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Prepared for the U.S. Department of Energy
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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 and 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 peakshaving facility release prevention systems benefited from the technical direction and guidance provided by Dr. John M. Cece and Dr. Henry F. Walter of ESED.

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

1. Assessment of Research and Development (R&D) Needs in LPG Safety and Environmental Control (PNL-3991)
2. Assessment of Research and Development (R&D) Needs in Ammonia Safety and Environmental Control (PNL-4006)
3. An Overview Study of LNG Release Prevention and Control Systems (PNL-4014)
4. Analysis of LNG Import Terminal Release Prevention Systems (PNL-4152)

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:

- Storage Tank Analysis
- Fire and Vapor Control Assessment
- Human Factors in LNG Operations

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This detailed analysis of a representative LNG peakshaving facility benefited from several plant visits undertaken by the authors and other members of the project team and many communications with representatives of the natural gas industry. The authors want to acknowledge and thank all who hosted these visits and responded to inquires during the course of this study. This report contains the authors' interpretation of all such information received.

The authors also gratefully acknowledge the assistance provided by individuals at Pacific Northwest Laboratory (PNL) during the course of this study and preparation of the final report. J. G. DeSteese, Manager of the Liquefied Gaseous Fuels Safety Studies Project at PNL, provided support and guidance throughout the study. The editorial review of the final report prior to publication was contributed by Karen Yasutake and Cary Counts. Finally, Marlene Hale and Annette Taylor 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 for a reference LNG peakshaving facility. An overview assessment of the reference peakshaving facility (Pelto 1982), which preceeded this effort, identified 14 release scenarios which are typical of the potential hazards involved in the operation of LNG peakshaving facilities. 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 the reference peakshaving facility. In addition, the effectiveness of release prevention, release detection, and release control systems were evaluated.

The reference LNG peakshaving facility consists of five basic areas: gas treatment, liquefaction, storage, sendout/vaporization and shipping and receiving of LNG from a truck terminal. Of these five areas, the storage, vaporization, and transportation and transfer areas were determined to be the most significant to plant safety. Major failures in the storage area are estimated to occur between 10^{-5} and 10^{-6} times per year and releases of 10^6 to over 10^7 gallons (the entire contents of the tank) of LNG are possible. Between 10^{-1} and 10^{-3} major failures per year are estimated to occur in the vaporization area and releases of 10^3 to 10^4 gallons are possible. About 10^{-2} to 10^{-4} major failures per year can be expected from the transportation and transfer area with the maximum release being 10^4.

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 peakshaving facility 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 ESD is 10^{-1} to 10^{-4} per demand, depending on the particular system.

Design alternatives that could reduce the expected number of occurrences or reduce the size of the releases were identified for the storage area, the ESD system, and the transportation and transfer area. 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 a peakshaving facility. A more detailed technical and economic evaluation of these alternatives is necessary before the cost and benefits of the various alternatives can be compared. It is our opinion, based on our preliminary analysis, that for remotely located facilities many of these alternatives are not justified; 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 has been selected to accomplish the study objectives. A reference description of each type of LNG facility is developed. This system description is 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 are investigated and a decision is 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, analytical and experimental studies are recommended to fill information gaps.

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 discussion 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 peakshaving facility. The report first briefly summarizes the peakshaving facility 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 areas.

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 Peakshaving Facility. 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 SYSTEM DESCRIPTION

The peakshaving facility analysis of release prevention systems is based on a reference facility designed to deliver up to 225 MMscfd of gas to the pipeline during peak-demand periods. The plant consists of a 12.3-MMscfd gas treatment system, a 6.0 MMscfd liquefaction section, a 348,000-bbl storage tank, a 225-MMscfd vaporization system, and a truck terminal capable of shipping or receiving 350 gpm of LNG. The major operations and the safety systems for the plant are briefly described in the following paragraphs. Figure 1 provides a process flow diagram of the facility.

3.1.1 Gas Treatment System

Natural gas from the pipeline first enters a filter separator to remove any free liquids. The 500-psia gas then passes through one of two molecular sieve adsorbers where moisture and CO_2 are removed. Each adsorber is capable of handling 16 MMscfd of gas. The usual flow rate is 12.3 MMscfd. After passing through the adsorber, the gas is filtered to remove dust. Half the treated gas, ∿6 MMscfd, is routed as feed to the liquefaction unit. The rest of the gas is used to regenerate the off-line adsorber. The regeneration gas is first heated to about 550°F in a gas-fired salt bath heater and is then passed through the off-line adsorber. Next, the regeneration gas is filtered to remove free liquids. The gas is then compressed back to line pressure (about 870 psia), cooled in another fan cooler to under 120°F, and then reintroduced into the pipeline.

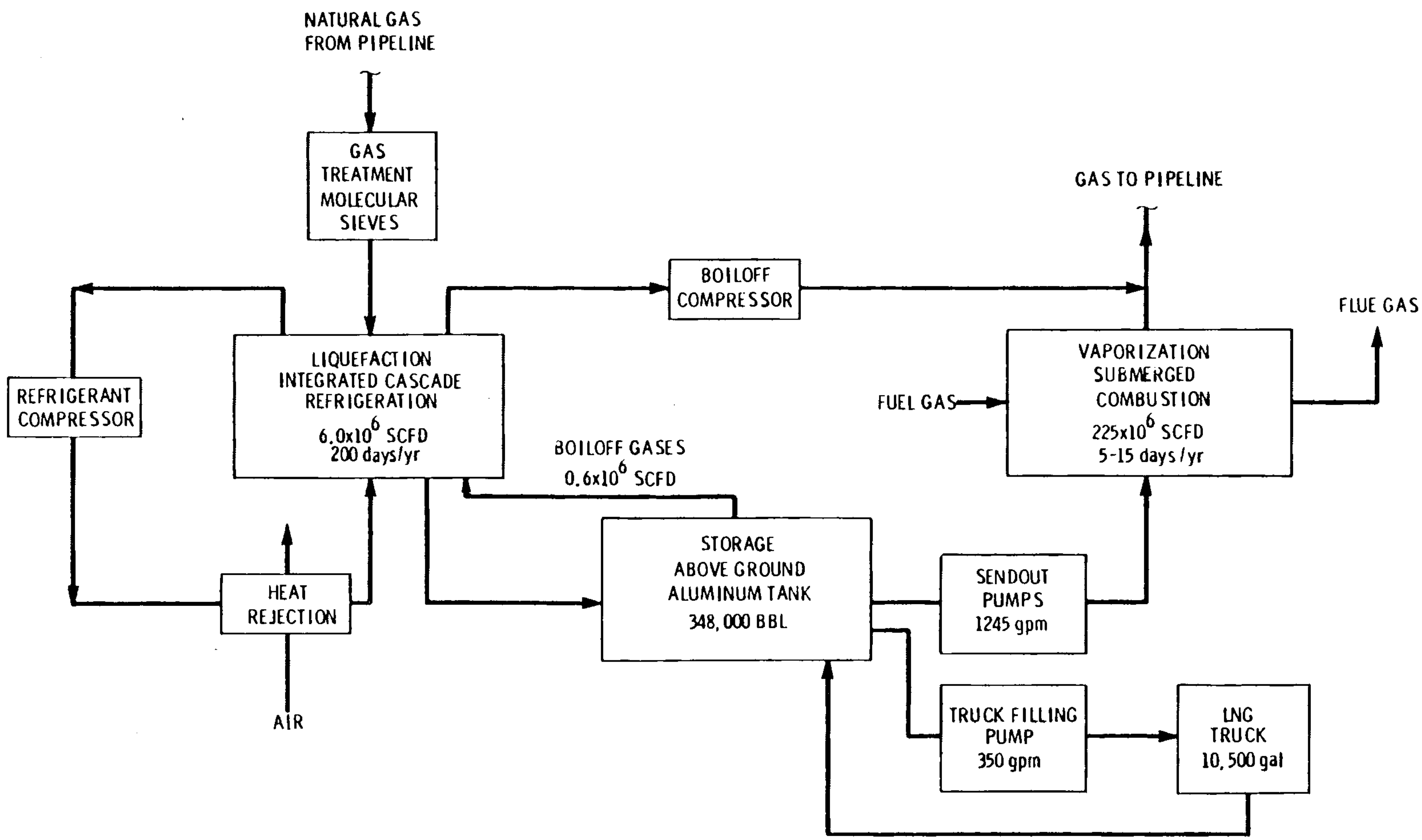


FIGURE 1. LNG Peakshaving Plant - Process Flow Diagram

3.1.2 Liquefaction System

After treatment, the natural gas is cooled and liquefied in a mixed refrigerant cycle to provide LNG for storage. The liquefaction unit is comprised of a cold box, refrigerant compressor and coolers, and refrigerant storage. The cold box consists of heat exchangers, separator vessels, and associated piping and instrumentation all enclosed in an insulated shell. All cold box equipment is constructed of stainless steel, except for the heat exchanger tubing which is aluminum. The natural gas feed enters the cold box at about 500 psia and is passed through a series of six heat exchangers where it is progressively cooled until it is liquefied. The liquefied gas leaves the cold box at about -260°F and about 475 psia. It is expanded to slightly above atmospheric pressure ($\sim$1 psig) as it is introduced into the storage tank.

