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Analysis of LNG Import- Terminal Release-Prevention Systems

E.G. Baker

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April 1982

Prepared for the U.S. Department of Energy
under Contract DE-AC06-76RLO 1830

Pacific Northwest Laboratory
Operated for the U.S. Department of Energy
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**ANALYSIS OF LNG IMPORT-TERMINAL
RELEASE-PREVENTION SYSTEMS**

E. G. Baker

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Pacific Northwest Laboratory
Richland, Washington 99352

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FOREWORD

This report is one of a series prepared by Pacific Northwest Laboratory (PNL) to communicate results of the Liquefied Gaseous Fuels (LGF) Safety Studies Project, being performed for the U.S. Department of Energy, Office of Environmental Protection, Safety and Emergency Preparedness (DOE/EP). The DOE/EP Office of Operational Safety, Environmental Safety Engineering Division (ESED), is conducting the DOE Liquefied Gaseous Fuels Safety and Environmental Control Assessment Program. The LGF Safety Studies project contributes research, technical surveillance and program development information in support of the ESED Assessment Program. This analysis of LNG import terminal release prevention systems benefited from the technical direction and guidance provided by Dr. Henry F. Walter and Dr. John M. Cece of ESED.

Completed effort in other tasks of the PNL project are reported in:

1. Assessment of Research and Development (R&D) Needs in Ammonia Safety and Environmental Control (PNL-4006)
2. An Overview Study of LNG Release Prevention and Control Systems (PNL-4014)

Work in progress includes more detailed studies of topics identified in the LNG facility overview study as being worthy of further investigation.

Other reports of this series are in preparation on the following subjects:

- Assessment of R&D needs in LPG Safety and Environmental Control
- Peakshaving Plant Release Prevention Analysis
- Storage Tank Analysis
- Fire Prevention and Control Assessment
- Human Factors in LNG Operations.

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The author also gratefully acknowledges the assistance provided by individuals at Pacific Northwest Laboratory (PNL) during the course of this study and preparation of the final report. J. G. DeSteeze, Manager of the Liquefied Gaseous Fuels Safety Studies Project at PNL, provided support and guidance throughout the study. P. J. Pelto provided technical insights about LNG facilities and also provided a technical review of the entire study. The editorial review of the final report prior to publication was contributed by Judy Danko and Cary Counts. Finally, Marlene Hale and Marilyn Linse contributed their word processing skills in the preparation of the final manuscript.

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1.0 SUMMARY

The purpose of this study is to provide an analysis of release prevention systems at LNG import terminals. An overview study of LNG import terminals (Pelto 1982) which preceded this effort identified 11 release scenarios which are typical of the potential hazards involved in the operation of LNG import terminals. These scenarios formed the basis for this more detailed study.

Failure modes and effects analysis and fault tree analysis were used to estimate the expected frequency of each release scenario for a reference import terminal. In addition, the effectiveness of release prevention, release detection, and release control systems were evaluated.

The reference LNG import terminal consists of three basic process areas: marine terminal and ship unloading, storage, and sendout/vaporization. Major failures in the marine terminal and ship unloading area are estimated to occur about 10^{-3} times per year and releases of 10^5 to 10^6 gallons of LNG are possible. Between 10^{-4} and 10^{-6} major failures per year are estimated to occur in the storage area and releases of over 10^7 gallons are possible (the entire contents of one tank). About 10^{-2} to 10^{-3} major failure per year should be expected from the sendout system (sendout pumps and high-pressure pipeline), with the maximum release size being 10^4 to 10^5 gallons. For the vaporizer section, we estimate that major failures will occur about 10^{-1} per year. However, the maximum release size in the vaporizer section is on the order of 10^3 gallons, significantly smaller than marine terminal and storage sections.

LNG release can be caused by failure of both passive and active components. For passive components a single failure is often enough to cause a release (e.g., a pipe rupture) however the probability of these failures is low. The failure rate of active components (e.g., control system failure) is higher, but in most cases the failure of at least two components is necessary before a release occurs. In addition, the reference LNG import terminal has emergency shutdown (ESD) systems which, when activated, can shut down and isolate the facility quickly. This will significantly reduce the size of the release compared to the possible

release sizes referred to previously. The probability of a release not being stopped and isolated by the ESD is 10^{-1} to 10^{-4} per demand, depending on the particular system.

Two scenarios were selected as being most significant with respect to safety at an LNG import terminal: 1) rupture of the main transfer line and 2) gross failure of a storage tank. These two scenarios have the largest maximum spill sizes and are considered to be the only two that present a significant potential hazard to the public. Design alternatives that could reduce the expected number of occurrences or reduce the size of the releases were identified for both scenarios. The effectiveness of these alternative systems was evaluated qualitatively and, where possible, quantitatively. Several of these design alternatives have the potential to significantly reduce the probability of a large release of LNG occurring at an import terminal. A more detailed technical and economic evaluation of these alternatives is necessary before the cost and benefits of the various alternatives can be compared. Based on the preliminary analysis undertaken in this study, many of these alternatives probably are not justified for remotely located facilities. However, for facilities located in highly populated areas, these design alternatives deserve serious consideration.

2.0 INTRODUCTION

The LNG industry employs a variety of release prevention and control mechanisms which contain LNG during transfer and storage and which detect and control an LNG release if it occurs.

The LNG Release Prevention and Control Task in the LNG Safety Studies Project has the basic objective of developing an adequate understanding of LNG release prevention and control systems and the factors which may nullify their usefulness. Some more specific objectives include:

- Identifying the important features and possible weak links of release prevention and control systems.
- Identifying data needs and information gaps in the release prevention and control area and providing recommendations for obtaining the necessary additional information through data gathering, analytical studies and experimental studies.
- Identifying potential areas where release prevention and control systems can be effectively improved in terms of safety and cost/benefit.

A staged approach was selected to accomplish the study objectives. A reference description of each type of LNG facility was developed. The system description was then used to perform an overview or first level analysis (initially a preliminary hazards analysis followed by a failure mode and effect analysis) to identify information needs and potential release prevention and control areas which may merit more detailed study. The feasibility and methods of obtaining the required additional information were investigated and a decision was made whether to perform a more detailed assessment (possibly a refined failure mode and effect analysis or, if the system detail and data warrant it, a fault tree/event tree type analysis). In conjunction with this assessment, recommendations regarding analytical and experimental studies which could be undertaken to fill information gaps were developed.

The overview assessments for each of the basic types of LNG facilities have been completed. These include:

- Export Terminal
- Marine Vessel
- Import Terminal
- Peakshaving Facility
- Truck Tanker
- Satellite Facility

The overview assessment report includes a reference system description, a preliminary hazards analysis (PHA), and a list of representative release scenarios. The system description outlines the basic process flow, plant layout, and process description. The PHA identifies the critical release prevention operations. The list of representative release scenarios provides a format for discussing potential initiating events, effects of the release prevention and control systems, information needs, and possible design changes to prevent or reduce the consequences of potential release. The representative release scenarios will form the basis for the next stage of analysis.

This report presents the more detailed analysis of the release prevention systems for an LNG import terminal. The report first briefly summarizes the import terminal overview assessment and the analysis approach used for the present study. The results of a release scenario analysis are discussed next. Estimated frequencies and release quantities are given for each release scenario along with an identification of critical release prevention components. The final section of the report analyzes release prevention system design alternatives for key release scenarios.

3.0 ANALYSIS APPROACH

As discussed in the introduction, this report covers the third step in a staged approach used to evaluate LNG release prevention. As a first step, we developed a system description of a reference LNG Import Terminal. This system description was used to perform an overview or first level safety analysis. The reference system description and the overview analysis are included in a separate report (Pelto 1982). The next two subsections provide a brief summary of the system description and a review of the overview analysis.

The third subsection describes analysis techniques used in this phase of the project -- failure modes and effects analysis (FMEA) and fault tree analysis.

3.1 SUMMARY OF SYSTEM DESCRIPTION

The reference LNG import terminal consists of three basic process areas: marine terminal and ship unloading, storage, and sendout/vaporization. Figure 3.1 is a process flow diagram of the facility which is described briefly in the following paragraphs.

3.1.1 Marine Terminal and Ship Unloading

The marine terminal for the LNG import facility consists of a dock and a 6,000-ft trestle supporting a roadway and four transfer lines. The four major transfer lines include a 42-in. main LNG transfer line from the trestle to the LNG storage facility, a 16-in. vapor return line to maintain adequate pressure in the ship's storage tanks, a 4-in. LNG recirculation line to maintain the LNG unloading lines cold when not unloading a vessel, and a 10-in. Bunker "C" fuel oil line.

Two sets of four 16-in. diameter articulated LNG unloading arms are located on deck at the terminal. Each unloading arm connects with a 24-in. transfer line which connects to a 42-in. header. The 42-in. header feeds the 42-in. main transfer line to shore. The 16-in. vapor return line connects to two 16-in. vapor loading arms. The 42-in. transfer line, the 16-in. vapor return line, and the 4-in. recirculation line all contain expansion bellows every 240 ft. The unloading arms and the three cryogenic pipelines are all of stainless steel construction.

FIGURE 3.1. Process Flow Diagram for LNG Import Facility

The marine terminal includes enough berths to accommodate two vessels, one on each side of the trestle. However, only one ship can unload at a time. A control tower overlooking the ship operations is also included at the terminal.

3.1.2 LNG Storage

LNG storage consists of two flat-bottom, double-walled, above-ground LNG storage tanks with a capacity of 550,000 bbl each. The inner tank is constructed of 9% nickel-steel; the outer tank is of carbon steel. The dimensions of the tank are:

Inner diameter	215 ft
Outer diameter	225 ft
Inner shell height	88 ft, 1 in.
Outer tank shell height	98 ft
Overall tank height	146 ft, 6 in.

The outer tank has a dome-shaped carbon steel roof. Suspended from the outer tank's roof is a metal deck which serves as the inner tank ceiling. Open pipe vents allow vapor to circulate between the inner and outer tanks. The outer tank has a maximum and minimum design pressure of 2.0 psig and 0 psig respectively. It is protected by both vacuum and pressure relief valves. The inner tank is not subjected to differential gas pressure and is stressed only by the liquid head in the tank. All piping connections to the inner tank enter through the roof of the storage tank and are supported by independent structures.

Expanded perlite, a non-flammable material, is spread evenly over the suspended ceiling and in the annular space between the inner and outer shells as insulation. A resilient fiberglass blanket is wrapped around the outside of the inner shell to alleviate the effects of the movement of the inner shell due to thermal cycling. Foamglass blocks, a load-bearing insulation, are employed as insulation for the tank bottom.

A 4-ft thick reinforced concrete slab supports each tank. An electrically heated sandbed is located between the tank and the concrete slab. This prevents the soil from freezing beneath the tank and causing frost heave. Engineered fill is located directly beneath the slab.

A concrete dike wall surrounds each storage tank. Each dike will hold approximately 1-1/3 times the capacity of one storage tank. The inside of the dike wall is lined with insulating material to reduce the evaporation rate of LNG in the event of a tank failure. A weather shield extends from the top of the concrete dike to the outer tank roof in order to keep precipitation from falling into the annular space.

Normal storage tank boiloff and vapor from LNG tanker unloading operations are handled by a gas compressor system. The vent gas compressors take cold vapor from the tanks and send it to the fuel gas and the pipeline compressors, return it to the ships storage tanks during unloading, and return it to the storage tanks as necessary to maintain the tank pressure. There are four, single-stage, centrifugal, vent-gas compressors. One is required to handle normal boiloff, three are required during tanker unloading, and one is a spare. Excess gas from the

vent gas compressors is heated and then compressed by three, gas-driven, reciprocating fuel gas compressors. Plant fuel is taken from the discharge of these units. Any additional gas is further compressed and discharged into the pipeline by three gas-driven, reciprocating pipeline compressors.

Each storage tank contains two submerged primary sendout pumps which boost the LNG to 60 psig. These are followed by 10 secondary pumps. The secondary pumps are submersible, pot-mounted, 15-stage units which raise the LNG up to 1300 psig.

3.1.3 LNG Vaporizers

Vaporization for the import terminal is provided by nine vaporizers: five falling-film open-rack seawater vaporizers, and four submerged combustion gas-fired vaporizers. These vaporizers provide the plant with a total output capacity of 1 billion scfd of gas.

In the open-rack vaporizers, LNG is introduced through manifolds at the bottom of banks of vertical panels constructed of special extruded fins. The LNG passes upward inside the tubes where it is heated by the water which falls as a film over the outside of the panels. LNG emerges in its gaseous form at the top.

The submerged combustion vaporizers are designed such that the burners exhaust hot combustion gases downward through a downcomer and into a water bath below the liquid surface. The exhaust forms bubbles in the water causing turbulence, mixing, and a "lifting" action. This lifting action forces the water up at a high velocity through an annular space created by a weir around the downcomer. The water flows over the top of the weir and into a more quiescent tank. A heat exchanger tube coil for the LNG is located in the annular space between the weir and the downcomer where it is scrubbed by the warm gas-water mixture thus transferring the heat to the LNG and vaporizing it. The vaporizers consume between 1.5% and 2.0% of the LNG vaporized as fuel.

The inlet piping and the outlet piping to the outlet header are all stainless steel construction on both the open rack and submerged combustion units. The heat exchange panels in the open rack vaporizers are aluminum and the LNG tube coil in the submerged combustion units is stainless steel. Gas piping downstream of the outlet headers is carbon steel.

3.1.4 Safety Systems

The plant emergency shutdown system (ESD) has three shutdown circuits: the Master Emergency Shutdown (MES), the Vaporizer Emergency Shutdown (VES), and the Offshore Emergency Shutdown (OES). These systems automatically shut-down and isolate portions of the facility or, in the case of the MES, the entire facility. It takes about 30 seconds to shut down the plant once the ESD is activated. The shutdown systems are activated by detectors located throughout the plant, by certain process variables, or by plant operators.

Combustible gas detectors, UV flame detectors, and temperature sensors are located throughout the plant area. These detectors activate alarms which indicate the exact location of a spill or fire on a graphic panel in the control room. The fire control and prevention system consists of fixed and portable dry chemical extinguishers, expansion foam generators, and a fire water system.

All gas lines and gas handling equipment can be vented to the vent stack through the nitrogen-purged vent gas header. Gas is not normally vented except in the case of an emergency shutdown when the Master Emergency Shutdown automatically vents all gas lines and gas handling equipment. The LNG sendout pumps are vented back to the storage tank. The storage tank pressure relief valves relieve to the atmosphere. The relief valves will open only in emergency situations to protect the tank from overpressure. Any sections of LNG lines, or vessels which can be isolated with LNG in them and allowed to warm are protected by relief valves which vent to the storage tank.

All piping or equipment in contact with LNG or cold vapor is constructed of stainless steel, 9% nickel steel, or an aluminum alloy. A minimum of flanged connections are used; welded connections are used where possible. All low pressure valves in cryogenic service are butterfly valves and all high pressure cryogenic valves are globe valves. A minimum of bellows expansion joints are used in the plant; expansion loops or designed-in flexibility are used where possible.

3.2 REVIEW OF THE OVERVIEW ANALYSIS

The purpose of the overview analysis was to facilitate more detailed analyses, by separating the particular release prevention systems and equipment which merit additional study from those that do not have a significant effect on the safety of the facility.

The first step in the overview analysis was to identify the systems and components which contain LNG or natural gas and determine the flow rate and inventory for each one. These results are shown in Table 3.1. The storage tanks and the unloading system have the highest flow rates and inventories.

This was followed by a preliminary hazards analysis (PHA). The effect of initiating events such as equipment failures, operator errors, and external events were qualitatively analyzed. For active type components (e.g., motors, electrical components, etc.) it was typically found that two or more separate failures were required to result in any significant release. For most of the passive type components (e.g., piping, process vessels, etc.), a single failure can result in an initial release, but emergency shutdown systems are designed to stop the initial release. One obvious exception is the gross failure of the inner storage tank.

