Distances

Release Case	Irritant – CO 350ppm Dist (m)													
	C4	D2	D4	D7	D10	E2	F1							
1.1	1125.00	2026.00	1680.00	1388.00	1113.00	4412.00	>10000							
1.2	456.00	824.00	826.00	548.00	449.00	1366.00	6711.00							
1.3	217.00	450.00	354.00	256.00	211.00	596.00	2053.00							
1.4	96.00	213.00	146.00	109.00	90.00	293.00	687.00							
2.1	853.47	2153.99	1423.82	1038.71	854.87	3744.67	7405.35							
2.2	363.24	833.68	574.12	431.14	357.36	1505.77	3258.27							
2.3	176.46	388.22	272.95	204.85	170.53	656.89	1759.78							
2.4	78.48	166.46	116.88	87.73	72.98	262.84	804.61							
3.1	629.43	1266.68	815.99	755.52	698.00	3566.06	9770.00							
3.2	297.43	500.64	431.22	351.71	285.80	893.19	3912.56							
3.3	142.80	241.18	216.44	164.35	134.64	374.91	1395.71							
3.4	62.26	121.78	93.85	68.34	56.39	145.69	443.40							
4.1	46.00	36.00	43.00	56.00	76.00	35.00	32.00							
4.2	46.00	36.00	43.00	56.00	76.00	35.00	32.00							
4.3	30.00	18.00	25.00	38.00	48.00	17.00	15.00							
4.4	21.00	14.00	20.00	23.00	20.00	12.00	10.00							
5.1	448.00	779.00	632.00	530.00	430.00	1324.00	4979.00							
5.2	190.00	351.00	299.00	215.00	178.00	498.00	1515.00							
5.3	93.00	187.00	140.00	103.00	85.00	234.00	584.00							
5.4	40.00	85.00	59.00	44.00	37.00	116.00	221.00							
6.1	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3							
6.2	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3							
6.3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3							
6.4	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3	Note#3							

Note #1: *COG releases are buoyant - - long distance impacts n/a*
Note #2: *LFL for COG are centreline plume (which rises), whilst toxic measured at 1m height*
Note #3: *Not toxic (asphyxiant only - - not land use safety planning issue)*
Note#4: *Report of results to two decimal points is strictly for calculation purposes and does not imply the accuracy of the models used.*

Appendix D
Failure Rate, Probability Data and Risk Estimation

D1. Gas holder Failures

In the one referenced paper found on gas holders (Bernatik & Libisova, 2004), it is quoted that the frequency of gas holder ruptures was 5 x 10^{-6} events/year, and was sourced from the TNO Purple Book (TNO, 1999). In consulting the TNO reference, this number would appear to be the generic number used for stationary atmospheric tanks (which would seem to have as the main data basis, tanks storing a liquid).

Therefore, as no reliability data or generic data appears to exist specifically for gas holders, an internal report by BHPE (1992) determined the likelihood of the existing gas holder releases by assigning frequencies to base events which have the potential to result in significant releases. These assigned frequencies were expressed in terms of orders of magnitude and were based on risk engineering judgement.

Using similar reasoning for the BOS Gas holder, the frequency estimation is shown in **Table D1**.

Table D1: Failure Frequency Estimate for BOS Gas holder

Base Event	Minor (100mm hole)	Significant (224mm Hole)	Major (447mm hole)	Catastrophic (1000mm hole)
Corrosion	10^{-3}/yr			
Structural Failure			10^{-5}/yr	
Overpressurisation / Vacuum				< 10^{-6}/yr
Aircraft Impact				< 10^{-6}/yr
Explosion/Missiles		10^{-4}/yr		
Earthquake				< 10^{-6}/yr
Total Frequency	10^{-3}/yr	10^{-4}/yr	10^{-5}/yr	10^{-6}/yr

The basis for the failure frequency show in **Table D1** is as follows:

- Corrosion could result in holes forming in the holder. For the existing COG holder, minor leaks have occurred around corroded rivets, and in the extreme case, a 50mm corrosion hole has occurred. Given that the LDG will provide a much less corrosive environment than COG, in addition to the holder containing a synthetic rubber lining, it would be expected to be at least two order of magnitude less likely event for a hole twice this size.