The mixed refrigerant, which is made up of nitrogen, methane, ethylene, propane, butane, and pentane, is cooled and condensed in stages and then expanded to provide cooling in the cold-box heat exchangers. The refrigerant is then recompressed by a two-stage compressor with inter and after fan coolers for heat rejection. The boiloff gases from the LNG storage tank also provide cooling for the refrigerant in three cold-box heat exchangers.

3.1.3 Storage System

The LNG from the liquefaction system is stored in a flat-bottom, double-walled, above-ground LNG storage tank with a capacity of about 350,000 bbl ($\sim$14.6 million gallons). The inner tank is constructed of aluminum-magnesium alloy AA5083 which has excellent low temperature ductility. Carbon steel is used for the outer tank. The tank dimensions are:

inner tank diameter:	164 ft.
outer tank diameter:	173 ft.
inner tank height:	97 ft.
outer tank height:	134 ft.

The annular space between the inner and outer tank walls is filled with expanded perlite insulation. A resilient fiberglass blanket 12 inches thick is attached to the outside of the inner tank wall to alleviate the effects of the movement of the inner wall due to thermal cycling.

The outer tank has a lap-welded, dome-shaped, steel roof. Suspended from the roof framing of the outer tank is a lap-welded metal deck that serves as a ceiling for the inner tank. Perlite insulation is spread evenly over the deck. Open pipe vents are installed in the deck to allow product vapor to circulate freely in the insulation space to keep the insulation dry. Superheated vapors remain stratified in the upper space, while colder, saturated vapors are below the deck. The butt-welded outer steel shell and lap-welded steel roof provide permanent weather protection for the tank insulation as well as an air-tight seal.

The outer tank rests on a concrete ringwall foundation while the inner tank rests on load-bearing insulation placed on the foundation soil. The bottom of the inner tank is a thin section of aluminum alloy AA5083 that serves only as a seal and is not subject to significant stress. Electrical resistance heating coils are embedded in the foundation soil to prevent freezing of moisture and possible "heaving."

The storage tank is designed to operate at 1.0 psig, with a maximum design pressure of 2.0 psig. The maximum external design pressure is 1 oz. gauge. Tank pressure is controlled by an automatic adjustment of the boiloff compressor recycle rate. The tank is equipped with two pressure relief valves venting to the atmosphere. In the event of an underpressure, gas from the pipeline is brought back into the tank and, if underpressure limits are still exceeded, two vacuum relief valves admit air to the tank. In the event of an emergency, the tank is isolated by block valves on the inlet and outlet liquid lines. The liquid level in the storage tank is monitored and controlled by a servo-powered, displacer-type liquid level device, a differential pressure gauge, and a closed overflow line equipped with a temperature sensor.

Boiloff gases from the storage tank are warmed, compressed to pipeline pressure by one of two compressors, and cooled prior to discharge to the pipeline. Each compressor is capable of handling 1.2 MMscfd of gas. The boiloff gas design rate is about 0.6 MMscfd, with an additional 0.3 MMscfd of flash gas during liquefaction. During liquefaction, the boiloff and flash gases are routed to the coldbox to provide extra cooling as described previously.

3.1.4 Vaporization System

LNG is pumped from the storage tank to the vaporizers by three vertical submerged, pot-mounted pumps. Each pump has a capacity of 75 MMscd or 625 gpm for a total rated sendout capacity of 150 MMscfd with one pump as a spare. The operating temperature is -260°F and the discharge pressure is 945 psia.

The LNG is vaporized in tube bundles submerged in a heated water bath, after which the vaporized natural gas is reintroduced into the pipeline. The vaporizers burn natural gas and bubble the resulting combustion gases through the water bath to heat the water. The three vaporizers have a capacity of 75 MMscfd each. With one vaporizer as a spare, the vaporization capacity of the plant is 150 MMscfd. All vaporization equipment normally carrying LNG is constructed of cryogenic materials to the first flange on the vaporizer outlet.

3.1.5 Transportation and Transfer System

The transportation and transfer system at an LNG peakshaving facility consists of an LNG truck trailer, a transfer system, and a control system. These systems are described in the following subsections.

3.1.5.1 LNG Truck Trailer

Specially designed cryogenic trailers are used to transport LNG to and from peakshaving plants. The main features are a double-walled liquid containment system separated by a perlite-insulation filled annular space. A 50-micron vacuum is established in the annular space for further insulation. This efficient insulation system allows trips of up to four weeks without loss of cargo due to boiloff. The trailer has a capacity of 10,500 gallons, is 40 ft. long, and weighs approximately 21,500 lbs empty and 60,000 lbs fully loaded.

The inner vessel is constructed of cryogenic materials, either 5083-0 Al-Mg alloy, or 9% Ni steel. The outer vacuum jacket is a high strength-low alloy steel. The inner tank is supported by low thermal conductivity support members within the outer jacket. Three flow baffles are utilized inside the inner vessel to prevent sloshing of the liquid cargo. Due to their high center of gravity, these trailers are susceptible to overturning accidents. However, the double-wall construction is very resistant to puncturing and loss of cargo.

The main fill line on the trailer is a 3-inch line passing through the lower half of the shell and is fitted with a manual throttling valve, a remote-operated shutoff valve, and a line safety valve. Other lines included on the trailer are a 2-inch pressure build line which supplies LNG to a pressure build-up coil that vaporizes LNG during unloading to maintain adequate trailer pressure. The trailer is also equipped with three manual trycock valves (liquid level indicators) and numerous pressure relief devices.

3.1.5.2 Transfer System

The transfer system consists of stainless steel liquid and vapor lines connecting the loading/unloading terminal to the facility's storage tanks. The three transfer lines include a 3-inch loading line, a 3-inch unloading line, and a 2-inch vapor return line. These lines are connected to the LNG trailer by flexible metal hoses. A 350 - gpm transfer pump loads LNG into the trailers. Vapor pressure inside the trailer is used to unload LNG at approximately the same rate. Manual shutoff valves are provided in all lines. In addition, liquid and vapor lines also employ remote operated emergency valves.

The terminal area is graded and diked so accidental spills flow away from trailers. Trailers are always grounded and chocked to minimize possible ignition sources in case of a spill during LNG transfer. Dry chemical fire extinguishers and water turrets are provided in the transport terminal area. A closed-circuit television camera continuously monitors the terminal area during cargo transfer and an operator watches for liquid and vapor leaks. Weight scales are installed in the terminal and are used to indicate liquid level in the trailer.

3.1.5.3 Control System

The transport terminals control system consists simply of the pump on/off control, manual valves on all three transfer lines, and remote operated shutoff valves on the liquid and vapor lines. The liquid level in the truck is determined by the weight scales and by opening the 87% full and 90% full trycock valves. Loading pumps are sized to provide the correct flow rates for normal trailer filling. Unloading rates are determined by either the manual valve in the unloading line or the throttle valve on the pressure build line.

Only passive pressure control devices are required on LNG trailers. These include safety valves and burst discs, all of which vent to a common elevated stack.

3.1.5.4 Emergency Shutdown System - Transportation and Transfer

The Emergency Shutdown (ESD) system can be activated manually at the truck terminal or in the control room. The ESD is capable of shutting down or isolating portions of the transportation and transfer system within 30 seconds of activation. If the ESD system fails, it is assumed that ten minutes are required for operators to manually isolate a release.

Combustible gas detectors are located throughout the plant and are assumed to activate an alarm at the truck terminal, alerting the operator to activate the ESD. There is also an alarm at the terminal, activated by a flow detector, which informs the operator of sendout pump performance. This is required at most peakshaving plants because the sendout pump is not visible from the terminal.

Upon activation, the ESD shuts down the transfer pump and isolates the storage tank and truck terminal. Block valves automatically close at the storage tank and at the terminal to stop the flow of LNG through the liquid lines. The ESD also automatically shuts down vapor return valves at the terminal. Trucks are equipped with remotely operated shutoff valves in the 3-inch liquid line, 2-inch pressure build line, and the vapor return line. Manually operated valves are also located at the storage tank, terminal, and trailer as a backup system in case of ESD or remote valve failure. It is assumed that a ten-minute manual shutdown time is required if the ESD does not isolate a release.

3.1.6 Safety Systems

Combustible gas detectors, UV flame detectors, and temperature sensors are located throughout the plant area. In the event of off-standard conditions, these detectors activate alarms in the control room. They can also be set to automatically activate the emergency shutdown system or the fire control system.

The ESD system circuits are energized with 120-V AC power from a separate uninterruptable power supply. When these circuits are deenergized, all valves go to their failsafe positions. This shuts down all equipment, isolates major equipment, isolates the plant from the pipeline and vents gas from all gas handling equipment and lines through the relief header to the vent stack. The ESD can be operated manually from push button stations in the main control panel and at the two exit gates. The ESD is operated automatically by activation of ultraviolet flame detectors.

The fire control system consists of fixed and portable dry chemical fire extinguishers, high-expansion foam systems, Halon fire suppression systems, and a fire-water system. Automatic venting and isolation systems help to prevent accumulations of flammable gas mixtures in enclosed areas and facilitate extinguishment of any fire.

The LNG storage tank and sendout pumps share a spill basin that drains into a diked impoundment basin. The dike walls average 17 ft in height. The impoundment basin is capable of holding about 480,000 bbl, or 1-1/3 times the capacity of the storage tank. High-expansion foam generation systems installed in the spill basin area can be activated either manually or automatically.