TABLE 3.1. System Capacities and Flow Rates (Pelto et al. 1982)

System	Component	No. of Components	Component Capacity	Flow Rates		Operating Condition	
				In	Out	Pressure	Temperature
Marine Terminal and Unloading System	42-in. LNG Transfer Line		10,000 bbl	53,000 gpm	53,000 gpm	100 psig	-258°F
	4-in. LNG Recirculation Line		93 bbl	1,500 gpm	1,500 gpm	60 psig	-258°F
	16-in. Vapor Return Line		2,800 scf	16 MMscfd	16 MMscfd	10 psig	-258°F
Storage	Storage Tank	2	550,000 bbl	53,000 bbl	4,550 gpm	0.8 psig	-258°F
Vaporization System	Seawater Vaporizers	5	37 bbl	4,550 gpm	550 MMscfd	1,300 psig	-252°F to 30°F
	Submerged Gas-Fired Vaporizers	4			450 MMscfd	1,300 psig	-252°F to 30°F
Compressors	Boiloff Compressor	4		26-52 MMscfd ^(a)	26-58 MMscfd	10 psig	-200°F to 152°F
	Fuel Gas Compressor	3		5-36 MMscfd	5-36 MMscfd	150 psig	-285°F
	Pipeline Compressor	3		5-36 MMscfd	5-36 MMscfd	1,300 psig	200°F
Sendout Pumps	Primary In-tank Pumps	2			4,000 gpm	60 psig	-258°F
	20-in. Tank Outlet Line		29 bbl	4,000 gpm	4,000 gpm	60 psig	-258°F
	Secondary Pumps	10		4,550 gpm	4,550 gpm	1,300 psig	-252°F
	25-in. Secondary Pump Sendout Line		303 bbl		4,550 gpm	1,300 psig	-252°F

(a) Flow rate during unloading operations.

At the conclusion of the overview assessment, we developed a list of representative release scenarios that can occur at an LNG import terminal. Some modifications were made to this list in this analysis, but the modified list shown in Table 3.2 is generally the same as the one developed in the overview.

TABLE 3.2. Representative Release Scenarios for an LNG Import Terminal

Marine Terminal and Unloading Section

1. Loading Arm Rupture
2. Transfer Line Rupture

Storage Section

3. Gross Failure of a Storage Tank
4. Storage Tank Overfill
5. Storage Tank Overpressure (including rollover)
6. Storage Tank Underpressure
7. Inlet/Outlet Line Rupture
8. Sendout Pump Rupture

Vaporization Section

9. Tube or Panel Rupture
10. Control Failure and Outlet Line Rupture
11. Main Gas Line Rupture

3.3 RELEASE SCENARIO ANALYSIS TECHNIQUES

The purpose of the release scenario analysis was to provide an estimate of the probability of the release scenarios so that a relative comparison could be made. We considered several possible analysis methods to accomplish this.

Historical system operating data can be used to estimate future accident frequency potentials. However, there are at least two problems in extrapolating historical experience to future operations of LNG systems. The number of LNG systems in operation and the number of accidents are not sufficient from a statistical viewpoint to estimate accident frequencies. Additionally, the LNG industry operates in a continuously changing environment with respect to technology and regulations. Only the most recent operating history may be applicable to future considerations.

Another approach to quantify identifiable hazards is to analyze the failure rate of individual components of the system being studied. By carefully utilizing generic component failure data, one can estimate system failure frequencies. Unfortunately one of the shortcomings of using generic failure rate information is that some components of LNG systems do not have generic counterparts in other industries. Even for components with generic counterparts in other industries the available data on rate of failure is often limited. Despite these drawbacks it was determined that this technique could provide valuable information for relative comparisons of release scenarios and release prevention systems. Two analysis techniques were used to evaluate the release scenarios previously identified--failure modes and effects analysis (FMEA), and fault tree analysis.

3.3.1 Failure Modes and Effects Analysis

The first step in determining the probability of system failure is to examine the ways the system, and in turn, each component, can fail. A failure modes and effects analysis (FMEA) is an inductive method that systematically analyzes component failure modes and identifies the resulting effect on the system. A FMEA is usually presented in tabular form and can include failure rates for each of the failure modes. Failure modes and effects analyses for the unloading, storage, and vaporization sections of the reference LNG import terminal are included in Appendix A.

Over a dozen sources were used to obtain the failure rate information included in the FMEAs. Only one of these sources dealt specifically with land-based LNG facilities (Welker 1979). For most components, generic failure rate information was used. In most instances these failure rates came from studies in the nuclear industry (USNRC 1975) and the chemical processing industry (Lees 1971, Lees 1973, Kletz 1973, Kletz 1975). In addition, some information was obtained from a study of safety on LNG ships (Welker 1976). Most of the failure rate information in this last study is generic and was obtained from previously mentioned sources. Table 3.3 contains a summary of the failure rate information used in the release scenario analyses.

TABLE 3.3. Generic Failure Rates for Components of LNG Import Terminals

Component	Failure Mode	Faults/Hr	Reference
Pump	Rupture	1×10^{-8}	(SAI 1975, Browning 1978)
	Fails to Stop	1×10^{-7}	(Welker 1976)
Compressor	Rupture	1×10^{-8}	(SAI 1975, Browning 1978)
	Fails to Run Normally	3×10^{-4}	(Welker 1979)
Vaporizer	Tube or Panel Rupture	1×10^{-5}	(Welker 1979)
	Control System Failure	1×10^{-4}	(Welker 1979)
Pipe Section 3 in dia.	Rupture	1×10^{-10}	(USNRC 1975)
Storage Tank	Rupture	1×10^{-9}	(SAI 1975, Atullah 1980)
		1×10^{-10}	
	Cold Spot	1×10^{-5}	(Welker 1979)
Valve	Rupture	1×10^{-9}	(USNRC 1975, Welker 1979)
	Fails Closed or Is Misdirected Toward Closed	5×10^{-5}	(Lees 1973, Lawley 1974, Browning 1973)
	Fails Open or Is Misdirected Toward Open	5×10^{-5}	(Lees 1973, Lawley 1974, Browning 1973)
Expansion Joints	Rupture	1×10^{-7}	(Welker 1976, SAI 1975)
Pipe Fittings (flanges, elbows, tees, etc.)	Rupture	1×10^{-8}	(USNRC 1975)
Loading Arm	Rupture	3×10^{-7}	(Welker 1976, SAI 1975)
Sensors/Detectors			
Flow	Fail Dangerously	2×10^{-4}	(Lees 1973, Lawley 1974, Browning 1973, 1978, Anyakora 1971)
Level	Fail Dangerously	2×10^{-4}	(Lees 1973, Lawley 1974, Browning 1973, 1978, Anyakora 1971)
Pressure	Fail Dangerously	1×10^{-4}	(Lees 1973, Kletz 1977, Anyakora 1971)
Temperature	Fail Dangerously	1×10^{-4}	(Lees 1973, Browning 1978, Anyakora 1971)
Combustible Gas	Fail Dangerously	4×10^{-5}	(Welker 1979, St. John 1978)
UV Radiation	Fail Dangerously	1×10^{-4}	(Welker 1976, 1979, Lees 1973, Anyakora 1971)
Low Temperature	Fail Dangerously	1×10^{-4}	--

TABLE 3.3. (contd)

Component	Failure Mode	Faults/Hr	Reference
Controller, Limit Switch (decision-making unit)	Fail Dangerously	3×10^{-5}	(Lees 1973, Browning 1973, 1978, Anyakora 1971, Fisher 1973)
Alarm	Fails to Operate.	5×10^{-5}	(SAI 1975, Lawley 1974, Browning 1978)
Relief Valve	Fails to Open	1×10^{-6}	(Lawley 1974, Kletz 1972, 1977, USNRC 1975)
	Opens Prematurely	1×10^{-5}	(Lawley 1974, USNRC 1975)
ESD Circuitry (based on failure of a relay to energize or to open)	Fails to Energize	1×10^{-5}	(USNRC 1975, Welker 1976, Lees 1973, Anyakora 1971)
	Fails to De-energize (fail-safe system)	1×10^{-7}	(USNRC 1975, Welker 1976, Lees 1973, Anyakora 1971)
Operator ^(a)	Fails to Respond Correctly to Changes in Important Process Variable, Complex System	3×10^{-1}	(Kletz 1972, 1973, 1975)
	Fails to Respond Correctly to Changes in Important Process Variables, Simple System	3×10^{-1}	(Lawley 1974, Kletz 1973, 1975)
	Fails to Respond Promptly and Correctly to Emergency Alarms, Simple System	1×10^{-2}	(Kletz 1975)
	Monitor or Inspection Error, Fails to Notice a Release, or Severe Equipment Problems	1×10^{-1}	(Welker 1976, Kletz 1972, 1973, 1975)
	Fails to Follow Standard Operating Procedure, Testing, or Maintenance Procedures	5×10^{-2}	(Kletz 1972, 1973, 1975)

(a) Faults per demand.

As indicated previously some components of LNG systems do not have generic counterparts in other industries. This makes the analysis of these components more difficult, if not impossible. LNG storage tanks are the most obvious example of this problem. The range shown in Table 3.3 is considered to be a best estimate of the failure rate.

Because operation of LNG import terminals is not completely automated, the operator is a "component" of the system. Some data on human reliability has been developed (Keltz 1973, Keltz 1975, USNRC 1975). Unfortunately, operator responses are often not discrete events which can be described simply as a

success or failure. If the operator takes a "long" time to detect a release or properly respond to an emergency situation, then the operator has not satisfied the design philosophy of the system. This type situation was considered a failure in this analysis.

Operator failure rates for various tasks involved in operation of an LNG terminal are also included in Table 3.3. These rates are based on available data on human reliability and our engineering judgment as to the difficulty of the task.

For simple release scenarios consisting of a single component failure (a pipe break, for instance), the number of failures that can be expected over a specified time interval is simply the failure rate multiplied by the time interval. For more complex release scenarios fault tree analysis was used to estimate the expected number of failures.

3.3.2 Fault Tree Analysis

Fault tree analysis is a deductive process. The analyst assumes the occurrence of an event as the top undesired event, constituting system failure. More often than not in our analysis the top event was a large release of LNG from a particular system. After selecting the top event the analyst systematically works backward to identify component faults (basic events), which could cause or contribute to the undesired top event.

Standard symbols shown in Figure 3.2 are used to express the relationship of individual component failures (basic events) to the overall system failure (top event). Multiple events which individually cause or contribute to the top event are connected by an OR gate. Events which must occur concurrently in order to cause or contribute to the top event are connected by an AND gate.

For simple fault trees, the expected number of system failures can be calculated directly from the component failure rates following the logic of the fault tree. For independent basic events connected by OR gates, the failure rate of the system can be obtained by summing the component failure rates. For independent basic events connected by an AND gate, the system failure rate is equal to the product of the component failure rates. For generally reliable systems the expected number of failures is equal to the failure rate multiplied by the specified time interval.

EVENT REPRESENTATIONS

THE RECTANGLE IDENTIFIES AN EVENT THAT RESULTS FROM THE COMBINATION OF FAULT EVENTS THROUGH THE INPUT LOGIC GATE.

THE HOUSE IS USED AS A SWITCH TO INCLUDE OR ELIMINATE PARTS OF THE FAULT TREE AS THOSE PARTS MAY OR MAY NOT APPLY TO CERTAIN SITUATIONS.

THE CIRCLE DESCRIBES A BASIC FAULT EVENT THAT REQUIRES NO FURTHER DEVELOPMENT. FREQUENCY AND MODE OF FAILURE OF ITEMS SO IDENTIFIED ARE DERIVED FROM EMPIRICAL DATA.

LOGIC OPERATIONS

AND GATE DESCRIBES THE LOGICAL OPERATIONS WHEREBY THE COEXISTENCE OF ALL INPUT EVENTS IS REQUIRED TO PRODUCE THE OUTPUT EVENTS.

THE DIAMOND DESCRIBES A FAULT EVENT THAT IS CONSIDERED BASIC IN A GIVEN FAULT TREE. THE POSSIBLE CAUSES OF THE EVENT ARE NOT DEVELOPED WHETHER BECAUSE THE EVENT IS OF INSUFFICIENT CONSEQUENCE OR THE NECESSARY INFORMATION IS UNAVAILABLE.

OR GATE DEFINES THE SITUATION WHEREBY THE OUTPUT EVENT WILL EXIST IF ONE OR MORE OF THE INPUT EVENTS EXISTS.

FIGURE 3.2. Fault Tree Symbols

For more complex fault trees cut set analysis can be used to determine the expected number of system failures. A minimal cut set is a collection of primary events (component faults) such that if they all occur they are sufficient

to cause the top event and the simultaneous existence of each is necessary to cause the top event. The expected number of system failures (top events) can be calculated by summing the expected number of system failures resulting from each cut set. For most LNG systems the minimal cut sets can be developed from the fault trees by inspection. The expected number of system failures for each cut set can then be calculated using the equations in Table 3.4. These equations were adapted from those developed by Fussell (1975) to hand calculate system reliability and safety characteristics.

For our analysis we were primarily interested in the expected number of large releases of LNG, which was the top event in many of the fault trees. A large release of LNG generally requires loss of LNG containment (e.g., a pipe rupture) plus failure to stop the release in a short time period. Emergency systems are designed to automatically shutdown the system and isolate the release. For these system and others that operate only in response to an initiating event, the probability of failure on demand is the average unavailability of the system over the specified time interval. For many of the release scenarios, calculating the expected frequency of top events required calculating the expected number of failures of processing equipment and instrumentation (initiating events) and the unavailability of emergency and monitoring equipment.

Many component faults can be detected and repaired. To determine the unavailability of these components, the dead time (time the fault event exists) must be known. For faults that are detected immediately, the dead time is equal to the repair time. Other faults will be detected during the course of plant operations and the dead time is equal to the detection time plus the repair time. Fault detection and repair times used in this study are shown in Table 3.5 and 3.6.

Some LNG emergency systems have components whose faults are unannounced (not detectable). Because these systems are important to the safety of the plant, these components are tested at regular intervals to detect faults. For these components, the unavailability is lowest right after the test and gradually increases until the next test. The average unavailability of these components can be approximated by using a mean dead time equal to one-half the test interval plus the repair time (Kletz 1973).

TABLE 3.4. Equations Used to Calculate System Safety Characteristics

Basic Events		Minimal Cut Sets	Top Events
<u>Non-Repairable Events</u>			
$\bar{a}_i \leq \lambda_i$ (very close if $\lambda_i t \ll 0.1$)		$a_k = \prod_{i=1}^{n_k} \bar{a}_i$	$a_T \leq \sum_{k=1}^n \bar{a}_k$
$enf_i = \bar{a}_i$		$enf_k = \int_0^t rof_k dt$	$\lambda_T \leq \sum_{k=1}^n \lambda_k$
		$rof_k \leq \bar{a}_k \sum_{i=1}^{n=k} \frac{\lambda_i}{\bar{a}_i}$	$enf_T \leq \sum_{k=1}^n enf_k$
		$\lambda_k = rof_k$	$rof_T \leq \sum_{k=1}^n rof_k$
<u>Repairable Events</u>			
$\bar{a} \leq \lambda_i T_i$ (very close if $t \gg 2T_i$)			
$enf_i = \int_0^t rof_i dt = \int_0^t (i - \bar{a}_i) \lambda_i dt$			
$enf_i = \lambda_i t$ (when $\bar{a}_i$ is small)			
<u>Definitions</u>			
λ	Failure Rate	Defined so λdt is the probability the fault event will occur between t and dt given it has not occurred before t .	
λ_i	Average Failure Rate		
$\bar{a}$	Unavailability	Probability the fault event exists at some specified time.	
T	Mean Dead Time	Average time the fault event exists.	
enf	Expected Number of Failures	Average number of occurrences of the fault event during a specified time interval.	
rof	Rate of Failure	Expected number of occurrences of the fault event per unit time.	
t	Time		

TABLE 3.5. Fault Detection Times for Some Components of LNG Systems

Components	Detection Time (hrs)
Process Equipment and Instruments Where a Serious Fault Will Be Obvious Immediately to the Operator Control Systems, Pumps, Compressors, Critical Process Indicators, etc.	0-1
Process Equipment and Instruments Where a Serious Fault Will Be Detected by the Operator during the Course of Normal Plant Operation (Process Indicators)	1-10
Process and Emergency Equipment and Instruments Where a Serious Fault Will Be Detected by the Operator Only When an Abnormal or Intermittent Condition Exists (Process Indicators and Alarms, Limit Switches, Trip Valves, etc.)	10-100
Emergency Equipment and Instruments Where A Serious Fault Will Be Detected By the Operator Only When an Abnormal or Intermittent Condition Exists (Process Indicator, Limit Switches, Alarms, Valves, Relief Valves)	80 ^(a) -1000 ^(b)

(a) 80 hours corresponds to weekly testing.