- It is expected that a number of protection systems will exist (similar to the existing gas holders) to prevent overpressurisation of the gas holder. These would include automatic activation of valving to vent excess LDG to flare; manual activation of valving by BlueScope Steel personnel who will continuously monitor the LDG system; automatic shut-off gas holder by inlet valve at high gas holder level. The estimated frequency of overpressurisation incidents is less than 10^{-6}/yr based on a failure probability of 0.01 for each protection system and a demand rate of 10/yr .

- Explosions from incidents outside the gas holders area may produce damaging overpressures and/or fragment missiles resulting in a breach of a gas holder plate in the form of a crack or hole. Molten slag/water explosions at slag dumps can produce missile hazards. Existing slag pits are located approximately well away from the gas holder. Significant releases of flammable gases in confined areas can result in explosions upon igniting. The proposed area around the gas holders is essentially open and free of congestion. A rail line passes approximately 250m from the gas holder. It is non-credible that a rolling stock derailment will impact on the holder due to the low locomotive speeds and the straight run. It is therefore considered that the frequency of gas holder releases from missiles/explosion incidents would not exceed 10^{-4}/yr.

- Structural failures of the gas holders can result from a range of factors including mechanical defects, corrosion, design errors, poor maintenance, poor construction standards and ground instability. Maintenance standards, corrosion and mechanical defects are also causes of failure. Given the expected high standards with this new plants installation and operation, it is estimated that the likelihood of structural failure of a gas holder would not exceed 10^{-5}/yr.

- The estimated frequency of a catastrophic failure of the gas holders due to earthquakes is in the order of 10^{-6}. This frequency is the product of a major earthquake frequency and the probability of major damage to the holder due to the earthquake. The Newcastle earthquake of 1989 has an intensity of Richter Magnitude of 5.5 (the largest recorded in NSW).

As for failure of the rubber seal, it is noted that an expected manufacturer[5] indicates that it is expected to last more than 200,000 strokes (ie a service life of over 10years). A generic seal failure number is given by NPRD(1995) as 4.79×10^{-2} /yr, however, given the reinforcement provided to the seal by the polymide layer reinforcing material, a significant failure (say equivalent to a big leak - 220mm) of a well maintained and serviced gas holder seal is taken to be two order of magnitude less (ie: 4.79×10^{-4} / yr, or in terms of order of magnitude, 10^{3}/yr).

Pipe Failures

Considerable data is available in published literature on pipe failure rates, although very little seems to be available for the large pipe diameters that apply in this study. Estimates vary considerable as demonstrated by a review by Hawksley (Lees, 1996). Generic pipe failure rates which have been used in a number of Australian and overseas QRAs (eg DNV Technica, 1993, ICI Mond Databook) have been used in this study and are presented in **Table D2**.

[5] *http://www.contitech.de/ct/contitech/themen/produkte/membranen/gasspeichermembranen/gasspeicher_e.html*

Table D2 - Generic Pipe Failure Data		
Failure / Probability Type	**Numerical Value use in this Study**	**Basis / Reference**
Rupture (100% of cross sectional area)	2.2×10^{-8} L/D per year	DNV Technica (1993)
Catastrophic Leaks (20% of cross sectional area)	5.0×10^{-8} L/D per year	(nb: The above data has been generalised to include the reliability of butt welded joints and flanged joints)
Big Leaks (5% of cross sectional area)	1.2×10^{-7} L/D per year	L = pipe length D = pipe diameter
Small Leaks (1% of cross sectional area)	2.8×10^{-7} L/D per year	

Ignition Probabilities

Ignition of a gas release may occur either at the point of the release or at some distance from it. Ignition can be due to the presence of a range of sources that can provide sufficient energy to ignite the vapour. These can include:

- Engines and exhausts
- Electrical equipment
- Hot surfaces
- Human activity engaged in hot work such as welding, cutting and grinding
- Sparks
- Static electricity
- Hot particles

A gas release may also self-ignite due to the generation of static electricity or frictional energy at the high pressure leak site or because the gas release is above its auto-ignition temperature.

Hydrogen has a very low minimum ignition energy and it would be expected that gas mixtures containing high concentrations of hydrogen would have similarly low ignition energies.

Reasonable estimates of ignition probabilities for flammable releases are difficult to obtain. The probability is very dependent on plant specific factors such as process conditions, plant equipment and configuration, and standards of safety management amongst others. Many studies have been undertaken to derive generic ignition probabilities and guidelines.