The trucking terminal is diked and trenched and is equipped with several dry chemical extinguishers. The spill basin capacity is greater than that of a tank trailer plus the loading/unloading transfer lines.

3.2 REVIEW OF THE OVERVIEW ASSESSMENT

The overview assessment previously performed identified the particular release prevention systems which were more important to plant safety from those which were not significant. The first step was to identify the systems and components which contain natural gas or LNG and determine the flow rates and inventories for each. These results are shown in Table 1. The storage tank and vaporization system have the highest flow rates and inventories.

TABLE 1. System Process Operations Conditions

System	Major Components	Number of Components	Component Capacities	Flow Rates In	Flow Rates Out	Operating Conditions Pressure	Operating Conditions Temperature
Gas Treatment	Adsorbers	2	17,000 scf	12.3 MMscfd (100 gpm)	12.3 MMscfd (100 gpm)	500 psia	68°F
Liquefaction	Cold Box	1	--	6.3 MMscfd (50 gpm)	6.3 MMscfd (50 gpm)	485 psia	-257 to 106°F
Storage	Storage Tank	1	348,000 bbl	6.3 MMscfd (50 gpm)	200 MMscfd (1660 gpm)	15.8 psia	-257°F
	Sendout Pumps	3	--	150 MMscfd (1245 gpm)	150 MMscfd (1245 gpm)	900 psia	-257°F
	Boiloff Compressors	2	--	0.9 MMscfd	0.9 MMscfd	870 psia	120°F
Vaporization	Submerged Combustion Vaporizers	4	--	225 MMscfd (1870 gpm)	225 MMscfd	900 psia	-257 to 70°F
Transportation and Transfer	Truck Trailer	1	10,500 gal	42 MMscfd (350 gpm)	42 MMscfd	15 psia	-257°F

The next step in the overview assessment was a preliminary hazards analysis (PHA). The effects of initiating events such as equipment failures, operator errors, and external events were qualitatively analyzed. The storage system, the vaporization system and the transportation and transfer system have the potential for the largest LNG releases from a peakshaving facility. Key storage section release prevention components include the inner and outer tank structure, the pressure control system, the tank discharge line, and the storage tank pump vessel. Important vaporization system release prevention components include the vaporizer heat exchanger tubes and water bath tank, the vaporizer discharge line, and the temperature controller on the discharge line. Key transportation and transfer release prevention components include the double-shell truck tank and the pressure relief devices. In addition, the operator interface can have a significant effect on release prevention for all systems in the facility.

Release scenarios representing the spectrum of potential releases from a peakshaving facility were developed in the overview analysis. Table 2 presents the representative release scenarios for the storage, vaporization, and transportation sections. These scenarios form the basis for release scenario analyses described in this report.

TABLE 2. Representative Release Scenarios for an LNG Peakshaving Facility

Storage Section

1. Gross Failure of Storage Tank
2. Storage Tank Overfill
3. Storage Tank Overpressure
4. Storage Tank Underpressure
5. Inlet Line Rupture
6. Outlet Line Rupture
7. Sendout Rump Discharge Line Rupture

Vaporization Section

8. Tube Rupture
9. Control Failure and Outlet Line Rupture

Transportation Section

10. Liquid Loading/Unloading Line Failure
11. Flexible Loading/Unloading Hose Failure
12. Vapor Return Line Failure
13. Truck LNG Tank Failure
14. Trailer Pressure Build-up Coil Failure.

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 we felt 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 storage, vaporization and transportation and transfer sections of the reference LNG peakshaving facility 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 (Anyakora 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 contains a summary of the failure rate information used in the release scenario analyses.

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   is our best estimate of the failure rate.

Because operation of LNG peakshaving facilities is not completely automated, the operator is a "component" of the system. Some data on human reliability has been developed (Kletz 1973, Kletz, 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. In our analysis we considered this as a failure.

Operator failure rates for various tasks involved in operation of an LNG terminal are also included in Table 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 we used fault tree analysis 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.

TABLE 3. Generic Failure Rates for Components of LNG Peakshaving Facilities

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^{-4}	(Welker 1979)
	Control System Failure	1×10^{-3}	(Welker 1979)
Pipe Section >3 in dia.	Rupture	1×10^{-10}	(USNRC 1975)
Storage Tank	Rupture	1×10^{-9} 1×10^{-10}	(SAI 1975, Atallah 1980)
	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. (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	$3.\times 10^{-2}$	(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.

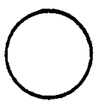
More often than not in our analysis the top event was a large release of LNG from a particular system. After selecting the top event the analyst systematically works backward to identify component faults (basic events), which could cause or contribute to the undesired top event.

Standard symbols shown in Figure 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.

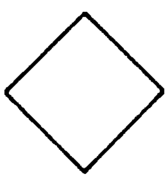
EVENT REPRESENTATIONS

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

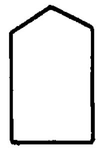
THE CIRCLE DESCRIBES A BACIC FAULT EVENT THAT REQUIRES NO FURTHER DEVELOPMENT. FREQUENCY AND MODE OF FAILURE OF ITEMS SO IDENTIFIED ARE DERIVED FROM EMPIRICAL DATA.



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.

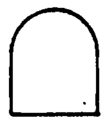


THE HOUSE IS USED AS A SWITCH TO INCLUDE OR ELIMINATE PARTS OF THE FAULT TREE AS THOSE PARTS MAY OR MAY NOT APPLY TO CERTAIN SITUATIONS.



LOGIC OPERATIONS

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



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

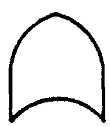


FIGURE 2. Fault Tree Symbols

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.

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 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 shut down the system and isolate the release. For these systems 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.

TABLE 4. Equations Used to Calculate System Safety Characteristics

Basic Events	Minimal Cut Sets	Top Events
Non-Repairable Events		
$\bar{a}_i \leq \lambda_i$ (very close if $\lambda_i t < 0.1$)	$a_k = \prod_{i=1}^{n_k} \bar{a}_i$	$a_T \leq \sum_{k=1}^{n} \bar{a}_k$
$enf_i = \bar{a}_i$	$enf_k = \int_0^t rof_k dt$	$\lambda_T \leq \sum_{k=1}^{n} \lambda_k$
	$rof_k \leq \bar{a}_k \sum_{i=1}^{n=k} \frac{\lambda i}{\bar{a}_1}$	$enf_T \leq \sum_{k=1}^{n} enf_k$
	$\lambda_k \simeq rof_k$	$rof_T \leq \sum_{k=1}^{n} rof_k$
Repairable Events		
$\bar{a} \leq \lambda_i T_i$ (very close if $t > 2T_i$)		
$enf_i = \int_0^t rof_i dt = \int_0^t (1-\bar{a}_i)\lambda_i dt$		
$enf_i = \lambda_i t$ (when $\bar{a}_i$ is small)		

Definitions

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	

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 time used in this study are shown in Tables 5 and 6.

TABLE 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 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

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 immediately 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).

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 ten 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 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 peakshaving facility. The fault trees and supporting calculations for the results reported in this section are given in Appendix B. Table 7 summarizes the results of the release scenario analysis.

4.1 EMERGENCY SHUTDOWN SYSTEMS

An emergency shutdown at an LNG peakshaving facility 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 atmosphere.

Most detectors and sensors provide visual and audible alarms within ten 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 ten 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 will cease almost immediately upon shutdown. Overall, less than one minute is required to totally shut

TABLE 7. Summary of Results of LNG Peakshaving Facility Release Scenario Analysis

Release Scenarios	Expected No. of Events Per Year	Release Occurs And Is Not Stopped in 1 Min. by ESD Expected No. of Events Per Year	Maximum Release Size (Equivalent Gallons of LNG) 1 Min.	10 Min.	Material Released	Critical System Components
Storage System						
Gross Failure of Storage Tank	1×10^{-5}	-	-	1.5×10^{7}	LNG	Storage Tank
Storage Tank Is Overfilled	3×10^{-4}	-	(a)	(a)	Natural Gas/LNG	Operator, Level Detectors
Storage Tank Is Over-pressured and Relief Valves Open	1	-	5-2400[b]	50-24000[b]	Natural Gas	Pressure Detector, Pressure Controller, Operator
Storage Tank Is Over-pressured and Relief Valves Fail	1×10^{-6}		95-285[c]	1.1×10^{6} - 3.4×10^{6}[c]	Natural Gas/LNG	Relief Valves
Storage Tank Is Under-pressured and Relief Valves Open	5×10^{-3}		-	-		Pressure Detector, Pressure Controller
Storage Tank Is Under-pressured and Relief Valves Fail	5×10^{-9}		95-285[c]	1.1×10^{6} - 3.4×10^{6}[c]	Natural Gas/LNG	Relief Valves
Rupture of Storage Tank Inlet Line	5×10^{-4}	6×10^{-4}	50	500	LNG	Expansion Joint, Operator
Rupture of Storage Tank Outlet Line	5×10^{-5}	1×10^{-5}	28000	280000	LNG	Expansion Joint, Internal Valve, Operator
Rupture of Pump Discharge Line	2×10^{-3}	2×10^{-5}	625	6250	LNG	Expansion Joint, Operator