(b) 1000 hours corresponds to a test interval of three months.

TABLE 3.6. Fault Repair Time for Some Components of LNG Systems

Component	Failure Mode	Average Repair Time (hrs)
Pumps, Compressors	Fails to Run Normally	50
Valves	Fails to Operate Correctly	24
Instrumentation Systems (detectors, controllers, alarms)	Fails to Operate Correctly	8-24

When components are down for testing or out of service for some other reason the unavailability is equal to 1.0. This should be included in determining the overall average unavailability. If a system is down for testing 1 hr/yr, the unavailability due to testing is 1×10^{-4} . For most emergency systems at LNG facilities, we assumed that testing took place only when the processing facility is shut down.

4.0 RELEASE SCENARIO ANALYSIS

This section summarizes our attempt to predict the probability of each of the release scenarios using FMEA and fault tree analysis. This analysis generally consisted of two parts: 1) predicting how many times each scenario would occur in one year, and 2) determining the reliability and efficiency of the emergency sensors and shutdown systems.

Generally the functions of the sensors and the shutdown systems did not vary much for each scenario. For this reason, a general subsection on the emergency shutdown systems (ESD) is included prior to discussion of the release scenarios.

Where possible, the following information was developed for each release scenario.

1. The number of times the scenario will occur in one year.
2. The size of the release assuming the ESD is activated promptly and functions properly.
3. The probability of ESD failure.
4. The size of the release after 10 minutes of uncontrolled flow.
5. The critical components in both the process and emergency systems.
6. Possible operator actions in the event of ESD failure.

The size of the potential releases was calculated based on the following assumptions.

1. Maximum design flow rates and inventories.
2. When possible, the release occurs from the point in the system which results in the largest spill.
3. Guillotine pipe breaks.

The probability of some of the simple release scenarios was calculated directly from the FMEA (pipe ruptures, vaporizer tube or panel ruptures, etc.) For the more-complex scenarios, fault tree analysis was used.

A word of caution concerning the use of these results should be included here. The purpose of this analysis was not to identify all possible scenarios (an impossible task, as no extent of analysis will assure that all failure modes have been examined), but to examine potential scenarios to provide relative comparisons for release-prevention system effectiveness. The following sections discuss the release scenario analysis for the basic process areas of the reference import terminal. The fault trees and supporting calculations for the results reported in this section are given in Appendix B. Table 4.1 summarizes the results of the release scenario analysis.

4.1 EMERGENCY SHUTDOWN SYSTEMS

An emergency shutdown at an LNG import terminal consists of three elements: detection, activation, and isolation. Emergency situations are detected by process instrumentation (flow, temperature, pressure, and level sensors), emergency instrumentation (combustible gas, fire, and low-temperature detectors), and the plant operators. The process and emergency sensors are connected to controllers or limit switches which, when a preset value is exceeded, activate audible and visible alarms in the control room. The emergency shutdown system (ESD) is activated either automatically by the controllers or limit switches, or manually by the operator in response to an alarm or visual recognition of the emergency. The ESD circuit consists of a combination of relays, signal transmission lines, and power supplies which when activated stops the flow of LNG or natural gas by shutting down pumps and compressors and closing block valves. The ESD also opens vent valves to vent various pieces of equipment to the vent stack.

Most detectors and sensors provide visual and audible alarms within 10 seconds. For automatic activation of the ESD, a discretionary time delay of 30 seconds is provided to minimize accidental activation. This delay can be terminated at any time by manual activation of the ESD controls. Closure time for all valves except those in the LNG transfer line is less than 10 seconds. LNG transfer line valves have a programmed closure to minimize fluid hammer and close within 20 seconds. Significant flow through all pumps and compressors

TABLE 4.1. Summary of Results of LNG Import Terminal Release Scenario Analysis

Release Scenarios	Expected Number of Events per Year	Release Occurs and Is Not Stopped in 1 Minute by ESD	Maximum Release Size (Equivalent Gallons of LNG)		Material Release	Critical System Components
		Expected Number of Events per Year	1 Minute	10 Minutes		
Marine Terminal and Unloading System						
1. Rupture of Loading Arm or Components During Unloading	2×10^{-3}	2×10^{-5}	15,000	130,000	LNG	Loading Arm, Swivel Joints, Operator
2. Rupture of Main Transfer Line or Components During Unloading	4×10^{-3}	1×10^{-6}	500,000	970,000	LNG	Expansion Joints
Storage System						
3. Gross Failure of a Storage Tank	1×10^{-5}	--	--	23,000,000	LNG	Storage Tanks
4. Storage Tank Is Overfilled	2×10^{-3}	--	-- (a)	-- (a)	Natural Gas/LNG	Level Detectors and Transmitters, Filling Valve, Operator
5. Storage Tank Is Overpressured During Filling and Vent/Relief Valves Open	5×10^{-1}	--	400-4,000 (b)	4,000-40,000 (b)	Natural Gas	Pressure Detector, Pressure Controller, Operator
5a. Storage Tank Is Overpressured During Filling and Vent/Relief Valves Fail	$1 \times 10^{-6(d)}$	--	200-600 (c)	1,200,000-3,500,000 (c)	Natural Gas/LNG	Vent/Relief Valves
6. Storage Tank Is Underpressured and Vacuum Relief Valves Open	2×10^{-3}	--	--	--	--	Pressure Controller, Pressure Detector
6a. Storage Tank Is Underpressured and Vacuum Relief Valves Fail	$1 \times 10^{-8(e)}$	--	200-600 (c)	1,200,000-3,500,000 (c)	Natural Gas/LNG	Vent/Relief Valves
7. Rupture of Storage Tank Inlet Line	2×10^{-6}	4×10^{-8}	61,000	538,000	LNG	Filling Valves, Pipe Fittings, Operator
8. Rupture of Storage Tank Outlet Line	3×10^{-5}	5×10^{-7}	5,000	41,000	LNG	Outlet Valves, Pipe Fittings, Operator
9. Rupture of Secondary Pump or Associated Piping and Valves	2×10^{-3}	3×10^{-5}	5,000	41,000	LNG	Secondary Pumps, Section and Discharge Valves
Vaporization Section						
10. Vaporizer Tube (Sub. Comb.) or Plate (Open Rack) Rupture	5×10^{-1}	5×10^{-2}	1,700	9,100	Natural Gas/LNG	Heat Exchanger Tubes or Plates
11. Rupture of Vaporizer Outlet Line From Cold Gas (Vaporizer Control Failure)	8×10^{-2}	8×10^{-4}	2,000	11,000	Natural Gas/LNG	Temperature Control System, Temperature Detectors, Operator
12. Rupture of Main Vaporizer Header and Gas Line to the Pipeline	1×10^{-4}	1×10^{-6}	16,000	98,000	Natural Gas	Piping

(a) Could cause failure of outer shell and roof. Failure of inner tank is possible.

(b) Lower number is typical vapor generation rate during filling. Higher number is maximum relief valve capacity.

(c) Will cause rupture of outer tank roof/wall joint. Failure of inner tank is unlikely. Release rate is for an open top tank for 100 hours (time to pump out the tank is full).

(d) Assumes four relief valves or the vent valve required to relieve pressure.

(e) Assumes only one vacuum relief valve is necessary to equalize pressure.

will cease almost immediately upon shutdown. Overall, less than 1 minute is required to totally shutdown the plant when an emergency situation occurs. These are the design shutdown times thought to be typical of new LNG import terminals.

The reference LNG import terminal has three major shutdown circuits: the Offshore Emergency Shutdown (OES), the Vaporizer Emergency Shutdown (VES), and the Master Emergency Shutdown (MES). The OES has two levels of shutdown: Level I, which stops all unloading operations and Level II, which disconnects the ship in addition to stopping all loading operations. The MES shuts down the entire plant and the VES stops sendout and vaporization operations.

During normal operation the ESD circuits at the reference import terminal are energized from a separate "Uninterruptable Power Supply." When the ESD is activated the circuits are de-energized and the emergency actions described above are initiated. This arrangement is often referred to as "fail-safe" because many system faults (such as switches or relays failing to make contact, breaks in signal transmission lines, power failure, etc.) will result in spurious shutdowns and the ESD can be repaired. This is inherently safer than circuits which must be energized to operate because system failures cannot be detected until the system is activated (Fisher 1973, Bennett 1972, Kletz 1972). However, the higher rate of spurious trips associated with fail-safe systems can be a nuisance to plant operation. Figure 4.1 shows a simple fail-safe emergency shutdown system.

Some faults in the ESD circuitry will not be detected until the system is tested. Failure of normally closed relay contacts to open on demand was used to represent these faults. Such faults occur about 1×10^{-7} /hr. The OES can be checked prior to each unloading operation; however, for this study a testing frequency of once per month was assumed. The average unavailability of the ESD circuitry for the OES is then 4×10^{-5} . The VES and MES circuitry cannot be checked for operability at specified intervals since checking the circuitry would result in an unscheduled shutdown. A test interval of 3 months, which gives an average unavailability of 1×10^{-4} was assumed. Spurious trips resulting from detection equipment or ESD circuitry faults will occur about once per year. If the LNG processing systems are operated while the ESD is being repaired, the unavailability of the system will increase to 1×10^{-3} .

DURING NORMAL OPERATION CURRENT IS FLOWING IN BOTH CIRCUITS A AND B

Components:

1. Detector	4. Relay
2. Controller/Limit Switch	5. Uninterruptable Power Supply
3. Power Supply	6. Valve Pump, Compressor, etc.

FIGURE 4.1. Simple Fail-Safe Emergency Shutdown System

Figure B.1 shows fault trees for failure of the OES, MES, and VES shutdown systems. While it is desirable that pumps (or compressors) are shut down and block valves are closed, it was considered a successful shutdown if either one occurs and flow is stopped. As a result, ESD circuitry failures dominate over processing equipment failures.

Although ESD systems are activated automatically in some cases, most often they are activated by the operator. Unfortunately, operator responses are not discrete events which can be described simply as a success or failure. If it takes the operator too long to detect and properly respond to an LNG release, then the operator has failed to satisfy the design philosophy of the safety system. This type situation was considered to be a failure.

The probability that the plant will not be shut down quickly and safely in the event of an emergency varies from 10^{-1} to 10^{-4} per demand depending on the situation and the system involved. Figure B.2 is a fault tree for a fully

automatic emergency shutdown. The failure rate for fully automatic emergency shutdown is 10^{-3} to 10^{-4} per demand. The critical failures in fully automatic systems are sensor and limit switch failure and ESD circuitry failure.

The failure rate for an automatic or manual detection, manual activation emergency shutdown is about 10^{-2} per demand except for some situations where the cause of the emergency is not obvious and the failure rate is higher--about 10^{-1} per demand. In both cases operator response to the emergency situation is the critical step in the shutdown.

A fault tree for a manually activated emergency shutdown is shown in Figure B.3. The following section discusses release scenarios at LNG import terminals, including the effects of the emergency shutdown systems. Specific ESD fault trees are included in the fault trees developed for many of these scenarios.

4.2 MARINE TERMINAL AND SHIP UNLOADING SECTION

Only two release scenarios were evaluated for this section of the plant: rupture of an unloading arm during LNG transfer and rupture of the main transfer line during LNG transfer. Generic failure data suggest both of these events will occur approximately 10^{-3} times per year.

4.2.1 Loading Arm Rupture

Potential failure modes for the loading arms include rupture of the arm due to excessive ship motion, fatigue from thermal and pressure cycling, failure of the swivel joints, and rupture of the valves associated with the loading arms. A fault tree for a large release of LNG from the loading arms is shown in Figure B.4. Supporting calculations are shown in Table B.2. Based on generic failure rate data this will occur about 2×10^{-3} times per year.

Loading arms are typically equipped with sensing devices that indicate when the loading arms are approaching their maximum reach. These sensors sound audible alarms and automatically activate the ESD. Analysis indicates that in the event of excessive ship motion the probability of the ESD failing to shut down unloading operations and disconnect the loading arms is 1×10^{-4} per demand.

If a loading arm ruptures, the operator can activate the emergency shutdown system and if the ESD operates as designed, the unloading system will shut down in less than 1 minute, releasing about 15,000 gallons of LNG. If the unloading system is not shut down and isolated, about 130,000 gallons will be released in 10 minutes. Our analysis indicates that the probability of this occurring is 1×10^{-2} per demand, resulting in a frequency of 2×10^{-5} times per year for this scenario. Operator recognition of the release and subsequent activation of the ESD is the critical step. Containment for small or medium size spills is provided under the dock. Spills from the loading arms would normally occur over water between the ship and the dock. If the ESD fails, a ship crewman can manually shutdown the pumps, close the main ship unloading valve, or close the valves on the pumps to stop flow.

If excessive ship motion is the cause of the loading arm rupture, it is reasonable to assume that more than one and possibly all of the loading arms may fail. If all four arms rupture, 60,000 gallons would be released in 1 minute and 520,000 gallons would be released in 10 minutes.

4.2.2 Transfer Line Rupture

Generic failure rate data indicate the most likely cause of failure of the transfer line is a rupture of an expansion joint. The failure rate for expansion joints is 2 or 3 orders of magnitude higher than other components of the line. A fault tree for failure of the transfer line is shown in Figure B.5. Supporting calculations are given in Table B.3.

In the event an expansion joint or other transfer line component ruptures (about 4×10^{-3} times per year), the pressure in the line will fall sharply and a pressure limit switch will activate the ESD system and sound alarms in the control room. A flow limit switch will also activate alarms and depending on the location of the break, gas detectors or low-temperature detectors may be activated. If the unloading system is shutdown and isolated in 1 minute, the size of the release would be about 500,000 gallons. If the flow is not stopped, nearly 1,000,000 gallons would be released in 10 minutes. This will occur approximately 3×10^{-4} times per demand, resulting in a probability of 1×10^{-6} per year for this scenario. If the ESD fails, an operator can manually close the

main block valve in the transfer line or the loading arm valves to stop flow. In addition, a ship crewman can stop flow as indicated under "Loading Arm Rupture."

For the reference facility, no containment is provided under the main transfer line until it reaches the shore. The transfer line can be drained into the containment area under the dock, but this will hold only a small percentage of the LNG in the line.

4.3 LNG STORAGE

LNG storage tanks deserve special consideration because of the magnitude of potential releases from them. Each tank when full contains 550,000 barrels. A simplified fault tree for failure of an LNG storage tank is shown in Figure B.6 in Appendix B. A major release from an LNG tank can result from internal events (metal fracture, overpressure, underpressure, tank overfill, or piping failure) or external events (earthquake, severe weather, sabotage, adjacent fire, or airplane impact).

Unfortunately, the operating history of LNG tanks is too short to provide meaningful failure information. These tanks are unique to the point that generic data for other low-pressure storage tanks is not entirely applicable; however, we were able to make some comparisons with this information. In addition, we analyzed several operating scenarios (internal events) that could affect the structural integrity of the tanks. The mechanical design aspects of LNG storage tanks and their ability to withstand both internal and external events are covered in a separate report (Fecht et al. 1982) and are covered only briefly in this report.

Cooldown and heatup of storage tanks are delicate operations and could cause the tanks to fail, but the quantity of LNG present during these operations is small. These procedures were not considered further.

4.3.1 Gross Failure of a Storage Tank

The only failure (loss of liquid containment) of an LNG land based storage tank occurred in Cleveland in 1944. The cryogenic shell of the tank that failed

was constructed of 3 1/2% Ni steel, a material no longer considered acceptable for LNG tanks. There has not been a failure of an LNG tank constructed of currently acceptable materials (9% Ni steel, aluminum, or concrete) that has resulted in a large release of LNG. One LNG tank collapsed during construction and another during maintenance. In neither case was any LNG in the tank.