The results of a review of these studies undertaken by Cox et. al. (1991) has been used to estimate ignition probabilities. This correlation is based predominantly on hydrocarbon leaks below their auto-ignition temperature. It has then been assumed that for 50% of the time, the ignition will explode, otherwise the ignition will result in a flash fire, but both end in a jet fire. This approach tends to predict that ignition likelihood is less than the probabilities used in the Bertatik & Libisova (2004) paper. Given that the land use concerns are due to toxic gas releases, this is a more conservative approach.

The probability that protection systems will fail to rapidly control and contain the hazard is taken as 1.0 (i.e. there is a 100% chance that, is an event occurred, the protection systems in place would fail to adequately perform). The protection system has not been defined specifically at this stage, but it could include a shut-off valve, operation response, etc.

Weather Data
The probabilities for the various wind speed / stability / direction combinations was sourced from BHPE (1992) and is summarised in **Table D3**.

Risk Estimations
To calculate the risk results, the risk program developed by the Warren Centre for Advanced Engineering, Sydney University was used. Details regarding the program are explained by Pearce (1986). This program performs a risk summation for a large number of individual points on a grid pattern around a facility. Individual risk contours can then be drawn connecting points all locations of equal risk. The continuous changing risk slope due to the probit was transformed into discrete levels (10%, 30%, 50%) such that they were suitable for input into the package.

Table D3: Probability of Pasquill Windspeed/Stability Categories for 16 wind/directions (%)																
Pasquill Windspeed Stability Category	**N**	**NNE**	**NE**	**ENE**	**E**	**ESE**	**SE**	**SSE**	**S**	**SSW**	**SW**	**WSW**	**W**	**WNW**	**NW**	**NNW**
C 4 m/s	0.5298	1.5771	1.8834	0.9768	0.7651	0.8592	1.2477	1.4245	1.1770	1.0476	1.3066	1.2592	0.5532	0.3534	0.1883	0.2590
D 2 m/s	0.1412	0.2354	0.0824	0.0706	0.1177	0.0706	0.0706	0.0824	0.0942	0.2236	0.5179	0.4120	0.1412	0.0942	0.1530	0.1412
D 4 m/s	1.1300	1.8833	0.8357	0.4591	0.3178	0.4402	0.8003	1.2594	1.5536	2.7660	2.2245	1.7655	1.1418	0.4002	0.6827	1.1299
D 7 m/s	0.9769	2.8719	1.5419	0.2825	0.1530	0.2001	0.7062	2.4600	3.1662	2.2599	0.6944	2.9788	2.2128	0.4119	0.3414	0.8593
D 10 m/s	0.1059	2.4953	0.9534	0.1059	0.1648	0.1766	0.4002	1.2359	1.9815	1.0711	0.2237	1.6479	2.4364	0.1648	0.0235	0.0471
E 2 m/s	0.9299	0.7651	0.2118	0.1176	0.0588	0.1884	0.3296	0.3413	0.6356	1.3771	2.6131	3.5311	1.3418	0.4002	0.6238	1.0123
F 1 m/s	0.6945	0.5179	0.1531	0.1530	0.1295	0.1765	0.1883	0.2002	0.3532	0.6945	2.2481	3.1661	1.6125	1.0240	0.9652	0.6710