TABLE 7. con't

Vaporizer Section						
Vaporizer Tube (Sub. Comb.) Rupture	1×10^{-1}	1×10^{-3}	1250	12500	Natural Gas/LNG	Heat Exchanger Tubes, Operator
Rupture of Vaporizer Outlet Line from Cold Gas (Vaporizer Control Failure)	9×10^{-2}	9×10^{-4}	1250	12500	Natural Gas/LNG	Vaporizer Outlet Temp. Controller, Operator
Transportation and Transfer						
Rupture of 3" Liquid Loading Line During Loading	1×10^{-4}	2×10^{-6}	460	3600	LNG	Expansion Joints, Transfer Pump Operator, Gas Detectors
Rupture of 3" Liquid Unloading Line During Unloading	6×10^{-5}	1×10^{-6}	870	8200	LNG	Expansion Joints, Operator, Gas Detectors
Rupture of 350 gpm Sendout Pump During Loading	1×10^{-4}	1×10^{-5}	1800	18200	LNG	Transfer Pump, Gas Detectors, Flow Detector, Operator
Rupture of 2" Vapor Return Line During Loading	1×10^{-4}	2×10^{-6}	1100 ft^3	11000 ft^3	Natural Gas	Expansion Joints, Operator, Gas Detectors
Rupture of Flexible Metal Hose During Load/Unload	6×10^{-4} 3×10^{-4}	1×10^{-5} 6×10^{-6}	Load: 470 Unload: 1600	Load: 3600 Unload: 10500	LNG LNG	Flexible Hose, Connectors, Operator
Rupture of Pressure Buildup Coil During Unloading	6×10^{-7}	6×10^{-9}	200	2000	LNG	Valves, Operator
Trailer Accidents	1.7×10^{-2}	-	10500	-	LNG	Driver, Double-Walled Tank, Pressure Relief Devices

(a) Could cause failure of outer shell and roof. Failure of inner tank possible.
(b) Lower number is typical of 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 200 hours (time to pump tank down).

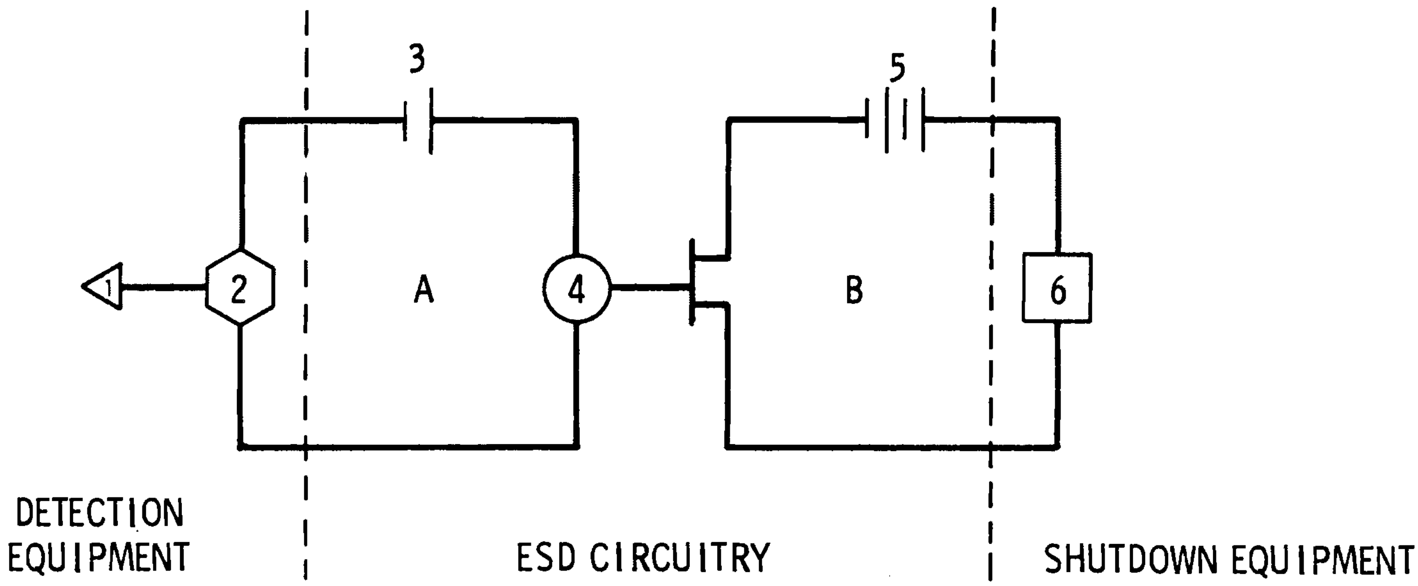
down the plant when an emergency situation occurs. These are the design shutdown times thought to be typical of new LNG peakshaving facility.

The reference LNG peakshaving facility has two major shutdown circuits: the Vaporizer Emergency Shutdown (VES) and the Master Emergency Shutdown (MES). The MES shuts down the entire plant and the VES stops sendout and vaporization operations.

During normal operation the ESD circuits at the reference peakshaving facility 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 3 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 MES circuitry cannot be checked for operability at specified intervals since checking the circuitry would result in an unscheduled shutdown. We assumed a test interval of three months for both VES and MES which gives an average unavailability of 1×10^{-4}. 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}.

Figure B.1 and B.2 show fault trees for failure of the Emergency Shutdown Systems. While it is desirable that pumps (or compressors) are shut down and block valves are closed, we considered it a successful shutdown if either one occurs and flow is stopped. As a result, ESD circuitry failures dominate over processing equipment failures.



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

Components:

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

FIGURE 3. Simple Fail-Safe Emergency Shutdown System

Athough 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. We considered this 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^{-2} per demand depending on the situation and the system involved. Figure B.1 is a fault tree for a fully automatic emergency shutdown. The failure rate for fully automatic emergency shutdown is 10^{-1} to 10^{-2} 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-2. The following section discusses release scenarios at LNG peakshaving facilities, 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 STORAGE SECTION

LNG storage tanks deserve special consideration because of the magnitude of potential releases from them. Each tank when full contains 348,000 barrels. A simplified fault tree for failure of an LNG storage tank is shown in Figure B.3 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 (Bampton 1980) 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.

4.2.1 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 Inc. 1971). Atallah (1980) reports the frequency of fires/explosions in LNG tanks to be 3.3×10^{-6} per year.

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.

4.2.2 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 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 tank has two liquid level indicators, one displace-type and one differential pressure gauge. Both level indicators sound alarms in the control room when a pre-set maximum level is reached. If the operator does not stop flow into the tank, a low temperature sensor will sound another alarm in the control room. This alarm is set for 95 feet, or two feet below the top of the inner tank.

The storage tank is filled at a rate of 1710 bbl LNG/day for 200 days/yr. Based on generic failure data, a tank will be overfilled 3×10^{-4} times per year. A fault tree for overfilling a storage tank is shown in Figure B.4. The critical components in the system are level indicators, 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.

4.2.3 Storage Tank Overpressure

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

LNG vapor is removed from the storage tank by boiloff compressors. The pressure in the tank is controlled by adjusting the boiloff compressor recycle rate. The tank has a pressure controller which maintains the absolute pressure in the tank at 15.7 psia by adjusting the boiloff compressor recycle rate. The tank also has a gauge pressure control system. Two 12 in. pressure relief valves operate to relieve overpressurization. Data from peakshaving plants indicate that relief or vent valves are opened approximately once per year. A fault tree for a storage tank reaching vent pressure is included in Figure B.5 .

The probability of overpressure not being relieved by both of the pressure relief valves is 1×10^{-6} times per year.

The amount of vapor released from the relief valves will vary depending on the situation. The maximum venting capacity of the two 12 in. valves is 2400 equivalent gal/min.

The vapor-generation rate from an open-top tank can be used to approximate the vapor release rate from a tank whose roof has failed from overpressure. In the time required to pump out the tank, 1.0×10^6 to 2.7×10^6 equivalent gallons of LNG would be vaporized.

4.2.4 Storage Tank Underpressure

The minimum design pressure for the reference facility storage tank is 0 psig. The tank can withstand only about 1 oz gauge external pressure before it will collapse.

In the event of the underpressure, gas from the pipeline is brought back into the tank through the 8 in. vapor outlet line. If this is insufficient to prevent underpressure damage, two 12 in. vacuum relief valves admit air into the tank.

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 storage tank reaching vacuum relief pressure is shown in Figure B.6. Normal operating conditions present no problems with respect to reducing tank pressure.

The probability of the storage tank reaching vacuum relief pressure is 5×10^{-3} times per year. Failure rate for the pressure relief valves is 1×10^{-6} per demand resulting in a valve failure 5×10^{-9} times per year.

4.2.5 Inlet Line Rupture

The inlet line to the storage tank is 3 in. in diameter and enters through the bottom of the outer tank and top of inner tank. It consists of about seventeen 20-ft. pipe sections with expansion joints between them. The storage tank is filled 200 days per year. Analysis indicates the inlet line will rupture 5×10^{-4} times per year, resulting in a 50 gallon LNG spill if stopped in one minute or 500 gallons if stopped in 10 minutes. The probability of the release occurring and not being stopped in one minute is 6×10^{-6} times per year. A fault tree analysis for rupture of the inlet line is shown in Figure B.7. No major damage to the storage tank is expected.