The operating history of land-based LNG storage tanks constructed of currently acceptable materials is approximately 1000 years. The accumulated years of all cryogenic tank experience is somewhat greater and several serious failures have occurred, primarily with smaller spherical tanks (A. D. Little 1971). Atallah (1980) reports the frequency of fires/explosions in LNG tanks to be 3.3×10^{-6} per year. Table 4.2 lists some serious incidents that have occurred with cryogenic storage tanks, including LNG tanks.

A review of serious incidents associated with petroleum tanks at refineries in the past 10 years, as reported in Chemical Engineering, Hydrocarbon Processing, and the Oil and Gas Journal, indicates such incidents occur about 1×10^{-4} times per tank-year. Atallah (1980) reports a similar frequency for fires/explosions in petroleum refineries. These incidents almost always include failure of a storage tank and a large fire. Other failures that did not cause significant damage may have occurred but may have not been reported.

In their risk assessment of the proposed import terminal at Oxnard, California, SAI assumed that LNG tank ruptures occur approximately 1×10^{-6} times per tank year (Science Applications, Inc. 1975). Based on data for petroleum refinery tanks and our analysis of some operating scenarios that could lead to failure of the storage tank (discussed below), it appears that more than 1×10^{-6} failures (maybe 1×10^{-5}) per tank year can be expected.

4.3.2 Storage Tank Overfill

An overflow of liquid onto the insulation above the suspended ceiling is a highly critical situation. As the liquid spills over into the relatively warm vapor space above the insulation, a rapid evolution of vapor will occur. This will probably cause opening of the tank pressure relief valves, and if their capacity is insufficient the tank roof may fail from overpressure. The carbon

TABLE 4.2. Serious Incidents That Have Occurred with Cryogenic Storage Tanks (Little 1971, Vielvoye 1979, Chem. Eng. 1978)

- Failure of an LNG storage tank at Cleveland Ohio (1944). Tank was constructed of 3 1/2% Ni steel.
- Failure of a large refrigerated propane storage tank in Saudi Arabia (1962).
- Rupture of the outlet line of a 21,000-bbl LOX tank at Kennedy Space Center in Florida (1966) over 17,000-bbl of liquid was released.
- Failure of a liquid withdrawal in the annular space of a 7000-bbl LOX sphere caused failure of the outer carbon steel shell (1968).
- Failure of the roof joint of an ethylene tank from over-pressure in Montreal (1970).
- The inner tank of a 250,000 bbl LPG storage tank leaked into the annular space. The leak was detected and a spill did not occur.
- A liquid nitrogen storage tank ruptured and the rupture caused failure of a pipeline from an adjacent liquid oxygen tank.
- The inner tank of a 630,000 bbl LNG storage tank in Abu Dhabi LNG, leaked into the annular space causing a cold spot on the outer tank wall (1979). The leak was detected and the tank was drained without a spill.

steel roof may also fail from contact with the cold vapor. If a major overfill occurs, LNG will flow into the annular space between the tank walls and failure of the carbon steel outer shell will result. Failure of the outer tank shell will not necessarily cause an LNG-releasing failure of the inner tank, but it is a definite possibility.

The storage tanks at the reference import terminal each have three float-type level indicators. Two of these indicators sound high-level alarms in the control room when a pre-set maximum level is reached. Both indicators sound high-high level alarms at a somewhat higher level. If the operator still has

not stopped flow into the tank, the third level indicator activates a level switch which automatically shuts down flows into and out of the tank. The high-level alarm is set for 86 feet, or 2 feet below the top of the inner tank. If tank filling is not stopped, liquid will overflow the inner tank about 10 minutes after the high-level alarm is activated.

Storage tanks at the reference facility will be filled a total of 100 times per year. Based on generic failure data, a tank will be over-filled 2×10^{-3} times per year. A fault tree for overfilling a storage tank is shown in Figure B.7. Supporting calculations are shown in Table B.4. The critical components of this system are the level indicators and switches and the operator. Various problems with float type indicators in LNG storage tanks have been reported (Chelton 1979). If these indicators are less reliable than their counterparts in other applications, then generic failure rate data may underestimate the probability of overfilling an LNG storage tank.

If a tank is overfilled, there will be several indications (other than level indicators and alarms) that a serious problem exists (high-pressure alarm, relief valves opening, etc.) and the operator can stop flow before a serious overflow into the annular space occurs. Because the exact consequences of a serious overflow are not known, it is impossible to predict how often this would result in gross failure of the storage tank.

4.3.3 Storage Tank Overpressure

The maximum design pressure for the reference facility storage tanks is 2.0 psig. At a somewhat higher pressure (around 5-7 psig) the roof-shell joints on the outer tank will fail and the tank roof will explode outward, exposing the inner tank. A large, continuous release of vapor would result. To prevent this, the storage tanks have sophisticated pressure control and pressure relief systems.

LNG vapor is removed from the storage tanks by boiloff compressors. A portion of the vapor removed is returned to the tanks to maintain the pressure in the tanks. The main pressure controller maintains the pressure in the boiloff header by modulating the amount of gas sent to the fuel gas and pipeline compressors. Each tank has a pressure controller which maintains the absolute

pressure in the tank at 15.5 psia by controlling the amount of vapor returned to the tank. In addition, each tank has a gauge pressure control system. At 1.3 psig the gauge pressure control system opens a 20-in.-diameter vent valve. If the pressure continues to increase, six 12-in. pressure relief valves will open. Data from peakshaving plants indicate that relief or vent valves are opened approximately once per year (Welker 1979).

Some of the conditions that can result in an increase in storage tank pressure are insulation failure, excessive vapor displacement or flash vaporization during filling, rollover, pressure control system failure, liquid overfill, and a drop in barometric pressure.

Localized spot failure of the insulation are common in LNG storage tanks, occurring at a rate of about 1×10^{-1} per tank-year (Welker 1979). They result in a cold spot on the outer shell and a small increase in boiloff which can be handled easily by the pressure control system. Even large insulation failures would not increase the boiloff enough to cause severe pressure control problems. The effect of insulation failures on the outer carbon steel shell will be a more serious problem than controlling the pressure in the tank.

Failure of the pressure control system is not too serious unless the tank is being filled. If the tank is not being filled and no boiloff is removed from the tank, it will take several days for the tank to reach the vent pressure and only one vent or relief valve is necessary to handle the gas flow when the tank vents. If no vapor is removed from the tank while it is being filled, the tank will reach vent pressure in 2 to 30 minutes, depending on the filling rate, the vapor pressure of the LNG entering the tank, and the liquid level in the tank. The vent valve would be able to handle the gas flow, but if it failed, four relief valves would be required to handle the flow. A fault tree for failure of the pressure control system is included in Figure B.8. Supporting calculations are shown in Table B.5. Our analysis indicates that a tank will reach vent pressure 1×10^{-2} times per year due to failure of the pressure control system during tank filling.

Excessive vapor displacement, flash vaporization during filling, and rollover can be prevented if proper ship unloading procedures are followed. The probability of an operator failing to follow standard procedures is in the

range of 10^{-1} to 10^{-2} . This suggests the tank vent or relief valves will be opened approximately 0.5 times per year as a result of failure to follow correct tank filling procedures. The amount of gas vented and the vent rate depend on the liquid properties, the filling rate, and the conditions in the tank. The total amount of gas vented can be estimated with reasonable accuracy, but the vent rate is not easily determined.

Because the storage tanks are maintained at a constant absolute pressure, a decrease in barometric pressure will increase the gauge pressure in the tank. Should this occur, gauge pressure sensors in the tank will sound an alarm in the control room and the operator can adjust the absolute pressure controller to account for this change. Failure to do this can result in the tank pressure reaching vent pressure. This will occur about 3×10^{-2} per demand, based on operator failure rates.

To calculate the unavailability of the pressure relief valves, we used a failure rate of 1×10^{-6} /hr for faults that would prevent the relief valve from opening when required. If a test interval of 3 months is used, the probability of failure on demand is 1×10^{-3} . This is higher than failure rates used in studies of the nuclear industry (USNRC 1975) and somewhat lower than failure rates that have been used in the chemical industry (Kletz 1972).

If only one relief valve is required to handle an overpressure, the probability of the overpressure not being relieved is 1×10^{-19} (six relief valves plus one vent valve) assuming independent relief valve failures. In the case of overpressure during tank filling where either the vent valve or four relief valves are required, the probability of the overpressure not being relieved is 10^{-10} per demand, again assuming independent failures. Common-cause failures which result in several or all of the relief valves failing at once are more likely than the simultaneous random failure of several relief valves. Common-cause failures include a plugged relief header, improperly adjusted set pressure, closed relief valve block valves, and common manufacturing defects. A "rule of thumb" that has been used to predict the effectiveness of redundant components (Fussell 1980):

$$P_s = P_c^{\sqrt{n}}$$

where

P_s = the probability of redundant components failing concurrently

P_c = the probability of failure for a single component

n = number of components.

Using this rule of thumb, the probability of an overpressure not being relieved when only one relief or vent valve is required is about 5×10^{-9} . For an overpressure during tank filling, the probability the overpressure would not be relieved is 6×10^{-7} .

The amount of vapor released from the relief valve will vary depending on the situation but will be in the range of 50 to 500 MMSCFD (400-4000 gal/min equivalent). The lower number is typical of the boiloff rate during tank filling and the higher number is equal to the maximum capacity of the vent valve and all the relief valves.

The vapor-generation rate from an open-top LNG tank can be used to approximate the vapor release rate from a tank whose roof has failed from overpressure. This vapor generation has been calculated for a 240-ft-diameter tank on Staten Island (FPC 1974). The calculated value ranges from 12 to 36 lb/sec for typical ambient wind speeds. This is equal to about 200 to 600 equivalent gallons per minute. If the storage tank is full it would take about 100 hours to pump out the tank. In this time 1.2×10^6 to 3.6×10^6 equivalent gallons would be vaporized.

4.3.4 Storage Tank Underpressure

The minimum design pressure for the reference facility storage tanks is 0 psig. The tanks can withstand only about 2 inches of water external pressure before they collapse. Information obtained from personal communications with Kletz of Imperial Chemical Industries, LTD. indicates that the probability of the atmospheric pressure storage tanks failing from underpressure is 4×10^{-3} per tank-year. ICI indicated that most of these tanks do not have the safety

features included on LNG storage tanks to prevent underpressure. The storage tanks at the reference LNG import terminal have a guage pressure indicator which alarms in the control room if the tank reaches 0.3 psig. At 0.15 psig vapor is automatically admitted into the tank from the fuel gas system, and if the pressure continues to fall, three vacuum "breather" valves open at 0.03 psig to prevent the tank gauge pressure from becoming negative.

Conditions that could result in decreased tank gauge pressure include a rise in barometric pressure, failure of the pressure control system causing removal of too much vapor from the tank, and too rapid a withdrawal of liquid from the tank. A fault tree for pressure control system failure resulting in opening of the vacuum relief valves is shown in Figure B.9. Supporting calculations are shown in Table B-6. This will occur about 2×10^{-3} times per year at the reference facility. Our analysis indicates that in the event of a large increase in barometric pressure, the vacuum relief pressure will be reached at a probability of 3×10^{-3} per demand. Normal sendout to the vaporizers presents no problems with respect to reducing the tanks pressure; however, other operations such as tank-to-tank transfer or emergency tank pump-out could cause a reduction in tank pressure.

The probability of all three vacuum relief valves failing is 10^{-9} per demand, assuming independent failures. If more than one valve is required to open to stabilize the tank pressure, than the failure rate is correspondingly higher-- 10^{-6} per demand if two are required and 10^{-3} if all three are required. As with the pressure relief valves, common mode failures are more likely than simultaneous random failure of several relief valves. These include such things as a blocked header pipe, improperly adjusted valves, and closed relief valve block valves. Using the "rule of thumb" described earlier to account for common cause failures of redundant components, the probability of three vacuum relief valves failing is 6×10^{-6} and the probability of two valves failing is 6×10^{-5} .

4.3.5 Inlet and Outlet Line Leaks/Ruptures

All of the inlet and outlet lines to/from the storage tanks enter through the roof. Any significant release of LNG from these lines over the storage tank could crack the outer carbon steel roof. If one of the lines ruptures over the tank, the resulting release would likely cause failure of the tank roof. Generic

failure rate data indicate about 1×10^{-6} failure of this type will occur per tank-year. Large leaks from the inlet/outlet lines will occur approximately 1×10^{-2} times per year. These leaks may cause small cracks in the roof which could result in a gas leak from the tank. There have been several reported incidents of cracked roofs caused by small LNG leaks (Thwaites 1978).

Our analysis indicates a rupture of an inlet LNG line will occur about 2×10^{-6} times per year. If the unloading system is shut down and the release is isolated in 1 minute, 61,000 gallons would be released. If the system is not shut down for 10 minutes, 538,000 gallons will be spilled. This will occur about 4×10^{-8} times per year. Rupture of an outlet LNG line will occur about 3×10^{-5} times per year. If the sendout system is shut down and the release isolated in 1 minute, 5,000 gallons would be spilled. If the system is not shut down for 10 minutes, 41,000 gallons would be released. This will occur 2×10^{-2} per demand, resulting in a probability of about 5×10^{-7} per year for this scenario. If an inlet or outlet line ruptures, gas detectors in the storage tank dike area will warn the operator that an emergency condition exists and the operator would then have to activate the ESD. Fault trees for rupture of storage tank inlet and outlet lines are shown in Figures B.10 and B.11. Supporting calculations are given in Table B.7 and B.8.

The "riser" portion of the inlet and outlet lines are the most stressed pieces of piping in the main plant area (Contreras 1979). The risers extend over 100 feet in the air, from near ground level to the top of the tank. Failure of these risers can be caused by such things as geysering, fatigue from thermal cycling, and fluid hammer. Because of the unique stresses associated with the risers, it is possible that generic failure rate data underestimates the probability of their failure.

4.3.6 Second-Stage Sendout Pumps

A release of LNG from a rupture of a second-stage pump, the suction and discharge valves, or associate piping and fittings, will occur about 2×10^{-3} times per year. Failure modes include fatigue from pressure and temperature cycling, cavitation, and vapor lock. Cavitation and vapor lock are both common when pumping liquids near their saturation point. Both can cause the pump to rupture.

In the reference facility, the second-stage pumps are surrounded by their own dike. In the event of a release of LNG in the secondary pump area dike, low-temperature detectors would activate an alarm in the control room. If the release were large enough, the flow sensor in the sendout line might also activate an alarm (low flow). If the operator activates the ESD and it shuts down the sendout system in 1 minute, the maximum spill size would be about 5,000 gallons. If the ESD system fails, over 40,000 gallons could be released in 10 minutes. This will occur about 2×10^{-2} per demand, resulting in a probability of 3×10^{-5} per year for this scenario. If the ESD fails, the operator can manually shut down the primary or secondary pumps (trip the circuit breaker) to stop the flow of LNG. Figure B.12 shows a fault tree for a large release of LNG from the second-stage sendout pumps. Supporting calculations are given in Table B.9.

The major danger associated with the secondary pumps is that if a large spill occurred and were ignited, the resulting fire could threaten the storage tanks. In the reference facility, the secondary pumps are located less than 100 feet away from a storage tank.

4.4 LNG VAPORIZERS

Of all the systems in an LNG import facility, the vaporizers have the highest failure rates. However, the consequences of major vaporizer failure are small compared to the unloading and storage sections and should have little effect on other sections of the facility or on areas surrounding the facility.

4.4.1 Vaporizer Tube or Panel Rupture

A recent study (Welker 1979) indicated that submerged combustion vaporizer tube failures occurred at a rate of approximately 10^{-4} per hour, which is equal to about once per vaporizer-year for full-time operation. This failure data was for peakshaving plants where vaporizers are used for only short periods of time each year. Failures were more common during startup than during steady state operation, so the failure rate will probably be less at import terminals. However, even if the failure rate is reduced by a factor of 10, a tube or panel rupture will occur about 0.5 times per year. The high failure rate is inherent

in the design of the submerged combustion vaporizer, i.e., a large number of small high-pressure tubes. The failure rate for panels in an open rack vaporizer is probably similar.