Appendix E
Frequency Analysis Summary

Table E1 – Event Frequency and Location Information

Failure Case	Unit/Pipe Failure Descriptor	Location		Scenario Description	Equivalent Consequence Modelling Scenario (EMS)	Vessel / Pipe (V or P)	Pipe Dimensions		Vessel/Pipe Frequency Base Failure Rate (/yr)	Probability Immediate or Delayed Ignition	Protection Devices Failure Prob.	Frequency Jet Fire	Frequency Flash Fire	Frequency Explosion	Frequency Toxic
		X	Y				Dia (mm)	Length (m)							
1.1	Gas holder Failure - Lower Steel Structure	313	1012	Rupture	EFS#1.1	V	1000.00	n/a	1.00E-06	0.26	1.00	2.56E-07	1.28E-07	1.28E-07	7.44E-07
1.2	Gas holder Failure - Lower Steel Structure	313	1012	Catastrophic Leak	EFS#1.2	V	447.21	n/a	1.00E-05	0.09	1.00	9.26E-07	4.63E-07	4.63E-07	9.07E-06
1.3	Gas holder Failure - Rubber Seal and Lower Steel Structure	313	1012	Big Leak	EFS#1.3	V	223.61	n/a	1.10E-03	0.04	1.00	4.24E-05	2.12E-05	2.12E-05	1.06E-03
1.4	Gas holder Failure - Lower Steel Structure	313	1012	Small Leak	EFS#1.4	V	100.00	n/a	1.00E-03	0.01	1.00	1.39E-05	6.97E-06	6.97E-06	9.86E-04
2.1	LDG Line - GH to CG	840	621	Rupture	EFS#2.1	P	680.00	1374	4.45E-05	0.20	1.00	8.88E-06	4.44E-06	4.44E-06	3.56E-05
2.2	LDG Line - GH to CG	840	621	Catastrophic Leak	EFS#2.2	P	304.11	1374	2.26E-04	0.07	1.00	1.63E-05	8.16E-06	8.16E-06	2.10E-04
2.3	LDG Line - GH to CG	840	621	Big Leak	EFS#2.3	P	152.05	1374	1.08E-03	0.03	1.00	3.26E-05	1.63E-05	1.63E-05	1.05E-03
2.4	LDG Line - GH to CG	840	621	Small Leak	EFS#2.4	P	68.00	1374	5.66E-03	0.01	1.00	6.16E-05	3.08E-05	3.08E-05	5.60E-03
3.1	LDG Line - OG to GH	440	858	Rupture	EFS#3.1	P	680.00	469	1.52E-05	0.15	1.00	2.24E-06	1.12E-06	1.12E-06	1.29E-05
3.2	LDG Line - OG to GH	440	858	Catastrophic Leak	EFS#3.2	P	304.11	469	7.71E-05	0.05	1.00	4.11E-06	2.06E-06	2.06E-06	7.30E-05
3.3	LDG Line - OG to GH	440	858	Big Leak	EFS#3.3	P	152.05	469	3.70E-04	0.02	1.00	8.22E-06	4.11E-06	4.11E-06	3.62E-04
3.4	LDG Line - OG to GH	440	858	Small Leak	EFS#3.4	P	68.00	469	1.93E-03	0.01	1.00	1.55E-05	7.76E-06	7.76E-06	1.92E-03
4.1	COG Line	1376	376	Rupture	EFS#4.1	P	680.00	100	3.24E-06	0.11	1.00	3.60E-07	1.80E-07	1.80E-07	2.88E-06
4.2	COG Line	1376	376	Catastrophic Leak	EFS#4.2	P	304.11	100	1.64E-05	0.04	1.00	6.61E-07	3.30E-07	3.30E-07	1.58E-05
4.3	COG Line	1376	376	Big Leak	EFS#4.3	P	152.05	100	7.89E-05	0.02	1.00	1.32E-06	6.60E-07	6.60E-07	7.76E-05
4.4	COG Line	1376	376	Small Leak	EFS#4.4	P	68.00	100	4.12E-04	0.01	1.00	2.49E-06	1.25E-06	1.25E-06	4.09E-04
5.1	BFG Line	1376	376	Rupture	EFS#5.1	P	680.00	100	3.24E-06	0.15	1.00	4.91E-07	2.45E-07	2.45E-07	2.74E-06
5.2	BFG Line	1376	376	Catastrophic Leak	EFS#5.2	P	304.11	100	1.64E-05	0.05	1.00	9.02E-07	4.51E-07	4.51E-07	1.55E-05
5.3	BFG Line	1376	376	Big Leak	EFS#5.3	P	152.05	100	7.89E-05	0.02	1.00	1.80E-06	9.01E-07	9.01E-07	7.71E-05
5.4	BFG Line	1376	376	Small Leak	EFS#5.4	P	68.00	100	4.12E-04	0.01	1.00	3.40E-06	1.70E-06	1.70E-06	4.08E-04
6.1	CH4 Line	1376	376	Rupture	EFS#6.1	P	300.00	200	1.47E-05	0.31	1.00	4.58E-06	2.29E-06	2.29E-06	1.01E-05
6.2	CH4 Line	1376	376	Catastrophic Leak	EFS#6.2	P	134.16	200	7.45E-05	0.11	1.00	8.42E-06	4.21E-06	4.21E-06	6.61E-05
6.3	CH4 Line	1376	376	Big Leak	EFS#6.3	P	67.08	200	3.58E-04	0.05	1.00	1.68E-05	8.42E-06	8.42E-06	3.41E-04
6.4	CH4 Line	1376	376	Small Leak	EFS#6.4	P	30.00	200	1.87E-03	0.02	1.00	3.18E-05	1.59E-05	1.59E-05	1.83E-03