4.2.6 Outlet Line Rupture

The outlet line from the storage tank is 12 in. in diameter and exits through the bottom of the inner tank. LNG is drawn from the tank for vaporization about 20 days per year. The outlet line will rupture about 5×10^{-5} times per year, resulting in a release of 28,000 gallons of LNG if stopped in one minute. If the release is isolated in ten minutes, 280,000 equivalent gallons of LNG will be released. The probability of the release occurring and not being stopped in one minute is 1×10^{-5} times per year. A fault tree analysis for rupture of the outlet line is shown in Figure B.8. Critical system components are expansion joints, 12 in. internal valve, and operators.

4.2.7 Rupture of Pump Discharge Line

The three LNG sendout pumps are vertically submerged, pot-mounted LNG pump systems. The pumps and the motor drive are hermetically sealed in a vessel and submerged in LNG. This design eliminates the extended pump shaft and the associated seal. The pump and motor surroundings are 100% rich with LNG and will not support combustion. The pumps are mounted in a suction pot below grade to provide sufficient suction head for operation. Two pumps operate 20 days per year with the third pump as a spare.

A rupture of a pump discharge line will occur 2×10^{-3} times per year resulting in a 625 equivalent gallon release if isolated in one minute. If the release is not stopped in one minute, about 6250 gallons of LNG will be released in ten minutes. This will occur 2×10^{-5} times per year. Figure B.9 shows a fault tree analysis for a large release of LNG from the pump discharge line.

4.3 VAPORIZATION SECTION

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

4.3.1 Vaporizer Tube Rupture

The vaporizers for the referenced peakshaving facility are three submerged combustion units, two of which operate 20 days per year with the third unit as a spare. A recent study (Welker 1979) indicated that submerged combustion vaporizer tube failures occurred at a rate of approximately 1×10^{-4} per hour or 5×10^{-2} times/yr. The high failure rate is inherent in the design of the vaporizers, i.e., a large number of small high-pressure tubes.

Because the vaporizers operate at high pressure (900 psig), gas or liquid will be released at a high rate from a single tube until the system is depressurized. 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 one minute, 1250 equivalent gallons of LNG will be released. If the vaporizers are not shut down, over 12500 equivalent gallons of LNG will be released in ten minutes.

The probability of a rupture in a tube and the release not being stopped in one minute is 1×10^{-3} per year. Critical components are the vaporizer tubes and operators. A fault tree analysis for vaporizer tube failure is shown in Figure B.10.

4.3.2 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. 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. A temperature sensor in the gas outlet line will activate a low-temperature alarm in the control room if cold gas reaches the outlet.

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 about 9×10^{-2} times per year. If the ESD is activated automatically by the process sensors, such failures will only occur about 2×10^{-3} times per year. If the ESD is activated after the spill occurs and shuts down the system in one minute, 1250 equivalent gallons of LNG will be released. If the system is not shut down for ten minutes, 12500 gallons will be released. This will occur 9×10^{-4} times per year. Figure B-11 shows a fault tree for vaporizer outlet line rupture.

4.4 TRANSPORTATION AND TRANSFER SYSTEM

Representative release events for the transportation and transfer system were shown previously in Table 2. These included releases from the LNG semi-trailer, from piping and valves at the transport terminal, and from lines leading to and from the storage tank. Some release scenarios apply to other parts of the facility while others are unique to the transportation and transfer system.

4.4.1 LNG Trailers

Large, double-walled LNG semi-trailers are used at peakshaving plants to ship from as few as 50 to as many as 2300 truckloads per year. (DOE, 1978) The average number of deliveries a typical plant makes is about 500 annually, which works out to be approximately 30% of the storage tank. Assuming 90% of incoming empty trailers are cold (1/2 hour to load) and 10% are warm (four hours to load), a typical transport terminal operates 340 hours per year. An average transit distance of 86 miles was calculated using data from the Distrigas Terminal, Everett, Mass. (A.D. Little, Inc., 1978) Although this is not a peakshaving plant, it is capable of up to 20,000 truck shipments per year with 14,000 not uncommon. Since LNG trucking is used most extensively in the northeast, 86 miles is assumed to be a typical trucking distance. The following sections will examine release scenarios which apply to LNG trucking.

4.4.1.1 Highway Accidents

No scenario is given for a truck accident. There have been 14 LNG truck accidents since trucking began in the late 1960's. This includes approximately 26 million miles of LNG transport. (A. D. Little, Inc., 1978) Of these accidents, one occurred at the loading terminal and the rest on the highway. Although nine accidents resulted in rollover of the trailer, due to the structural integrity of the double-walled construction, none involved a major spill or fire. From this data, the probability of a truck being in an accident is 5×10^{-7} per mile. Using this frequency, the expected number of accidents involving trucking from a typical peakshaving plant is 1.7×10^{-2} per year.

A. D. Little, Inc., (1978) several years ago, made an attempt to theoretically evaluate the probabilities of an LNG truck being in an accident serious enough to breach the cargo tank. Their results are as follows:

(1) Catastrophic spill; 40 m^3 (10,500 gal) instantaneous spill
probability = 5×10^{-9} per mile

(2) Serious spill; 4 m^3 (1,000 gal) instantaneous or 40 m^3 in 5 minutes
probability = 2.5×10^{-8} per mile

(3) Minor spill; valve leakage, small tank puncture
probability = 1×10^{-7} per mile.

Due to uncertainties both in accident rates and conditional spill probabilities, this data is presented for relative comparison only.

4.4.1.2 LNG Trailer Overpressure

Overpressure of an LNG trailer can occur only if pressure builds up in the tank and relief valves, burst discs, and blowdown valves fail simultaneously. Overpressure while in transit is difficult to achieve because of the high efficiency of the insulation system. A loaded trailer is capable of going four weeks before boil-off gases cause the trailer to reach vent pressure. A trailer involved in a fire most likely will not overpressurize because the insulation system is also heat resistant and the steel outer shell is rugged enough to withstand the heat. Some trailers have been built with an aluminum outer shell and would probably fail if engulfed in flame. These trailers are no longer in LNG service. Therefore, trailer loading/unloading operations are the most likely source of overpressure events.

Operator actions are very important during the loading/unloading sequence. He must determine truck status (warm, cold, damaged, etc.) and take proper actions. There is an operational procedure that includes chocking, grounding, inspection, connecting hoses, opening valves, etc., that must be followed. Operators must also remain alert and monitor gauges and alarms at all times. Due to the diversity of tasks and possible high stress situations, human errors are more likely to contribute to a release sequence than a mechanical malfunction.

Overpressurization of a semi-trailer could cause a complete collapse and release the entire tank's contents. A fault tree for failure of an LNG trailer is shown in Figure B.12. The probability of a failure due to overpressurization is reduced considerably by the presence of a redundant pressure relief devices (safety valves, burst discs) on the trailer. Another safety device called the differential pressure control valve (not shown) also reduces the expected number of failures. This device automatically shuts down loading when trailer pressure reaches 20 psig.

LNG terminals are designed to contain a complete trailer failure. The loading area is diked and channeled so spills flow away from the trailer. Trailers are grounded and non-sparking tools used to reduce the number of ignition sources. Therefore, the expected number of highway accidents and the possibility of tank failure in uncontrolled situations leads one to believe the dominant LNG trailer failure mode is highway accidents.

4.4.2 Liquid Lines

The reference description includes two 3-inch liquid lines, one for unloading trailers and one for loading, which connect the storage tank to the truck terminal. These lines are typically 300 ft long and made of 20 ft sections. An expansion joint is installed every five sections. Both remotely and manually operated block valves are installed at the storage tank and at the terminal in each line. The loading line also contains a transfer pump. Although unloading is rarely done at a peakshaving plant, situations exist where it may be necessary, e.g., if the liquefaction unit breaks down.

A release from a loading or unloading line will occur about 1×10^{-4} or 6×10^{-5} times per year, respectively. Failure modes include corrosion, metal fatigue, cavitation, and external events. Combustible gas detectors activate an alarm at the control panel in the event of a release. If the operator activates the ESD in one minute, the maximum spill size is 460 gallons if loading and 870 gallons if unloading. If the ESD system fails, up to 3600 gallons or 8200 gallons could be released during loading or unloading, respectively. This will occur approximately 1.8×10^{-2} times per demand, resulting in a probability of 2×10^{-6} per year if loading and 1×10^{-6} per year if unloading. The difference between these probabilities is due to a transfer pump located in the loading line. If the ESD fails to operate, manual valves at the storage tank and transport terminal can be closed to stop flow. Fault trees are shown in Figure B.13 for the loading line and Figure B.14 for the unloading line.

4.4.3 Sendout Pump

A 350 gpm sendout pump is used to load LNG into trailers. In a few isolated cases, pumps are also used to unload trailers, although pressure unloading is most common. A release of LNG from the sendout pump, suction line, and associated valves will occur about 1×10^{-4} times per year. The amount released in one minute is a maximum of 1800 gallons.

In the reference facility, the sendout pump's performance is measured by a flow detector which sounds an alarm at the control panel in case of a loss of flow. Gas detectors also alarm at the control panel in case of a release event. If the ESD is not activated, 18,200 gallons of LNG could be spilled in ten minutes. This will occur about 8.8×10^{-2} per demand or about 1×10^{-5} times per year at the reference facility. A fault tree for a large release from the sendout pump is shown in Figure B.15.

4.4.4 Vapor Return Line

The vapor return line is a 2-inch line that returns vapor to the storage tank during trailer loading operations. It is approximately 300 ft long and contains three expansion joints, a check valve, and two block valves. The expected number of release initiating events for this scenario is 1×10^{-4} per year. Failure modes include pressure and thermal cycling fatigue and corrosion.