Because the vaporizers operate at high pressure (1300 psig), gas or liquid will be released at a high rate from a single tube or panel rupture until the system is depressured. A flow sensor in the outlet line will sound a low flow alarm in the control room and the operator must determine what the problem is and then activate the ESD. If this is accomplished in 1 minute, only 1700 equivalent gallons of LNG will be released. If the vaporizers are not shut down, over 9,000 equivalent gallons of LNG will be released in 10 minutes. Our analysis indicates this will occur about 1×10^{-1} times per demand, resulting in a probability of 5×10^{-2} per year for this scenario. If the ESD fails, the operator can manually shut down primary pumps to stop the flow.

4.4.2 Vaporizer Control Failures

The other common failure associated with vaporizers is control failure. Such failures occur about 10^{-3} times per hour for submerged combustion and direct fired units at peakshaving plants (Welker 1979) and are more common during startup. A failure rate of 1×10^{-4} per hour might be expected for import terminals based on this data. Generic failure rate data for control system components also indicate such failures should occur at approximately this rate (1×10^{-4} /hour) for both submerged combustion and open rack vaporizers. These control failures generally result only in an unscheduled vaporizer shutdown. A control failure that reduces the amount of heat input to the vaporizer could allow cold gas to reach the outlet of the vaporizer, and if the system is not shut down before the cold gas reaches carbon steel components, a serious failure could result.

Both types of vaporizer generally have an additional temperature sensor in the gas outlet line (other than the temperature control sensor) which activates a low-temperature alarm in the control room. In addition, open rack vaporizers have a flow sensor in the sea water line which initiates a low flow alarm and a temperature sensor in the sea water outlet line which activates a

low-temperature alarm. The submerged combustion units have a temperature sensor in the bath water which activates an alarm when the water temperature is too low, as well as UV flame detectors to indicate when a burner goes out.

If the operator is required to activate the ESD in the event of a serious control failure, a failure of the carbon steel outlet line will occur approximately 8×10^{-2} times per year. If the ESD is activated automatically by the process sensors, such failures will only occur about 1×10^{-2} times per year. If the ESD is activated after the spill occurs and shuts down the system in 1 minute, 2,000 gallons will be released. If the system is not shut down for 10 minutes, 11,000 gallons will be released. In the event of ESD failure, the operator can manually shut down the primary pumps to stop the flow. Figures B-13 and B-14 show fault trees for vaporizer outlet line rupture. Supporting calculations are given in Tables B.10 and B.11.

4.4.3 Vaporizer Header or Gas Line Rupture

Based on generic failure rate data, a rupture of this type should be expected about 1×10^{-4} times per year. If the release is stopped by the ESD in 1 minute, 16,000 equivalent gallons of LNG will be released. If the system is not shutdown 88,000 equivalent gallons will be released in 10 minutes. This can be expected to occur about 10^{-6} times per year. If the ESD fails, the operator can manually shut down the primary pumps to stop the flow. If the ESD fails and if the check valve in the gas line does not close, the operator can manually close a block valve in the gas line to prevent back flow from the pipeline.

In the event of a rupture of the header or gas line, a low-pressure sensor in the gas line would activate an alarm in the control room. The operator would have to determine the cause of the alarm and then activate the ESD.

5.0 EVALUATION OF RELEASE PREVENTION ALTERNATIVES

In Figure 5.1 the results of the release scenario analysis are plotted on the basis of frequency versus the maximum release size. Two release scenarios are significantly out of line with the rest: 1) rupture of the main transfer line and 2) gross failure of the storage tank. These two scenarios also have the largest maximum spill sizes and are the only two that could seriously affect public safety. Consequently, these two scenarios are a logical starting point for an evaluation of release prevention system design alternatives.

Using the FMEA and the fault trees developed to analyze the release scenarios, we examined on a preliminary basis the affect of various design alternatives on the expected number of events and on the magnitude of the releases. The analysis was limited to the main transfer line, the storage tank, and the emergency shutdown system which affects many of the release scenarios. The following design alternatives were considered:

Emergency Shutdown System

Redundant Sensors

Automatic Activation

Main Transfer Line

Double Ply Expansion Joints

Double Wall Transfer Line

Use of Two Smaller Transfer Lines

Offshore Emergency Shutdown Alternatives.

Storage Tank

Double Roof Tanks

Cryogenic Outer Tank

Materials of Construction

Concrete Berm or Earthen Backfill

In Ground Tanks

Redundant Level, Pressure, and Temperature Sensing Systems

The following sections discuss the results of a preliminary evaluation of the design alternatives given above. Table 5.1 lists several which have the potential to reduce the probability of a large release of LNG at an import terminal.

FIGURE 5.1. Comparison of Release Size With Expected Number of Releases

TABLE 5.1. Design Alternatives for Reducing the Probability of LNG Releases at Import Terminals

<u>Preferred Alternative</u>	<u>Others</u>
Double-ply Expansion Joints	Single-ply Expansion Joints
Double Pipe Transfer Lines	Single Pipe Transfer Line
Two Small Transfer Lines	One Large Transfer Line
Prestressed Concrete or Inground Tanks	Metal Tanks
Tank Overflow Lines	--
High, Close-in, Concrete Dikes	Low Earthen Dikes
Concrete or Earthen Berms	--
Automatically Activated Emergency Shutdown Systems	Manually Activated Emergency Shutdown Systems

The fact that many of these alternatives do in fact reduce the expected number of releases or the size of the releases should not be taken as an endorsement of these designs. More information regarding the consequences of these releases, along with a more detailed technical and economic evaluation of the design alternatives, is needed before any recommendations can be made.

5.1 EMERGENCY SHUTDOWN SYSTEM (ESD) ALTERNATIVES

As indicated previously, providing for automatic activation of the ESD (as opposed to manual activation) results in a reduction in the probability of failure on demand of at least an order of magnitude.

For manual activation of the ESD, sensor and limit switch errors or failures are generally insignificant compared to operator errors, and redundant sensors would not significantly reduce the probability of failure on demand. However, for automatic activation of the ESD, sensors and limit switch errors are critical. Addition of redundant sensor systems can often reduce the probability of failure by another order of magnitude. In some of the release scenarios, two or more different types of sensors can detect the emergency condition and additional redundancy would provide little improvement (e.g., a transfer line rupture would be detected by both flow and pressure sensors).

5.2 EVALUATION OF DESIGN ALTERNATIVES FOR THE MAIN TRANSFER LINE

Of the release scenarios studied, rupture of the main transfer line was second only to gross failure of a storage tank in the magnitude of the potential release. Three alternatives which could reduce the probability of such a release or reduce its consequences are use of double-ply expansion joints, use of a double wall transfer line, and use of two smaller transfer lines as opposed to one large one. One less-rigorous design alternative that should be considered is an emergency shutdown system which does not tie in with the ship's shutdown system. Some older import terminals do not have this capability.

5.2.1 Double-Ply Expansion Joints

Expansion joints on the main transfer line at the Cove Point terminal are each made of two plies of stainless steel, each ply designed for 300 psig. The space between the plies is pressured with regulated 15 psig dry nitrogen through an orifice and is monitored by high-low pressure alarm switches. A rise in pressure indicates a fault in the inner ply; a drop in pressure, a fault in the outer ply (Osborn 1979).

Such a system would reduce the probability of major release from the transfer line at the reference import terminal from 1×10^{-3} per year to 3×10^{-5} per year, assuming the pressure sensing system would be checked every 3 months. The most likely cause of a large release would then be a rupture of one of the one hundred-fifty 42-in. -diameter pipe sections in the line. A fault tree for failure of a transfer line with double ply expansion joints is given in Figure B-15. Supporting calculations are given in Table B.12.

5.2.2 Double Pipe Transfer Line

Two different design alternatives were considered. The first consists of two concentric stainless steel pipes with the inner line containing the LNG and the annular space being purged continuously with nitrogen. The nitrogen gas would be monitored for combustible gas. Large faults in the outer line would be detected by low-pressure alarm switches monitoring nitrogen pressure in the annular. If this design were applied to the reference import terminal it would reduce the number of expected large releases from the transfer line from

1×10^{-3} per year to 2×10^{-7} per year. Failure of the expansion joints in the inner and outer lines would be the primary cause of failure. Figure B.16 is a fault tree for failure of a transfer line of this design. Supporting calculations are given in Table B.13.

The second design utilizes the annular space as the vapor return line. As with the first design, expansion joint failures would be the primary cause of transfer line releases. Because of the dual function of the transfer line, faults in the inner pipe would not be as easy to detect, and as a result the rate of failure is somewhat higher than the first design: approximately 1×10^{-6} per year. Figure B.17 gives a fault tree for this alternative. Supporting calculations are given in Table B.14.

5.2.3 Two Smaller Transfer Lines

Two smaller transfer lines (two 30-in. I.D. lines will replace one 42-in. line) can be used to provide the same unloading rate as one larger line. One line serves as the recirculation line when no ships are unloading. This type of unloading system is used at Cove Point and several export terminals. One justification for this type of system is that the components (particularly expansion joints and valves) of the smaller transfer line are more reliable, although no data exists to confirm this. If the failure rate for smaller expansion joints is one-half of that of the larger expansion joints, the expected number of failures will be equal for both system (because there are twice as many components in the system with two smaller transfer lines). However, the size of releases from the smaller transfer lines will be only about one-half the magnitude of those from a larger transfer line. In the event of a one line, the second line could be used to pump out the ruptured line, further reducing the amount of LNG released.

5.3 OFFSHORE EMERGENCY SHUTDOWN (OES) ALTERNATIVES

The OES at the reference import terminal is activated automatically in the event of a rupture by a pressure sensor in the line and can be activated manually

by the operator in response to low pressure and low flow alarms or visual recognition of a release. If the OES is not activated automatically, the probability of the release not being stopped by the OES is increased from 5×10^{-4} to 1×10^{-2} .

Some import terminals do not have the capability to shut down the ship's pumping operations. In this case, the only means to stop flow quickly in the event of a rupture is to close the transfer line block valve. If this valve is checked for operability prior to unloading each ship, such shutdown systems are not significantly less reliable than those which can shut down the ship's pumping operations. However, if this valve is checked only every 3 months, the probability that flow is not stopped by the OES in the event of a rupture increases from 5×10^{-4} to 5×10^{-2} .

5.4 EVALUATION OF DESIGN ALTERNATIVES FOR STORAGE TANKS

Several types of LNG storage tanks are in use around the world today. Each of these designs have certain advantages and disadvantages with respect to release prevention. We evaluated the various advantages and disadvantages of these designs with respect to both internal events and external events. In addition, the merits of additional instrumentation and controls that can be added to any of the tank designs were considered.

Six different types of LNG containers are briefly described below. These types of storage tanks, which are listed below, have all been used or proposed for use in commercial LNG facilities.

- A. An above-ground tank having an inner shell of 9% Ni steel or aluminum and an outer shell of carbon steel. Perlite insulation is contained between the two shells and on top of the suspended, inner ceiling.
- A-1. Similar to Type A, except the inner tank has its own self-supporting roof.
- B. An above-ground tank with two cryogenic self-supporting independent containers. The outer tank is prestressed concrete and the inner tank can be 9% nickel steel, aluminum, or prestressed concrete.
- B-1. A tank similar to Type A, but with an outer wall of 9% Ni steel or aluminum.
- C. Type B tanks or Type A tanks with a high independent concrete dike can be backfilled to the full height of the tank or dike, or a concrete berm

can be constructed around the tank. This berm can also be located between the inner and outer tank walls.

- D. An inground tank with insulation and a thin membrane of stainless steel or Invar for the walls and floor.

Types A and A-1 are always surrounded by dikes which hold the entire contents of the storage tank or tanks. This can either be a low earthen or a concrete dike located a large distance from the tank or a high concrete wall located close to the tank. Types B and B-1 already have an additional level of containment (the second cryogenic wall); however, a dike can be added to provide a third level of containment. Dikes are usually not used with Type C and D designs.

With Types A, A-1, and B-1 several possibilities exist with respect to the liquid outlet line. If the first-stage send-out pumps are located outside the tank, the outlet line must penetrate the inner tank. A shutoff valve can be located in the outlet line inside the tank. The first stage sendout pumps can be located in the storage tank, and the liquid outlet line can exit through the roof of the tank and no penetrations in the inner tank below the liquid level are required.

5.4.1 Internal Events

In the previous section, we analyzed several operating scenarios that have the potential to cause a significant release from an LNG storage tank. The scenarios were storage tank overfill, overpressure, underpressure, and inlet/outlet line ruptures. The susceptibility of tanks of various designs to these internal events is discussed below.

Storage Tank Overfill

This scenario is very serious for Type A tanks, as indicated previously. Liquid overflow onto the suspended ceiling and into the annular space between the tank walls can cause failure of the outer tank shell, the outer tank roof, and possibly the inner tank. For Type A-1 tanks, the liquid would flow out the vapor line. If a knockout pot with a liquid sensor is installed in the outlet line, flow can be stopped before serious damage to the blower results. No damage to the tank should result.

Overflowing the inner tank of a Type B, B-1, or C design will probably not cause failure of the outer tank. However, the vapor generated when the cold vapor enters the warm space above the suspended ceiling will lift the relief valves and could cause failure of the roof. A similar scenario would take place in a Type D design.

Storage Tank Overpressure and Underpressure

The relative advantages and disadvantages of the various tank designs for these scenarios are not significant. The design and maximum operating pressures for all the tank designs are similar.

Inlet and Outlet Line Leaks/Ruptures

The expected number of these failures will not vary for any of the tank designs. However, if the leak occurs over a tank with a carbon steel roof it may cause the roof to crack and release a small amount of gas. A rupture of one of these lines could cause failure of the tank roof. A roof constructed of, or overlayed with a material suitable for cryogenic service would be protected from this type of failure.

For those tanks with outlet lines which penetrate the inner tank below the liquid level, a potentially more serious problem can occur. If the outline ruptures between the tank and the first block valve, the entire content of the tank would be released. This would occur about 1×10^{-5} /yr. Most tanks of this design have a shutoff valve inside the tank to prevent release of the tank contents if this occurs. However the reliability of these valves is questionable and testing them is difficult. If they are tested once per year, the probability they will fail to close on demand is about 0.2.

5.4.2 External Events

External events such as fires, explosions, sabotage, and severe weather may be a significant cause of LNG tank failures. The susceptibility of different tank designs to various hazards, including may external events, has been studied qualitatively (Fecht et al. 1982) and some of the general conclusions are presented here.

With respect to external events, the outer tank serves as a protective shield for the storage unit. The thin metal outer tank of Type A, A-1, and B-1 designs provides only minimal protection against these events. A high, close-in dike made of prestressed concrete make these tanks less susceptible to damage from fires, explosions, penetration by missiles, and severe weather than tanks with a low earthen dikes. However, the high, close-in, prestressed concrete dike is more likely to fail in the event of an earthquake.

Type B tanks (prestressed concrete outer wall) gives significantly more protection than metal tanks against most all external events. These tanks can be build with a single steel roof overlayed with concrete to provide even more protection against fires, explosions, sabatoge, and most types of severe weather.

Tanks of type C and D provide the ultimate in protection against most external events because only the top of the tank is exposed. However, an inground tank is more susceptible to damage from some types of severe weather such as floods and hurricanes.

Of the types of tanks considered here, Type B tanks with a concrete berm and a concrete overlay roof were judged most resistant to damage from external events followed by Type C tanks with concrete roofs. Type A tanks with low earthen dikes are the most susceptible to damage from external events.

5.4.3 Storage Tank Instrumentation and Controls

Sophisticated instrumentation systems are included in the reference import terminal tanks. Adding redundant sensors usually adds little to the capability of the systems. Adding additional shutdown switches and eliminating common cause failures can provide improvement in some cases.

Level Instrumentation

Each storage tank in the reference facility has three high-level alarms. The probability of the operator failing to respond to the high-level alarms is much higher than the probability of all three alarm systems failing off. Adding additional level detectors and alarms would not reduce the probability at overfill (10^{-3} per year for the reference import terminal). One level detector activates a high-level switch which automatically shuts off flow into the tank.

Equipping one of the other level indicators with a high-level shutdown switch would reduce the probability of overfill. Because these would be similar systems, common cause failures would probably limit the improvement to an order of magnitude (10^{-4} per year compared to 10^{-3}).