In the event of a vapor return line rupture, a check valve stops backflow into the trailer. The gas detection system sounds an alarm at the control panel. If the operator activates the ESD in one minute, 1100 ft^3 of natural gas could be released. If the ESD is not activated and it takes ten minutes to isolate the rupture, 11,000 ft^3 of natural gas would be released. This will occur about 1.8×10^{-2} times per demand or 2×10^{-6} times per year for this scenario. The release is isolated by closing vapor return valves at the terminal and storage tank, and shutting off the flow of LNG into the trailer. Figure B.16 shows a fault tree for a large release from the vapor return line.

4.4.5 Pressure Build-up System

LNG trailers are generally unloaded using the vapor pressure above the liquid to force it out. If the vapor pressure is too low, some liquid is routed to a pressure build-up coil by a 2-inch line. This liquid is vaporized and sent to the tank top to increase the pressure and the unloading rate (350 gpm typically). This system includes pipe sections, valves, and a check valve on the vapor side of the coil. A release of LNG from this sytem will occur about 6×10^{-7} times per year, releasing a maximum of 200 gallons if the ESD operates.

The pressure build-up system is shut down by operating valves, remote or manual, in the liquid side of the coil. There is also a separate pressure build-up valve which can be operated that will cut off LNG flow before the pressure build coil. If the ESD fails to operate, up to 2000 gallons could be released in ten minutes. This will occur about 1×10^{-2} times per demand, resulting in a probability of 6×10^{-9} per year for this scenario. If the ESD fails, there is a manual shutoff valve on the liquid line to stop the flow of LNG. A fault tree for failure of the pressure build-up system and ESD is shown in Figure B.17.

4.4.6 Flexible Metal Hose

A 3-inch, flexible metal hose is used to connect the LNG trailer to the transport terminal. The historic failure rate for this type of hose is 1.7×10^{-6}/hour.

In case of failure of the flexible hose, the ESD closes upstream valves and the pump is turned off to stop flow. Assuming it takes one minute to activate the ESD, 470 gallons could be released if loading and 1600 gallons if unloading. The probability of the ESD failing is about 2×10^{-2} per demand. If loading, a maximum of 3600 gallons could be released. If unloading, the entire tank's contents could be spilled in the terminal area. Assuming truck loading operations are conducted 340 hours per year and unloading operations are conducted 200 hours per year, this results in a probability of 1×10^{-5} and 6×10^{-6} events per year. In case of an ESD failure, operators can close manual valves at the terminal and at the truck to stop flow.

5.0 EVALUATION OF RELEASE PREVENTION ALTERNATIVES

The following sections discuss the results of a preliminary evaluation of the design alternatives given below:

Storage Tank

- In-Tank Pumps
- Double Ply Expansion Joint on Storage Tank Outlet Line
- Block Valve Upstream of the Outlet Line Expansion Joint

Emergency Shutdown System

- Automatically Actived Emergency Shutdown System.

Transportation and Transfer System

Driver Training
Design Changes to Semi-Trailers
Transport Terminal Barrier
Fail-Safe Transfer Line

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 DESIGN ALTERNATIVES FOR STORAGE TANK

Three design alternatives for the reference storage tank are discussed below.

5.1.1 In-Tank Pumps

The reference storage tank is pumped out by external pot mounted pumps that take suction from a withdrawal line that exits the bottom of the storage tank. The design alternative is to pump out the tank with in-tank pumps submerged in the LNG with the pump discharge line exiting the tank through the roof. If this design were applied to the reference peakshaving facility, it would reduce the one minute maximum release from 28,000 gallons of LNG to 2,000 gallons of LNG and the ten minute release from 280,000 gallons of LNG to 13,000 gallons of LNG. Figure B.18 gives a fault tree for this alternative.

5.1.2 Double Ply Expansion Joint

The reference storage tank has a single ply expansion joint on the storage tank withdrawal line at the point where the line exits the outer tank shell. The design alternative is to replace the single ply expansion joint with a double ply expansion joint. The two plies would be made of stainless steel with each ply designed for 300 psig. The space between the plies is pressurized 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 indicates a fault in the outer ply. (Osborn, 1979)

If this design were applied to the reference peakshaving facility, it would reduce the number of expected large releases from the storage tank outlet line from 5×10^{-5} per year to 1×10^{-5}/yr. A fault tree for the failure of the storage tank outlet line with a double ply expansion joint is given in Figure B.19.

5.1.3 Block Valve Upstream of Expansion Joint

The reference storage tank has a flapper valve inside the tank to close off the bottom withdrawal line. There are no other block valves downstream until after the expansion joint. The design alternative is to insert a block valve between the flapper valve and the expansion joint.

If this design were applied to the reference peakshaving facility, it would reduce the probability of a major release from the storage tank outlet line from 1×10^{-5} per year to 1×10^{-6} per year assuming the ESD stopped the release in one minute. A fault tree for the failure of the outlet line with a block valve upstream of the expansion joint is given in Figure B.20.

5.2 EMERGENCY SHUTDOWN SYSTEM (ESD) ALTERNATIVE

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., an LNG release may be detected by both a combustible gas detector and a low temperature detector).

5.3 EVALUATION OF RELEASE PREVENTION ALTERNATIVES IN THE TRANSPORTATION AND TRANSFER SYSTEM

Design alternatives are discussed for four separate areas of the transportation and transfer system.

5.3.1 Drivers

A problem associated with LNG trailers seems to be their unusually high center of gravity. Because of this, truck rollover resulted in 69% (9 of 13) of all LNG trailer accidents in transit. Three of these accidents were caused by failure to negotiate a highway turnoff and all resulted in rollover. Other accidents occurred because drivers had swerved to avoid a pedestrian (one case) or had driven off the road because of a tire blowout (two cases). These types of accidents could be reduced by an LNG trucking course in conjunction with a defensive driving course.

LNG trucking suffers from being seasonal work. Some good drivers may choose not to drive LNG trucks because the work is irregular. (GAO, 1978) Therefore, LNG drivers are not specifically assigned to an LNG truck. As of December 1978, driver training to some extent was being given to drivers who ship out of the Distrigas Terminal, Everett, Massachusetts. At that time the New England Gas Association was compiling a training manual in order to standardize the program. As drivers become aware of the specific problems encountered with LNG trucking and improve their defensive driving knowledge, the number of truck rollover accidents is expected to decrease.

5.3.2 Design Changes to Semi-Trailers

LNG cargo tanks are inherently resistant to failure caused by external events due to the strength of the double-walled construction. However, the piping and valve controls are contained in a compartment at the back of the trailer and may be susceptible to damage by rear-end collisions. A design change which would eliminate this type of damage would be to locate the valve controls under the trailer. Unfortunately, this would result in more complicated loading/unloading operations and may increase the expected number of failures at the truck terminal.

Another method of reducing the chances of a rear-end collision causing a release would be to use fail-safe valves currently used on trucks handling liquefied petroleum gas (LPG). These valves are fabricated with a weak point which shears preferentially but allows the valve to remain closed. Some LNG trailers are already equipped with this type of valve.

Difficulty has arisen with burst discs in the vent lines. Burst discs occasionally rupture at pressures below the design rating because of metal creep. This type of failure enroute may cause a release. In the event of a disc rupture, the trailer must be vented down to atmospheric pressure and a new one installed. All trucks should carry a spare burst disc and proper wrenches. Replacing the burst disc with an additional safety valve could eliminate this problem. It is not known if this would be compatible with current trailer designs.

Operator errors may contribute more to failure rates than mechanical malfunctions. Operators have a loading procedure which should be followed at all times. Failure to follow this procedure could result in unsafe conditions. Therefore, the addition of more interlocks on the trailer and transport terminal which would force operators to follow procedures could significantly reduce the number of accidents. For example, the addition of an interlock to prevent the pressure build-up coil from operating without the trailer being unloaded would reduce the probability of pressure build-up in the trailer. An interlock which would not allow the trailer to be loaded without the vapor return line open may also be desirable, considering the consequences.

5.3.3 Transport Terminal Barrier

A design improvement which could be incorporated into the loading terminal is a barrier to protect the piping and controls from a mishandled vehicle. Although an accident of this kind has not occurred, a substantial release of LNG could be the result. The barrier would have to be designed to stop a loaded trailer before contacting the LNG piping without severely damaging the trailer. A barrier system currently employed consists of hollow collapsible drums which slow the vehicle while transferring minimal force to the semi-trailer. These drums are connected to and backed by a steel support structure in concrete designed to withstand the full trailer force. (Howard, 1978) This type of barrier could be of great benefit to eliminate or reduce the chance of an incident occurring.

5.3.4 Fail-Safe Transfer Line

A fail-safe transfer line for hazardous fluids has been designed and tested. (Houghton, Simmons, Gonso, 1980) This line would replace the flexible metal hoses now used. It would not significantly reduce the probability of a line rupture but would decrease the response time, and therefore the release quantity, in case a rupture or leak occurs.

This system uses inlet and outlet flow meters to measure the pressure drop in the line. If a break or leak occurs and the pressure drop exceeds a predetermined level, the control module automatically closes inlet and outlet valves. It is assumed that the control module is also capable of shutting down the transfer pump to keep pressure from building in the liquid lines. Response time varies from 64 milliseconds for a guillotine break up to 16 seconds, depending upon the magnitude of the pressure drop.

6.0 CONCLUSIONS AND RECOMMENDATIONS

This report has analyzed release prevention systems of an LNG peakshaving facility. 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 9 summarizes this analysis.

The three plant areas identified as being most significant with respect to safety are storage, vaporization, and transportation and transfer areas. Gross failure of the storage tank, rupture of the storage tank outlet line, and rupture of an LNG semi-trailer tank are the three 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 a peakshaving facility have been identified. They are listed in Section 5. 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 peakshaving facility 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 may not be justified; however, for facilities located in highly populated areas, these alternatives deserve serious considerations.

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APPENDIX A

FAILURE MODES AND EFFECTS ANALYSIS

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis

Component	Failure: Mode	Failure: Frequency	Effect	Class	Compensating Provisions
2" LNG feed line from liquefaction unit, ~500' long	Leaks or ruptures	3×10^{-7} to 5×10^{-10} /hr	LNG release	Critical	ESD system, spill basin
2" flow control valve on LNG feed line (air operated)	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	LNG release	Critical	ESD system, spill basin
	Fails open	10^{-3} to 10^{-4}/demand	LNG flow is not stopped	Marginal	Other valves in liquefaction unit
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot fill tank	Safe	Pressure relief valves on line
2" flanged joint in LNG feed line	Leaks or ruptures	10^{-8}/hr	LNG release	Critical	ESD system, spill basin
LNG inlet spray header	Uneven LNG distribution	10^{-5} to 10^{-6}/hr	Uneven tank cooldown, possible tank failure and subsequent LNG release	Marginal	Tank thermocouples, ESD system
Inner shell of storage tank	Leaks or ruptures	2×10^{-6} to 3×10^{-9} /hr	LNG leaked to outer tank, probable failure of outer tank and release of LNG	Critical	Spill basin, tank pumpout capability
Suspended, insulated roof deck	Structural failure	10^{-8} to 10^{-10}/hr	Loss of tank dome insulation, probable dome failure and natural gas release	Critical	Tank pumpout capability
Outer shell of storage tank	Leaks or ruptures	2×10^{-6} to 3×10^{-9} /hr	Release of natural gas, possible failure of inner tank and release of total contents	Critical	Tank pumpout capability

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Tank foundation heating coils	Fail to operate	10^{-5} to 10^{-6}/hr	Freezing of foundation soil, possible tank failure and LNG release	Marginal	Heating coil replaceability, temperature instrumentation
Tank temperature instrumentation or linear movement indicators	Fail to operate properly	3×10^{-4} to 3×10^{-6}/hr	Uneven stresses on tank, possible tank failure and LNG release	Marginal	
12" pressure relief valve on tank (two)	Leaks or ruptures	5×10^{-8} to 1×10^{-8}/hr	Natural gas release	Marginal	Shut off valve prior to relief valve
	Fails open	10^{-5}/hr	Natural gas release, possible air admitted to tank	Marginal	Shut off valve prior to relief valve
	Fails closed	10^{-5}/demand	Possible overpressurization of tank, could lead to tank failure	Marginal	Redundant (two) relief valves, manual relief valve
12" vacuum relief valve on tank (two)	Leaks or ruptures	5×10^{-8} to 1×10^{-8}/hr	Natural gas release	Marginal	Shut off valve prior to relief valve
	Fails open	10^{-5}/hr	Natural gas release, possible air admitted to tank	Marginal	Shut off valve prior to relief valve
	Fails closed	10^{-5}/demand	Possible underpressurization of tank, could lead to tank failure	Marginal	Redundant (two) relief valves, natural gas addition capability, manual relief valve
12" shutoff valves on relief lines (two each for pressure and vacuum relief) manually operated	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	Natural gas release	Critical	Tank pumpout capability
	Fails open	10^{-6} to 10^{-7}/demand	Cannot isolate relief valve	Safe	
	Fails closed	3×10^{-4} to 3×10^{-5}/demand	Relief valve isolated	Marginal	Redundancy of relief systems

TABLE A.1 LNG Peakshaving Facility/Storage Section
Failure Modes and Effects Analysis (cont)

Component	Failure: Mode	Failure: Frequency	Effect	Class	Compensating Provisions
Control valve on natural gas addition line for vacuum relief (air-operated)	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	Natural gas release	Marginal	Shut off valves in line
	Fails open	10^{-3} to 10^{-4}/demand	Natural gas admitted to tank, possible overpressurization and subsequent release	Marginal	Shut off valves in line, pressure relief valves on tank
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot add natural gas to tank, possible underpressurization and subsequent release	Marginal	Vacuum relief valves on tank
Low pressure switch controlling natural gas addition to tank	Trips at too high pressure	4×10^{-5} to 1×10^{-8} /hr	Natural gas admitted to tank, possible overpressure and subsequent release	Marginal	Pressure relief valves on tank, shut off valves on line, tank pressure instrumentation
	Fails to trip	4×10^{-5} to 1×10^{-7} /hr	Natural gas not admitted to tank, possible underpressure and subsequent release	Marginal	Vacuum relief valves on tank
Tank liquid level instrumentation	Fails to operate properly	2×10^{-4} to 3×10^{-6} /hr	Possible overfilling of tank with subsequent failure and release of LNG	Critical	Redundancy of liquid level instrumentation
8" boiloff line, ∿500' long	Leaks or ruptures	3×10^{-7} to 5×10^{-10} /hr	Natural gas release	Critical	Shut off valves at ends of line
Boiloff routing selector switch	Fails to operate properly	3×10^{-5} to 3×10^{-6}/demand	Boiloff gases may be routed improperly	Marginal	Routing valves operable manually

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Boiloff routing valves (air operated, selector switch controlled):					
a) To cold box	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	Natural gas release	Critical	Shut off valves in line
	Fails open	10^{-3} to 10^{-4}/demand	Cannot isolate boiloff system from cold box	Safe	
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot route boiloff to cold box	Safe	Boiloff heat exchangers
b) From cold box	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	Natural gas release	Critical	Shutoff valves in line
	Fails open	10^{-3} to 10^{-4}/demand	Cannot isolate boiloff system from cold box	Safe	
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot route boiloff through cold box	Safe	Boiloff heat exchangers
c) To boiloff heat exchangers	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	Natural gas release	Critical	Shutoff valves in line
	Fails open	10^{-3} to 10^{-4}/demand	Cannot route boiloff through cold box	Safe	Boiloff heat exchangers
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot route boiloff directly to boiloff heat exchangers	Safe	Bypass through cold box
Boiloff line to or from cold box, ~50' long each	Leaks or ruptures	$3x10^{-8}$ to $1x10^{-10}$ /hr	Natural gas release	Critical	Boiloff routing valves

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Boiloff heat exchanger (two, one each in parallel boiloff systems)	Leaks or ruptures	10^{-4} to 10^{-5}/hr	Natural gas release	Critical	Shut off valves in line, redundant (two parallel) boiloff systems
	Fails to provide sufficient warming of boiloff gases	10^{-4} to 10^{-5}/hr	Possible brittle failure of downstream piping	Marginal	Redundant boiloff systems, temperature instrumentation
	Fouls or plugs	10^{-4} to 10^{-5}/hr	Cannot handle sufficient volume of boiloff gases, possible overpressure of lines	Marginal	Redundant boiloff systems
Boiloff heat exchanger inlet valve (air operated, temperature relay controlled) two, one each in parallel boiloff systems	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	Natural gas release	Critical	Shut off valves in line, parallel boiloff systems
	Fails open	10^{-3} to 10^{-4}/demand	Cannot bypass heat exchanger	Safe	Parallel boiloff systems
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot warm boiloff gases in heat exchanger	Marginal	Parallel boiloff systems, temperature instrumentation
Boiloff heat exchanger bypass valve (air operated, temperature relay controlled) two, one each in parallel boiloff systems	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	Natural gas release	Critical	Shutoff valves in line, parallel boiloff systems
	Fails open	10^{-3} to 10^{-4}/demand	Cold boiloff gases not warmed, possible brittle failure downstream	Marginal	Parallel boiloff systems, temperature instrumentation
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot bypass heat exchanger	Safe	Parallel boiloff systems

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Line valve, push button controlled (one prior to and one after compressor in each system)	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	Natural gas release	Critical	Manual shut off valves, parallel bioloff systems
	Fails open	10^{-3} to 10^{-4}/demand	Line cannot be blocked from control room	Marginal	Manual shut off valves, parallel boiloff systems
	Fails closed	10^{-3} to 10^{-4} /demand	Boiloff system inoperable due to stopped flow	Safe	Parallel boiloff systems
Temperature indicator - controller-alarm	Fails to operate properly	3×10^{-4} to 3×10^{-6}/hr	Boiloff gas to compressor either too warm or too cold, possible brittle failure downstream	Marginal	Temperature instrumentation after compressor, parallel boiloff systems
Boiloff compressor, multi-stage (one for each system)	Leaks or ruptures	10^{-6} to 10^{-8} /hr	Natural gas release	Critical	Gas detectors/Halon system in building, shut off valves in line, parallel boiloff systems
	Fails to operate	10^{-2} to 10^{-4}/hr	Boiloff gases cannot be sent to pipeline	Marginal	Parallel boiloff systems
Aftercooler (one for each system)	Leaks or ruptures	10^{-4} to 10^{-5}/hr	Natural gas release	Critical	Shutoff valves in line, parallel boiloff systems
	Fails to provide sufficient cooling	10^{-4} to 10^{-5}/hr	Boiloff gases sent to pipeline too warm	Marginal	Downstream temperature instrumentation, parallel boiloff systems
	Fouls or plugs	10^{-4} to 10^{-5}/hr	Cannot handle sufficient volumes of gas, possible overpressure of lines	Marginal	Pressure instrumentation, parallel boiloff systems