Another device used to prevent overfilling in some LNG tanks is a liquid overflow line. This line routes the LNG to some containment facility in the plant or to a flare (Contreras 1979). The capacity of the containment or flare facility is usually limited, so the line is equipped with a temperature sensor to indicate LNG flow in the line. This sensor typically activates an alarm warning the operator to shut down filling operations. For increased protection, it could automatically activate the emergency shutdown system.

Use of less sophisticated-level instrumentation systems would significantly increase the probability of overfill. A tank with a one-level alarm and no automatic shutdown would be overfilled about once per year. A tank with a two-level alarm (only one indicator) and an automatic shutdown would be overfilled about 10^{-2} per year.

Pressure Instrumentation

Each storage tank at the reference facility has an atmospheric pressure indicator/alarm and a gauge pressure indicator/alarm. In addition, the storage tanks are connected by the vapor withdrawal and vapor return headers. With respect to overpressure and underpressure, the storage system can really be considered as one large tank with four indicator/alarm systems. Additional redundancy would add little to the system. Each storage tank has six pressure relief valves and three vacuum relief valves. In most cases only one or two valves will be required, so redundancy is supplied.

The major cause of failure of the pressure indicator/alarm and relief valve systems will likely be common cause failures. Mistakes in testing, maintenance, and calibration can affect all of the pressure indicators or all the relief valves and significantly increase the probability of all of them failing simultaneously. Eliminating common cause errors and assuring the independence of each redundant component is the most effective means of decreasing the probability of a serious overpressure or underpressure.

Miscellaneous Instrumentation

Instruments are available which can be placed between the inner and outer tanks to detect leaks in the inner tank. Such an instrument could be useful in tanks with carbon steel outer tanks when early detection of a leak might prevent possible structural damage to the outer tank.

In tanks with two independent roofs the annular space between the tanks is purged with nitrogen to keep moisture out of the insulation. The purge gas can be analyzed to detect leaks in the inner tank.

Linear movement indicators can be attached to the bottom ring of the outer shell to detect any relative movement between the inner and outer shell. These would be particularly useful during cooldown and heatup operations.

6.0 CONCLUSIONS AND RECOMMENDATIONS

This report has analyzed the release prevention systems of an LNG import terminal. A series of potential release scenarios were analyzed to determine the frequency of the release events, the probability these releases are not stopped or isolated by emergency shutdown systems, the estimated release quantities, and the critical components of the system. Table 4.1 summarizes this analysis.

The two plant areas identified as being most significant with respect to safety are the unloading system and the storage system. Rupture of the main transfer line and gross failure of the storage tanks are the two release scenarios of primary safety interest. Reducing the rate of failure by improved design, better maintenance and testing, or adding redundancy of the critical system components for these plant areas and release scenarios will result in improved safety.

Several design alternatives which have the potential to significantly reduce the probability of a large release of LNG occurring at an import terminal have been identified. They are listed in Table 5.1. These design alternatives would reduce the probability of a large release of LNG by reducing the expected number of failures which could cause a release or by reducing the magnitude of releases that do occur. All of these alternatives are technically feasible and have been used or considered for use in at least one LNG facility.

A more rigorous analysis of the absolute risk of LNG import terminal operation is necessary before the benefits of these design alternatives can be determined. In addition, an economic evaluation of these alternatives must be made so the costs and benefits can be compared. It is our opinion, based on our preliminary analysis, that for remotely located facilities many of these alternatives are probably not justified; however, for facilities located in highly populated areas, these alternatives deserve serious considerations.

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APPENDIX A

FAILURE MODES EFFECTS ANALYSIS

TABLE A.1. Failure Modes and Effects Analysis of the Marine Terminal and Unloading System

Marine Terminal and Unloading System Component		Failure Mode	Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
			Minor	Major			
42" transfer line 40' section	(150)	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG release	C	ESD system, quick drain system
42" expansion joints	(25)	Leak or rupture	2×10^{-6}	1×10^{-7}	LNG release	C	ESD system, quick drain system
42" transfer valves (air operated)	(2)	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD system, quick drain system
		Fails open or is misdirected toward open (fails to close)		5×10^{-5}	LNG flow is not stopped	C	Can be closed manually
		Fails closed or is misdirected towards closed		5×10^{-5}	Can not unload ship	S	
16" loading arms (including swivel joints)	(5)	Leak or rupture	3×10^{-6}	3×10^{-7}	LNG release	C	ESD system, quick drain system
Drain valves to dump tank (including vapor line)	(7)	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD system
		Fails to open or is misdirected towards open		5×10^{-5}	LNG flows to dump tank	M	
		Fails closed or is misdirected toward closed		5×10^{-5}	LNG can not flow to dump tank in emergency	C	
24" loading lines (including vapor line)	(5)	Rupture or leak	10^{-9}	10^{-9}	LNG release	C	ESD system, quick drain system
24" loading valves (air operated)	(5)	Rupture or leak	10^{-7}	10^{-3}	LNG release	C	ESD system, quick drain system
		Fails open or is misdirected towards open		5×10^{-5}	Loading arms can not be isolated	M	
		Fails closed or is misdirected towards closed		5×10^{-5}	Can not unload ship	S	

TABLE A.1. (Contd)

Marine Terminal and Unloading System		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Relief valves on transfer line and loading lines (including vapor line) (8)	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD system, quick drain system
	Premature open		1×10^{-5}	LNG release	M	
	Fails to open		1×10^{-6}	Possible failure of piping from overpressure	C	
Vapor return compressor (1)	Leak or rupture	1×10^{-7}	1×10^{-8}	Release of natural gas	C	ESD system
	Fails to operate or run properly		10^{-3}	Vapor can not be returned to the ship	C	Pressure indicator/alarms on cargo tanks, ESD system
16" vapor return line 40' section (150)	Leak or rupture	1×10^{-9}	1×10^{-10}	Release of natural gas	C	ESD system
16" expansion joints (25)	Leak or rupture	2×10^{-6}	1×10^{-7}	Release of natural gas	C	ESD system, quick drain system
16" transfer valves (air operated) (2)	Leak or rupture	1×10^{-8}	1×10^{-9}	Release of natural gas	C	ESD system
	Fails open or is misdirected towards open		5×10^{-5}	Vapor flow is not stopped	M	
	Fails closed or is misdirected towards closed		5×10^{-5}	Vapor can not be returned to ship- could cause failure of cargo tanks from underpressure	C	Pressure indicator/alarms on cargo tanks, ESD system
Instrumentation and Emergency Shutdown System						
Ship motion (4 per arm) indicators 8 total	Fail to operate		$1 \times 10^{-2}/\text{demand}$	ESD is not activated in the event of excess ship motion	C	Two sets of redundant sensing devices on two of the loading arms
Pressure sensor in transfer line (includes transducer, amplifier, indicator)	Fails dangerously		1×10^{-4}	ESD is not activated in the event of loss of flow (rupture of transfer line)	C	Operator may be warned by other detectors and activate ESD manually

TABLE A.1. (Contd)

Marine Terminal and Unloading System		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Flow sensor in transfer line (includes transducer, amplifier, indicator)	Fails dangerously	2×10^{-4}		Operator is not warned in event of loss of flow	C	Operator may be warned by other detectors
Controllers or switches Motion Pressure Flow	Fail dangerously	3×10^{-5}		See corresponding sensor above		
Alarms Motion Pressure Flow	Fail to operate	5×10^{-5}		See corresponding sensor above		
Relays	Fails to energize	1×10^{-5}		If ESD is fail safe, system will shut down unnecessarily	M	
	Normally open contact fails to close	3×10^{-7}		This is not fail safe setup; failure will cause ESD not to be activated on demand		
	Normally closed contact fails to open	1×10^{-7}		This is fail safe setup; failure will cause ESD not to be activated on demand		
Gas detectors	Fail to detect gas	4×10^{-5}		Operator is not warned of release	M	
	Not in operation	--				
UV radiation (flame) detectors	Fails to sense flame	1×10^{-4}		Operator is not warned of a fire	M	
	Not in operation	--				
Low temperature detectors	Fails to detect low temperature LNG	--		Operator is not warned of release	M	

TABLE A.1. (Contd)

Marine Terminal and Unloading System		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Controlled closure mechanisms on transfer line valves	Valve closes too fast	--		Surge in transfer line pressure from fluid hammer	C	
	Valve closes too slow or will not close	--		System is not isolated	C	
Operator	Fails to activate ESD when warned by alarm	1×10^{-2} /demand		Release is not isolated	C	System will shutdown automatically in some instances
	Fails to observe a release	1×10^{-1} /demand		Release may not be detected and isolated	C	Detectors may warn operator of release
	Fails to start vapor return compressor	3×10^{-3} /demand		Could cause failure of ships tanks from under pressure	C	
	Fails to follow standard testing and maintenance schedules for instruments and shutdown systems	--		Increases probability of ESD failure	C	
	Fails to properly connect loading arms to ship	1×10^{-2} /demand		Small release of LNG	C	Hydraulic connections

TABLE A.2. Failure Modes and Effects Analysis of the Storage Tanks and Sendout System

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Storage tank inlet lines	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG release	C	ESD, dikes
Storage tank inlet valves	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD, dikes
	Fail open or are misdirected toward open	5×10^{-5}		Tank may be overfilled	C	Loading system can be shut down to stop flow
	Fail closed or are misdirected toward closed	5×10^{-5}		Tank can not be filled	S	
Storage tank inner tank	Leaks or ruptures	----		Can eventually cause failure of outer tank and release LNG to the atmosphere	C	Dikes
Outer tank	Cracks	----		Could cause failure of entire tank	C	Dikes
Primary pumps	Fail to run properly	3×10^{-5}		Can't empty storage tanks	M	
	Leak or rupture	1×10^{-7}	1×10^{-8}	Can't empty tank	M	
	Fail to start (same as relay)	1×10^{-5}		Tank can not be emptied	S	Two pumps per tank
	Fail to stop (same as relay)	1×10^{-7}		Flow from tank is not stopped. Tank could be drained and begin to heat up. Also flow would not be stopped in an emergency condition	M	LNG can be put back into the tank to keep it cool. Block valves will stop flow out of the tank.
Storage tank outlet lines	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG release, possibly on top of the tank causing roof to crack	C	ESD, dike

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Storage tank outlet valves	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD, dike
	Fails open or is misdirected toward open	5×10^{-5}		Storage tank is not isolated in an emergency	M	ESD also shuts down primary pumps
	Fails closed or is misdirected toward closed	5×10^{-5}		Storage tank can not be emptied	M	
Sendout line to secondary pumps	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG release	C	ESD, dike
Secondary pumps	Leak or rupture	1×10^{-7}	1×10^{-8}	LNG release	C	ESD, dike
	Fails to run properly	3×10^{-5}		LNG can not be fed to vaporizers	S	
	Fails to start (same as relay)	1×10^{-5}		Tank can not be emptied	S	Spare pumps are available
	Fails to stop (same as relay)	1×10^{-7}		Flow is not stopped in an emergency	M	Primary pumps and block valves will stop flow
Secondary pump suction and discharge valves	Leak or rupture	1×10^{-7}	1×10^{-8}	LNG release	C	ESD, dike
Minimum flow line	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG release	C	ESD
Minimum flow line valves	Leak or rupture	1×10^{-7}	1×10^{-8}	LNG release	C	ESD, dike
Pressure control valve - minimum flow line	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD, dike
	Fails open or is misdirected towards open	5×10^{-5}		Reduced LNG pressure or flow to vaporizers	M	
	Fails closed or is misdirected toward open	5×10^{-5}		Increased back pressure on pumps - resulting low flow can cause vapor lock or pump failure	C	Pump over-amp trip

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Sendout pump - isolation valves	Leak or rupture	1×10^{-8}	1×10^{-9}	Release of LNG	C	ESD also stops the sendout pumps
	Fails open or are misdirected toward open	5×10^{-5}		System is not isolated when ESD is activated	M	
	Fails closed or are misdirected toward closed	5×10^{-5}		Flow to the vaporizers is reduced or stopped	S	
Storage Tank Instrumentation						
Main storage tank pressure detector (includes transducer amplifier indicator)	Fails dangerously	1×10^{-4}		Increase or decrease in storage tank pressure	C	Individual storage tank pressure controllers and vent/relief valves
Main storage tank pressure controller (Modulates speed of boiloff compressor)	Fails dangerously	3×10^{-5}		Increase or decrease in storage tank pressure	C	Individual storage tank pressure controllers and vent/relief valves
Main control valve in boiloff line to compressors	Fails closed or is misdirected toward closed	5×10^{-5}		Tank pressure will increase	M	Will take several days to reach vent pressure
	Leak or rupture	1×10^{-8}	1×10^{-9}	Release of natural gas	C	
Pressure detector in vapor return line to ship	Fails dangerously	1×10^{-4}		Higher or lower pressure gas is returned to ship	M	Ship has pressure control valve
Pressure control valve in vapor return to ship	Fails closed or is misdirected towards closed	5×10^{-5}		Tank pressure will increase	C	
High or low pressure alarms	Fails to operate	5×10^{-5}		Operator is not warned of pressure variation	C	
Individual Storage tank pressure detectors	Fails dangerously	1×10^{-4}		Vent valve will not be activated or will be activated prematurely	C	Main storage tank pressure controller relief valves

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Individual storage tank pressure controllers	Fails dangerously	3×10^{-5}		Vent valve will not be activated or will be activated prematurely		Main storage tank pressure controller relief valves
Individual storage tank vent valves	Fail closed or are misdirected towards closed	5×10^{-5}		Tank will not vent or will vent at a reduced rate	C	Main storage tank pressure controller, relief valves
	Fail open or are misdirected towards open	5×10^{-5}		Tank will vent unnecessarily	M	
Storage tank level detectors	Fail dangerously	1×10^{-4}	1×10^{-5}	Storage tank may be overfilled	C	Three level detectors per tank
High or low level alarms	Fail to operate	5×10^{-5}		Operator is not warned of dangerous levels	C	Four high or high-high level alarms per tank. Two low level alarms per tank
Storage tank temperature detectors	Fail dangerously	1×10^{-4}		Potential rollover condition might not be recognized	M	
				Cooldown could proceed too fast causing stress on the tank	M	
Pressure relief valves	Fail to open	1×10^{-6}		Potential of tank failure from overpressure	C	Each tank has six valves
	Open prematurely	1×10^{-5}		LNG vapor release	M	
Vacuum relief valves	Fail to open	1×10^{-6}		Potential of tank failure from underpressure	C	Each tank has three valves
	Open prematurely	1×10^{-5}		LNG vapor release	M	
Storage tank level switches	Fail dangerously	3×10^{-5}		High or low level alarms are not activated, EDS is not activated	C	Two detectors activate high, high-high and low level alarms. Third detectors activates ESD.

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
<u>Gas Handling Equipment</u>						
Vapor blowers	Fail to operate properly	3×10^{-4}		Pressure in storage tank increases	M	Two compressors - one serves as backups
	Will not start or stops running (same as relay)	1×10^{-5}		Vapor can not be removed from storage tanks	C	
	Will not stop (same as relay)	1×10^{-7}		None	S	Pressure control valves
Vapor return line to ship	Leak or rupture	1×10^{-7}	1×10^{-8}	Release of natural gas	C	ESD will stop flow
	Leak or rupture	1×10^{-9}	1×10^{-10}	Vapor release, failure of shipboard tanks from underpressure if unloading is not stopped	C	ESD to stop unloading operations
Boiloff pipeline to pipeline and fuel gas compressors	Leak or rupture	1×10^{-9}	1×10^{-10}	Vapor release	C	ESD
Fuel gas compressors	Leak or rupture	1×10^{-7}	1×10^{-8}	Release of natural gas	C	ESD
	Will not start or stops running (same as relay)	1×10^{-5}		Boiloff can not be compressed and must be vented or pressure in tank will rise	M	Three compressors - one is required during normal operations, two for filling operations
	Will not stop (same as relay)	1×10^{-7}		None	S	
	Fails to run normally	3×10^{-4}		Can't compress boiloff	M	Three compressors - one is required during normal operations, two for filling operations

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Pipeline compressors	Leak or rupture	1×10^{-7}	1×10^{-8}	Release of natural gas	C	ESD
	Will not start or stops running (same as relay)	1×10^{-5}		Boiloff can not be compressed to pipeline pressure - must be vented if all compressors fail	M	Three compressors - one is required during normal operations, two for filling operations
	Fails to run properly	3×10^{-4}		Boiloff can not be compressed to pipeline pressure - must be vented if all compressors fail	M	Three compressors - one is required during normal operations, two for filling operations
	Will not stop (same as relay)	1×10^{-7}		Compressors may overheat	M	Operator can disconnect circuit breaker
Pipeline compressor inter and after coolers (fir-fans)	Leak or rupture	1×10^{-8}	1×10^{-9}	Vapor release	C	ESD
	Fail to operate (electric motor, or relay)	1×10^{-5}		Hot gas can cause compressor to overheat and fail	M	Compressor have over-temperature trip
Vapor blower fuel gas and pipeline compressor trips						
Sensors Pressure Temperature Vibration	Fails dangerously	1×10^{-4}		Compressors do not shut down and can fail	M	
Limit switch Pressure Temperature Vibration	Fails dangerously	3×10^{-5}		Compressors do not shutdown and can fail	M	

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)				
Component	Failure Mode	Minor	Major	Effect on System	Class	Compensating Provisions
<u>Emergency Shutdown System</u>						
Process detectors and controllers	- see storage tank instrumentation					
Gas detectors, UV flame detectors, Low temperature detectors	see Marine Terminal/Unloading System					
Relays	- see Marine Terminal/Unloading System					
Operator(s)	Fails to activate ESD when warned by alarm	1×10^{-2} /demand		Release is not isolated	C	System will shutdown automatically in some instances
	Fails to observe release	1×10^{-1} /demand		Release may not be detected and isolated	C	Detectors may warn operator of release
	Fails to follow standard maintenance and testing schedules for instruments and EDS systems	--		Increases probability of ESD failure	C	Assumptions: Normally inactive components - detectors, switches, relief valves, isolation valves, etc., checked every three months. Failure of normally active components - process instrumentation, control valves, pumps, etc., - will be detected immediately. Alarms checked once per week.

TABLE A.2. (Contd)

Storage Tank and Sendout		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Operator(s) (contd)	Fails to respond to high level in storage tanks	3×10^{-2}		Can result in overfilling storage tanks	C	High-high level automatically activates ESD
	Fails to respond to high or low pressure in storage tanks	3×10^{-2}		Can result in overpressuring or underpressuring storage tanks	C	Relief valves protect tanks
Individual storage tank control valves	Fail closed or are misdirected towards closed	5×10^{-5}		Tank pressure will decrease	C	Main storage tank pressure controller, relief valves
	Fail open or are misdirected towards open	5×10^{-5}		Tank pressure will increase	C	Main storage tank pressure controller, relief valves

TABLE A.3. Failure Modes and Effects Analysis of the Open-Rack Seawater Vaporizers

Open Rack Vaporizers		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
24-in LNG feed line from secondary pumps - 9 40-ft sections	Leak or rupture	1×10^{-6}	1×10^{-10}	LNG release	C	ESD system, dikes
24-in Isolation valve	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD system, dikes
	Fails open or is misdirected towards open	5×10^{-5}		LNG flow is not stopped ESD time may be increased	M	ESD also shutdown secondary pumps
	Fails closed or is misdirected towards closed	5×10^{-5}		LNG flow to vaporizers is stopped	S	
10-in LNG feed line to each seawater vaporizer - 5 each, 10-ft long	Leak or rupture	10^{-9}	10^{-10}	LNG release	C	ESD system, dikes
Relief valve in 10-in feed lines - 5 each	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD system, dikes
	Premature open	1×10^{-5}		LNG vented to storage tank	M	
	Fails to open	1×10^{-6}		Possible failure of piping from overpressure	C	ESD system, dikes
10-in Flow control valve in seawater vaporizer feed line - 5 each	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG release	C	ESD system, dikes
	Fails open or is misdirected toward open	5×10^{-5}		LNG flow rate to vaporizer is high	M	
	Fails closed or is misdirected toward closed	5×10^{-5}		LNG flow rate to vaporizers is low	S	
24-in Header to seawater vaporizers 4 sections	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG release	C	ESD system, dikes

TABLE A.3. (Contd)

Open Rack Vaporizers		Failure Rate (events/yr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
10-in Header to vaporizer 7 sections	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG release	C	ESD system, dikes
10-in Check valve in vaporizer feed line	Leak or rupture	1×10^{-8}	1×10^{-9}	LHG release	C	ESD system, dikes
	Allows backflow				M	
4-in LHG feed line to each vaporizer section - 30 each	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG release	C	ESD, dikes
4-in Block valve in feed line to each vaporizer section - 30 each	Leak or rupture	1×10^{-8}	1×10^{-9}	LHG release	C	ESD, dikes
4-in Expansion bellows in each feed line - 30 each	Leak or rupture	3×10^{-6}	1×10^{-7}	LHG release	C	ESD, dikes
4-in LHG inlet header to each vaporizer section - 30 each	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG release	C	ESD, dikes
Vaporizer heat exchanger panels - 180 per vaporizer	Leak or rupture	$1 \times 10^{-4*}$	$1 \times 10^{-5*}$	LHG release	C	ESD
		*Vaporizer-hr				
4-in Vapor outlet header for each vaporizer section - 30 each	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG vapor release	C	ESD
4-in Vapor outlet line from each section - 30 each	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG vapor release	C	ESD
4-in Block valve on vapor outlet lines - 30 each	Leak or rupture	1×10^{-8}	1×10^{-9}	LHG vapor release	C	ESD
12-in Vapor header from each vaporizer - 5 each	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG vapor release	C	ESD

TABLE A.3. (Contd)

Open Rack Vaporizers		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Flow Sensor and controller in 12-in vapor outlet line - 5 each	1. Sensor/transmitter failure	2×10^{-4}		Flow controller does not function properly. High or low flow alarms are not activated	C	Operator must recognize situation and control flow manually ESD
	2. Controller failure	3×10^{-5}				
Temperature sensor and controller in 12-in vapor outlet line - 5 each	1. Sensor/transmitter failure	1×10^{-4}		Temperature controller does not modulate seawater flow properly. High and low temperature alarms are not activated	C	Operator must recognize situation and control flow manually
	2. Controller failure	3×10^{-5}				
12-in Block valve on vapor outlet line from each vaporizer - 5 each	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG vapor release	C	ESD
12-in Check valve on vapor outlet line from each vaporizer - 5 each	Leak or rupture	1×10^{-7}	1×10^{-8}	LNG Vapor release	C	ESD
	Allows backflow			Will not normally cause release as block valves are used for shutoff	M	
Relief valve on 12-in vapor outlet line - 5 each	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG vapor release	C	ESD
	Premature open	1×10^{-5}		LNG vapor vented to vent stack header	M	
	Fails to open	1×10^{-6}		Possible failure of piping from overpressure	C	
30-in Vapor header to all five vaporizers 4 sections	Leak or rupture	1×10^{-9}	1×10^{-10}	LNG vapor release	C	ESD
30-in Isolation valve	Leak or rupture	1×10^{-8}	1×10^{-9}	LNG vapor release	C	ESD
	Fails or is misdirected toward open	5×10^{-5}		Vapor flow is not stopped	M	
	Fails closed or is misdirected towards closed	5×10^{-5}		Vaporizer can not be operated LNG vapor flow is stopped	S S	

TABLE A.3. (Contd)

Open Rack Vaporizers		Failure Rate (events/hr)		Effect on System	Class	Compensating Provisions
Component	Failure Mode	Minor	Major			
Seawater pumps at supply basin - 3 each, 2 required for full flow	Fails to run properly	3×10^{-5}		Seawater flow is reduced or stopped	M	ESD
9-ft Seawater supply line - 40 sections	Leak or rupture	1×10^{-9}	1×10^{-10}	Seawater flow is stopped	M	ESD
48-ft Feed line to each vaporizer - 5 each	Leak or rupture	1×10^{-9}	1×10^{-10}	Seawater flow to all vaporizers is stopped	M	ESD
Flow sensor in 48-ft feed line	1. Sensor/transmitter failure	2×10^{-4}		In event of loss of seawater flow, ESD is not activated. High or low flow alarms are not activated	C	
	2. Controller failure	3×10^{-5}				
Control valve in 48-in water feed line	Fails closed or is misdirected towards closed	5×10^{-5}		Seawater flow is insufficient. Exit gas may be too cold	C	
	Fails open or is misdirected towards open	5×10^{-5}				
48-in Seawater header - 5 each	Leak or rupture	1×10^{-9}	1×10^{-10}	Seawater flow to vaporizer is stopped	M	ESD
20-in Seawater inlet lines - 30 each	Leak or rupture	1×10^{-9}	1×10^{-10}	Seawater flow to vaporizer is stopped	M	ESD
20-in Block valves in seawater lines - 30 each	Leak or rupture	1×10^{-8}	1×10^{-9}	Seawater flow to vaporize is stopped	M	ESD
20-in Header for each vaporizer section - 30 each	Leak or rupture	1×10^{-9}	1×10^{-10}	LHG flow is vaporizer section is stopped	M	Operator can isolate each section of the vaporizer
8-ft Seawater header for each vaporizer panel	Leak or rupture	1×10^{-9}	1×10^{-10}	Bad flow distribution on heat exchange panel	M	Operator can isolate each section of the vaporizer

TABLE A.3. (Contd)

Open Rack Vaporizers		Failure Rate (events/hr)				
Component	Failure Mode	Minor	Major	Effect on System	Class Compensating Provisions	
Operator	Fails to observe a release	1 x 10 ⁻¹		Release continues	C	Vaporizer automatic controls and ESD
	Fails to activate ESD when warned by an alarm	1 x 10 ⁻²		Emergency continues	C	
	Fails to respond correctly to vaporizer process alarm (temperature, flow, etc.)	3 x 10 ⁻²		Process upset continues, could lead to a release	C	
	Fails to respond quickly and correctly to changes in important vaporizer operating variables such as loss of LNG flow due to panel rupture, seawater flow reduction, etc.	1 x 10 ⁻¹		Could lead to a release or could increase the size of a release	C	

APPENDIX B

FAULT TREES AND FAULT TREE CALCULATIONS

FIGURE B.1 Fault Tree for Failure of Emergency Shutdown Systems

FIGURE B.2. Fault Tree Failure of a Fully Automatic Emergency Shutdown

FIGURE B.3. Fault Tree for Failure of an Automatic or Manual Detection, Manual Activation Emergency Shutdown

FIGURE B.4. Fault Tree for Large Release from the Loading Arms

FIGURE B.5. Fault Tree for Large Release from the LNG Transfer Line

FIGURE B.6. Fault Tree for Large Release from an LNG Storage Tank

FIGURE B.7. Fault Tree for Storage Tank is Overfilled

FIGURE B.8. Fault Tree for Storage Tank Reaches Vent Pressure

FIGURE B.9. Fault Tree for Storage Tank Reaches Vacuum Relief Pressure

FIGURE B.10. Fault Tree for Large Release from Storage Tank Inlet Lines

FIGURE B.11. Fault Tree for Large Release from Storage Tank Outlet Lines

FIGURE B.12. Fault Tree for Large Release from Second Stage Sendout Pumps

FIGURE B.13. Fault Tree for Seawater Vaporizer Outlet Line Ruptures

FIGURE B.14. Fault Tree for Submerged Combustion Vaporizer Outlet Line Ruptures

FIGURE B.15. Fault Tree for Large Release from an LNG Transfer Line with Double Ply Expansion Joints

FIGURE B.16. Fault Tree for Large Release of LNG from a Double Pipe Transfer Line

FIGURE B.17. Fault Tree for Large Release of LNG from a Double Pipe Transfer Line (outer pipe used as vapor return line)

TABLE B.1. Supporting Calculations for Emergency Shutdown Fault Trees

<u>Offshore Emergency Shutdown</u>	λ	T^*	$\bar{a}$
Ship's pumps do not stop (5)	5×10^{-7}	15	8×10^{-6}
Ship's block valve does not close	5×10^{-5}	15	8×10^{-4}
Transfer line block valve does not close	5×10^{-5}	360	2×10^{-2}
			1×10^{-10}

*Pumps were stopped and block valves closed prior to unloading. Faults are unannounced during unloading.

	λ	τ	$\bar{a}$
ESD circuitry fails	1×10^{-7}	360	4×10^{-5}
OES equipment fails			4×10^{-5}

<u>Master Emergency Shutdown</u>	λ	T	$\bar{a}$
Ship's pumps do not stop (5)	5×10^{-7}	15	8×10^{-6}
Ship's block valve does not close	5×10^{-5}	15	8×10^{-4}
Transfer line block valve does not close	5×10^{-5}	15	8×10^{-4}
Filling valve does not close	5×10^{-5}	15	8×10^{-4}
			4×10^{-15}

	λ	τ	$\bar{a}$
ESD circuitry fails	1×10^{-7}	1000	1×10^{-4}
MES equipment fails			1×10^{-4}

<u>Vaporizer Emergency Shutdown</u>	λ	τ	$\bar{a}$
Primary pumps do not stop (2)	2×10^{-7}	100	2×10^{-5}
Storage tank outlet valve does not close	5×10^{-5}	100	5×10^{-3}
Vaporizer block valve does not close	5×10^{-5}	1000	5×10^{-2}
			5×10^{-9}

	λ	τ	$\bar{a}$
ESD circuitry failure	1×10^{-7}	1000	1×10^{-4}
VES equipment failure			1×10^{-4}

TABLE B.2. Supporting Calculations for LNG Loading Arm Fault Tree

Large Release From LNG Loading Arms

<u>Gas Detector System</u>	λ	τ	$\bar{a}$
Gas detector fails	4×10^{-5}	1000	4×10^{-2}
Control module fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			7.4×10^{-2}

Low Temperature Detector System

Temperature detector fails	1×10^{-4}	1000	1×10^{-1}
Control module fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			1.3×10^{-1}

ESD System

ESD Circuitry fails	1×10^{-7}	360	4×10^{-5}
Ship's pumps do not stop (5)	5×10^{-7}	15	8×10^{-6}
Ship's block valve fails to close	5×10^{-5}	15	8×10^{-4}
			6×10^{-9}
ESD fails to stop release			4×10^{-5}

Expected number of failures	1.4×10^{-6} (1500)	=	$2 \times 10^{-3}/\text{yr}$
Release is not isolated	$1 \times 10^{-2}/\text{demand}$		$(2 \times 10^{-5}/\text{yr})$

TABLE B.3. Supporting Calculations for LNG Transfer Line Fault Tree

Large Release From LNG Transfer Line

	λ
1 Block valve ruptures (2)	2×10^{-9}
2 Pipe section ruptures (150)	1.5×10^{-8}
3 Expansion joint ruptures (25)	2.5×10^{-6}
4 Drain valve ruptures (2)	2×10^{-9}
	2.5×10^{-6}

Pressure Detector System

	λ	τ	$\bar{a}$
5 Pressure detector fails	1×10^{-4}	7	7×10^{-4}
6 Limit switch fails	3×10^{-5}	1000	3×10^{-2}
7 Alarm fails	5×10^{-5}	80	4×10^{-7}
			3.4×10^{-2}

Flow Detector System

8 Flow detector fails	2×10^{-4}	7	1.4×10^{-3}
9 Limit switch fails	3×10^{-5}	1000	3×10^{-2}
10 Alarm fails	5×10^{-5}	80	4×10^{-3}
			3.4×10^{-2}

ESD System

16 ESP circuitry fails	1×10^{-7}	360	4×10^{-5}
15 Ship's pumps do not stop (5)	5×10^{-7}	15	8×10^{-6}
14 Ship's block valve fails to close	5×10^{-5}	15	8×10^{-4}
13 Transfer line block valve fails to close	5×10^{-5}	360	2×10^{-2}
			1.3×10^{-10}

Critical Cut Sets

3,(5,6),12	$7.5 \times 10^{-10}/\text{hr}$ ($1.1 \times 10^{-6}/\text{yr}$)
3,(5,6),(8,9,10),11	$2.3 \times 10^{-10}/\text{hr}$ ($.3 \times 10^{-6}/\text{yr}$)
	$1.4 \times 10^{-6}/\text{yr}$

Large release occurs	$2.5 \times 10^{-6}/\text{hr}$ (1500)	=	$4 \times 10^{-3}/\text{yr}$
Release is not isolated	$4 \times 10^{-4}/\text{demand}$		($1.4 \times 10^{-6}/\text{yr}$)

TABLE B.4. Supporting Calculations for LNG Storage Tank Overfill Fault Tree

Storage Tank is Overfilled (faults per fill)

Level Alarm System (Level indicators continuously monitor tank level. A level indicator failure will not be inherently obvious to the operator).

	λ	τ	$\bar{a}$
Level detector/indicator reads low	2×10^{-4}	50	1×10^{-2}
Level switch fails ⁽¹⁾	3×10^{-5}	1000	3×10^{-2}
Level alarm fails ⁽²⁾	5×10^{-5}	80	4×10^{-3}

Automatic Shutdown

Level detector/indicator reads low	2×10^{-4}	50	1×10^{-2}
Level switch fails ⁽¹⁾	3×10^{-5}	1000	3×10^{-2}
Filling valve fails to close	5×10^{-5}	7	4×10^{-4}

Critical Minimal Cut Sets

1,2,13	$(1 \times 10^{-2}) (1 \times 10^{-2}) (3 \times 10^{-2})$	$= 3 \times 10^{-6}$
3,11,13	$(3 \times 10^{-2}) (1 \times 10^{-2}) (3 \times 10^{-2})$	$= 9 \times 10^{-6}$
3,11,12	$(3 \times 10^{-2}) (1 \times 10^{-2}) (1 \times 10^{-2})$	$= 3 \times 10^{-6}$
1,2,12	$(1 \times 10^{-2}) (1 \times 10^{-2}) (1 \times 10^{-2})$	$= 1 \times 10^{-6}$
3,4,7,13	$(3 \times 10^{-2}) (3 \times 10^{-2}) (3 \times 10^{-2}) (3 \times 10^{-2})$	$= 8 \times 10^{-7}$
		$1.7 \times 10^{-5}/\text{fill}$
	$\times 100 \text{ fills/yr}$	$\sim 2 \times 10^{-3}/\text{yr}$

(1) Level switch failures will be detected only by testing or if a high level occurs. A combined detection and repair time of 1000 hrs was used.

(2) Alarms checked weekly

TABLE B.5. Supporting Calculations for LNG Storage Tank Reaches Vent Pressures Fault Tree

Storage Tank Reaches Vent Pressure

b) Pressure Alarm System

	λ	τ	$\bar{a}$
Pressure detector/indicator reads low	1×10^{-4}	30	3×10^{-3}
Pressure switch fails ⁽¹⁾	3×10^{-5}	1000	3×10^{-2}
Pressure alarm fails ⁽²⁾	5×10^{-5}	80	4×10^{-3}
			$\sim 4 \times 10^{-2}$

a) Pressure Control System

Pressure detector/indicator reads low	1×10^{-4}
Pressure controller controls high	3×10^{-5}
Control valve fails toward closed	5×10^{-5}
	$\sim 2 \times 10^{-4}$

Boiloff Compressor System

(2 of 4 compressors must operate -	3×10^{-4}	50	1.5×10^{-2}
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Cut Sets

	$\bar{a}_K$	λ_K
1,2,3	3.4×10^{-6}	2×10^{-7}
2,3,4	3.4×10^{-6}	2×10^{-7}
1,2,4	3.4×10^{-6}	2×10^{-7}
1,3,4	3.4×10^{-6}	2×10^{-7}

$$\lambda_K = \bar{a}_K \sum \frac{\lambda_i}{\bar{a}_i}$$

$$\lambda_T = \sum \lambda_K = 8 \times 10^{-7}$$

(1) Pressure switch failure will not be detected until testing or until a high pressure occurs. A combination deflection time plus repair time of 1000 hrs was used.

(2) Alarms checked weekly

TABLE B.6. Supporting Calculations for LNG Storage Tank Reaches Vacuum Relief Pressure Fault Tree

LNG Storage Tank Reaches Vacuum Relief Pressure

a) Pressure Alarm Systems	λ	τ	$\bar{a}$
Pressure detector/indicator reads high	1×10^{-4}	30	3×10^{-3}
Pressure switch fails ⁽¹⁾	3×10^{-5}	1000	3×10^{-2}
Pressure alarm fails ⁽²⁾	5×10^{-5}	80	4×10^{-3}
			$\sim 4 \times 10^{-2}$

b) Pressure Control System

Pressure detector/indicator reads high	1×10^{-4}
Pressure controller controls low	3×10^{-5}
Control valve fails toward open	5×10^{-5}
	$\sim 2 \times 10^{-4}$

c) Emergency Gas Repressurization System

Pressure detector/indicator reads high	1×10^{-4}	30	3×10^{-3}
Pressure switch fails ⁽¹⁾	3×10^{-5}	1000	3×10^{-2}
Valve does not open	5×10^{-5}	1000	5×10^{-2}
			$\sim 8 \times 10^{-2}$

Critical Minimal Cut Sets

1,4,11	$(1 \times 10^{-4}) (1 \times 10^{-2}) (8 \times 10^{-2}) = 8 \times 10^{-8}$
1,2,3,11	$(1 \times 10^{-4}) (2 \times 10^{-3}) (8 \times 10^{-2}) = 1.6 \times 10^{-8}$
5,10,11	$(2 \times 10^{-4}) (1 \times 10^{-2}) (8 \times 10^{-2}) = 1.6 \times 10^{-7}$
	$2.6 \times 10^{-7}/\text{hr}$
	$2.3 \times 10^{-3}/\text{year}$

(1) Pressure switch failures will not be detected until testing or until a low pressure occurs. A combination detection and repair time of 1000 hrs was used.

(2) Alarms checked weekly

TABLE B.7. Supporting Calculations for Large Release from Storage Tank Inlet Lines Fault Tree

Large Release From Storage Tank Inlet Lines

Gas Detector System

	λ	τ	$\bar{a}$
Gas detector fails	4×10^{-5}	1000	4×10^{-2}
Control module fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			7.4×10^{-2}

ESD System

ESD circuitry fails	1×10^{-7}	360	4×10^{-5}
Block valve fails to close	5×10^{-5}	360	2×10^{-2}
Ship's block valve fails to close	5×10^{-5}	15	8×10^{-4}
Ship's pumps do not stop (5)	5×10^{-7}	15	8×10^{-6}
			3×10^{-10}

ESD system fails to stop release			4×10^{-5}
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Expected number of failures	$1.6 \times 10^{-9}/\text{hr}$	$(1500 \text{ hrs/yr}) = 2.4 \times 10^{-6}/\text{yr}$
Release is not isolated	$1.7 \times 10^{-2}/\text{demand}$	$(4 \times 10^{-8}/\text{yr})$

TABLE B.8. Supporting Calculations for Large Release from Storage
Tank Outlet Lines Fault Tree

Large Release From Storage Tank Outlet Lines

Gas Detector System

	λ	τ	$\bar{a}$
Gas detector fails	4×10^{-5}	1000	4×10^{-2}
Control module fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			7.4×10^{-2}

ESD System

ESD circuitry fails	1×10^{-7}	1000	1×10^{-4}
Primary pumps do not stop (2)	2×10^{-7}	100	2×10^{-5}
Block valve fails to close	5×10^{-5}	1000	5×10^{-2}
			1×10^{-6}
ESD system fails to stop release			1×10^{-4}

Expected number of failures	$3.5 \times 10^{-9}/\text{hr} = 3.1 \times 10^{-5}/\text{yr}$
Release is not isolated	$1.7 \times 10^{-2}/\text{demand} \quad (5 \times 10^{-7}/\text{yr})$

TABLE B.9. Supporting Calculations for Large Release from Secondary Pumps Fault Tree

Large Release From Secondary Pumps

Low Temperature Detector System

	λ	τ	$\bar{a}$
Temperature detector fails	1×10^{-4}	1000	1×10^{-1}
Control module fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			1.3×10^{-1}

ESD System

ESD circuitry fails	1×10^{-7}	1000	1×10^{-4}
Primary pumps do not stop (2)	2×10^{-7}	100	2×10^{-5}
Block valve fails to close	5×10^{-5}	1000	5×10^{-2}
			1×10^{-6}

Expected number of failures	$1.7 \times 10^{-7}/\text{yr}$	$(1.5 \times 10^{-3}/\text{yr})$
Release is not isolated	$2.3 \times 10^{-2}/\text{demand}$	$(3.4 \times 10^{-5}/\text{yr})$

TABLE B.10. Supporting Calculations for Seawater Vaporizer Outlet
Line Rupture Fault Tree

Seawater Vaporizer Outlet Lines Ruptures

Vaporizer Control System

	λ	τ	$\bar{a}$
Temperature detector/indicator fails dangerously	1×10^{-4}		
Controller fails dangerously	3×10^{-5}		
Control valve fails toward closed	5×10^{-5}		
	1.8×10^{-4}		

Temperature Alarm System

Temperature detector fails dangerously	1×10^{-4}	30	3×10^{-3}
Limit switch fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			4×10^{-2}

Flow Alarm System

Flow detector fails dangerously	2×10^{-4}	30	6×10^{-3}
Limit switch fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
			4×10^{-2}

Operator fails to respond to alarms			1×10^{-2}
ESD equipment fails			1×10^{-4}

Expected number of failures = $1.8 \times 10^{-4}/\text{yr}$ (8760) (1×10^{-2})
 = .016/vaporizer
 x 5 vaporizers = .08/yr

TABLE B.11. Supporting Calculations for Submerged Combustion Vaporizer Outlet Line Rupture Fault Tree

Submerged Combustion Vaporizer Outlet Line Ruptures

<u>Vaporizer Control System</u>	λ	τ	$\bar{a}$
1 Gas temperature detector/indicator fails dangerously	1×10^{-4}		
2 Controller fails dangerously	3×10^{-5}		
3 Control valve fails toward closed	5×10^{-5}		
(A)	1.8×10^{-4}		
<u>Backup Gas Temperature Switch</u>			
7 Gas temperature detector/indicator fails dangerously	1×10^{-4}	30	3×10^{-3}
8 Limit switch fails	3×10^{-5}	1000	3×10^{-2}
(B)			3.3×10^{-2}
<u>Bath Temperature Alarm System</u>			
4 Bath temperature fails dangerously	1×10^{-4}	30	3×10^{-3}
5 Limit switch fails	3×10^{-5}	1000	3×10^{-2}
6 Alarm fails	5×10^{-5}	80	4×10^{-3}
(C)			3.7×10^{-2}
9 Operator fails to respond to alarm			1×10^{-2}
10 ESD equipment fails			1×10^{-4}

Critical Minimal Cut Sets

$$\begin{aligned}
 A,9,8 & \quad 1.8 \times 10^{-4} (1 \times 10^{-2}) (3.3 \times 10^{-2}) = 5.9 \times 10^{-8} \\
 A,10 & \quad 1.8 \times 10^{-4} (1 \times 10^{-4}) = 1.8 \times 10^{-8} \\
 A,C,8 & \quad 1.8 \times 10^{-4} (3.7 \times 10^{-2}) (3.3 \times 10^{-2}) = \underline{2.2 \times 10^{-7}} \\
 & \quad 3.0 \times 10^{-7}/\text{hr}
 \end{aligned}$$

Submerged combustion vaporizers operate 800 hrs/year

Expected number of failures $3 \times 10^{-7} (800) = 2.4 \times 10^{-4}/\text{year}$

TABLE B.12: Supporting Calculations for Main Transfer Line Rupture
(Double Ply Expansion Joints) Fault Tree

Rupture of Main Transfer Line During Filling
(Double Ply Expansion Joints)

1. Assume if one ply is faulty unloading operations are stopped and do not resume until repaired so $\tau = 0$
2. If not detected by the pressure sensor the fault is not repairable
3. Longest the fault will go undected is equal to the pressure sensor test interval so $\tau_{avg} = 1000$ hrs (3 month test interval)

	λ	τ	$\bar{a}$
Expansion joint 1 fails	1×10^{-7}		
Pressure sensor fails	1×10^{-4}	1000	0.1
Pressure switch fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	$\frac{4 \times 10^{-3}}{.13}$

Expansion joint 1 fails and fault is undetected	1.4×10^{-8}	1000	1.4×10^{-5}
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Expansion joint 2 fails during loading 1×10^{-7}

$$\begin{aligned} enf &= \lambda T = 1 \times 10^{-7} (1500 \text{ hrs/yr}) \\ enf &= 1.5 \times 10^{-4}/\text{yr} \end{aligned}$$

Expected number of failures during loading per expansion joint $= 1.4 \times 10^{-5} (1.5 \times 10^{-4}) = 2 \times 10^{-9}$
x 25 expansion joints $= 5 \times 10^{-8}/\text{year}$

Other failure modes (pipe and valve rupture) $= \underline{3 \times 10^{-5}/\text{yr}}$

TABLE B.13. Supporting Calculations for Main Transfer Line Rupture
(Double Wall Pipe) Fault Tree

Rupture of Main Transfer Line During Filling
(Double Wall Pipe)

1. If fault develops in either line the system is shutdown
2. The longest a fault will go undetected is equal to the sensor test interval so τ avg = 1000 hrs (3 month test interval)

	λ	τ	$\bar{a}$
Inner pipe component fails (expansion joint)	2.5×10^{-6}		
Gas detector fails	4×10^{-5}	1000	4×10^{-2}
Limit switch fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
Operator fails to shutdown			$\frac{1 \times 10^{-2}}{8.4 \times 10^{-2}}$
Inner pipe component fails and is not detected	2.1×10^{-7}	1000	2.1×10^{-7}
Outer pipe component fails	2.5×10^{-6}		
Pressure sensor fails	1×10^{-4}	1000	0.1
Limit switch fails	3×10^{-5}	1000	3×10^{-2}
Alarm fails	5×10^{-5}	80	4×10^{-3}
Operator fails to shutdown			$\frac{1 \times 10^{-2}}{.14}$
Outer pipe component fails and is not detected	3.5×10^{-7}	1000	3.5×10^{-4}

$$\begin{aligned}
 \text{enf}_K &= \int \text{rof}_K dT = a_K \sum \frac{\lambda_i}{\bar{a}_i} dT = 2.1 \times 10^{-4} (3.5 \times 10^{-4}) \frac{2.1 \times 10^{-7}}{2.1 \times 10^{-4}} \frac{3.5 \times 10^{-7}}{3.5 \times 10^{-4}} \\
 &= \int_0^{1500 \text{ hrs}} 7.4 \times 10^{-8} (2 \times 10^{-3}) dT = 2.2 \times 10^{-7} / \text{yr}
 \end{aligned}$$

TABLE B.14. Supporting Calculations for Main Transfer Line Rupture
(double wall pipe-outer pipe is vapor return)

Rupture of Main Transfer Line During Filling
(Double wall pipe - outer pipe used as vapor return line)

	λ	τ	$\bar{a}$
Outer pipe components fails	2.5×10^{-6}	0	0
Inner pipe component fails	2.5×10^{-6}	100	2.5×10^{-4}

$$\begin{aligned} \text{Expected number of failures} &= 2.5 \times 10^{-6}/\text{hr} \quad (2.5 \times 10^{-4}) \quad (1500 \text{ hrs/yr}) \\ &\approx 1 \times 10^{-6}/\text{yr} \end{aligned}$$

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16. Abstract The purpose of this study is to provide an analysis of release prevention systems at LNG import terminals. An overview study of LNG import terminals which preceeded this effort identified 11 release scenarios which are typical of the potential hazards involved in the operation of LNG import terminals. These scenarios formed the basis for this more detailed study. Failure modes and effects analysis and fault tree analysis were used to estimate the expected frequency of each release scenario for a reference import terminal. In addition, the effectiveness of release prevention, release detection, and release control systems were evaluated.					
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Summary

Introduction

Analysis Approach

Release Scenario Analysis

Evaluation of Release Prevention Alternatives

Conclusions and Recommendations

References

Appendix A - Failure Modes Effects Analysis

Appendix B - Fault Trees and Fault Tree Calculations