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis(cont)

Component	Failure: Mode	Failure: Frequency	Effect	Class	Compensating Provisions
Pressure indicator	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Operator misinformed, may not shut system down in off-standard conditions, possible overpressure of lines	Marginal	
Manual shutoff valves in boiloff lines	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	Natural gas release	Critical	Other shutoff valves, parallel boiloff systems
	Fails open	10^{-6} to 10^{-7}/demand	Cannot isolate parts of system	Safe	Other shutoff valves
	Fails closed	$3x10^{-4}$ to $3x10^{-5}$/demand	No flow through system	Safe	Parallel boiloff systems
Temperature indicator following aftercooler	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Boiloff gas to pipeline too hot or too cold	Marginal	High temperature alarm downstream, temperature indicator prior to compressor
Compressor recycle valve (air operated, controlled by pressure indicator - controller)	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	Natural gas release	Critical	Shutoff valves, parallel boiloff systems
	Fails open	10^{-3} to 10^{-4}/demand	Cannot control recycle	Marginal	Parallel boiloff systems
	Fails closed	10^{-3} to 10^{-4} /demand	No compressor recycle, possible insufficient sendout pressure	Safe	Parallel boiloff systems
Pressure indicator - controller	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Recycle not controlled properly, possible insufficient sendout pressure or overpressure of lines	Marginal	Parallel boiloff systems, pressure indicator after compressor

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure: Mode	Failure: Frequency	Effect	Class	Compensating Provisions
Check valve in boiloff sendout line	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	Natural gas release	Critical	ESD system
	Fails open	10^{-6} to 10^{-7}/hr	Possible backflow of gas from pipeline	Marginal	Two check valves in series in line, ESD system
	Fails closed	$3x10^{-4}$ to $3x10^{-5}$/demand	Boiloff gases cannot be sent to pipeline	Marginal	Parallel boiloff systems
High temperature alarm on sendout line	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Boiloff gases sent to pipeline too warm	Marginal	Temperature indicator following aftercooler
Flow recorder on sendout line	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Inaccurate measurement of gas sent out	Safe	
Pressure recorder on sendout line	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Possible overpressure of sendout line	Marginal	Pressure instrumentation/ control on pressure loop
12" expansion joint on LNG sendout line	Leaks or ruptures	10^{-5} to 10^{-8}/hr	LNG release	Critical	Spill basin, block valve in tank
12" block valve in tank on LNG sendout line	Leaks or fails open	10^{-3} to 10^{-4}/demand	Cannot close off send-out line	Marginal	Shutoff valves exterior to tank
	Fails closed	10^{-3} to 10^{-4} /demand	LNG sendout line blocked	Safe	Smaller auxiliary sendout line
Feed valve to LNG sendout pump (one per pump, three total)	Leaks or ruptures	$5x10^{-8}$ to $1x10^{-8}$ /hr	LNG release	Critical	Spill basin, block valve in tank
	Fails open	10^{-6} to 10^{-7}/demand	Cannot isolate pump	Marginal	ESD shutoff valve in line
	Fails closed	$3x10^{-4}$ to $3x10^{-5}$/demand	Cannot send out LNG with associated pump	Safe	Parallel sendout pumps

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
ESD shutoff valve in sendout line	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	LNG release	Critical	Spill basin, block valve in tank
	Fails open	3×10^{-3} to 3×10^{-4}/demand	Cannot shut down send-out line	Marginal	Block valve in tank
	Fails closed	10^{-3} to 10^{-4} /demand	Cannot send out LNG	Marginal	
LNG sendout pump (three in parallel)	Leaks or ruptures	10^{-6} to 10^{-8} /hr	LNG release	Critical	Spill basin, ESD shutoff valve
	Fails to operate	10^{-2} to 10^{-5}/hr	Cannot send out LNG with that pump	Safe	Parallel sendout pumps
Vapor return line from send-out pumps, ∿100' long	Leaks or ruptures	5×10^{-9} to 1×10^{-10} /hr	Natural gas release	Critical	Shutoff valves at pumps and tank
LNG sendout line to vaporizers, ∿500' long	Leaks or ruptures	3×10^{-7} to 5×10^{-10} /hr	LNG release	Critical	ESD system, spill basin
LNG recycle line to tank ∿100' long	Leaks or ruptures	5×10^{-9} to 1×10^{-10} /hr	LNG release	Critical	ESD system, spill basin
LNG flow control valve (two per pump: one in sendout line and one in recycle line)	Leaks or ruptures	5×10^{-8} to 1×10^{-8} /hr	LNG release	Critical	ESD system, spill basin
	Fails open	10^{-3} to 10^{-4}/demand	Unable to control flows	Marginal	Parallel sendout pumps
	Fails closed	10^{-3} to 10^{-4} /demand	Unable to send out or recirculate LNG	Safe	Parallel sendout pumps, manual valve parallel to sendout valve

TABLE A.1 LNG Peakshaving Facility/Storage Section Failure Modes and Effects Analysis (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Flow indicator-controller (one following each pump, three total)	Fails to operate properly	$3x10^{-4}$ to $3x10^{-6}$/hr	Unable to control LNG sendout and/or recycle	Marginal	Parallel sendout pumps
Gas detectors	Fails to detect gas	10^{-5} to 10^{-6}/hr	Activation of ESD delayed	Marginal	
UV flame detector	Fails to detect flame	$2x10^{-4}$ to $3x10^{-6}$ /hr	Activation of ESD delayed	Marginal	
Controlled closure mechanisms on valve	Valve closes too fast	10^{-4} to 10^{-5}/hr	Pressure surge in line from fluid hammer	Critical	Pressure relief valves on lines
	Valve closes too slowly	10^{-4} to 10^{-5}/hr	Time required to isolate system increased	Marginal	
Instrument leads	Signal transport failure	10^{-4} to 10^{-6} /hr	Loss of system control	Marginal	Manual overrides
ESD system	Operator fails to activate	10^{-3} to 10^{-4}/demand	Emergency condition continues	Critical	ESD can be activated from various plant locations by different people
	System fails to activate	10^{-4}/demand	Emergency condition continues	Critical	Manual shutdown capability

TABLE A.2 Failure Modes & Effects Analysis Vaporizer System

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
START-UP OR SHUTDOWN					
LNG Pump Discharge Pressure Setting	High	$3 \times 10^{-3}/d$	• No hazard if piping design adequate for maximum pump pressure. • LNG not fully vaporized. • Outlet low T increases burning rate.	Marginal	• Integrity of piping. • Adequate design for P. • Low outlet T activates ESD.
	Low	$3 \times 10^{-3}/d$	• Low flow through vaporizer. • Outlet temperature increases and burner rate goes to minimum.	Safe	• High T alarm in line. • Low flow alarm in pump outlet. • Low pressure alarm in pump outlet.
Outlet temperature setting	High	$3 \times 10^{-3}/d$	• Burner rate goes to maximum.	Safe	• High T alarm in stack.
	Low	$3 \times 10^{-3}/d$	• LNG not vaporized/possible cryogenic failure downstream. • Burner rate goes to minimum.	Marginal	• Low outlet T activates ESD
Throughput Flow Setting	High or turned up too rapidly.	$3 \times 10^{-3}/d$	• LNG not vaporized/possible cryogenic failure downstream • Burner rate goes to maximum.	Marginal	• Low outlet T activates ESD. • High T alarm in stack. • Large heat storage capacity of bath.
Block Valve Closing Various Locations	LNG trapped between valves.	10^{-2}/demand 1.8×10^{-3} demand	• LNG warms and vaporizes	Critical	• Relief valves between every pair of block valves.

TABLE A.2 Failure Modes & Effects Analysis Vaporizer System (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Tube Bundle	Failure or leak (from thermal shock, corrosion or external cause)	1.25×10^{-4}/hr	• Release of LNG or vaporized LNG into tank. Potential explosive mixture in tank.	Critical	• Combustible gas detector alarm. • Halon fire suppressant
Burner & Downcomers	Flameout	1.4×10^{-3}/hr	• Potential explosive mixture in tank. • LNG not vaporized, possible cryogenic failure downstream.	Marginal	• Burner UV flame detector-alarm. • Thermal storage in water bath. • Auxiliary electric heater. • Low outlet T activates ESD and alarm.
Burner Jacket Water Pump	Failure to start. Failure to run.	1×10^{-3}/demand 3×10^{-5}/hr	• Thermal stream failure of burner.	Marginal	• Low P switch on pump discharge opens valve to admit more water/If low P is due to pump failure, water level will rise and activate level alarm.
Burner Gas Supply	Flameout	1.4×10^{-3}/hr	• See Burner & Downcomers	Marginal	• Low P alarm on manifold.
Burner Air Supply (incl. Air Blower)	Flameout	1.4×10^{-3}/hr	• See Burner & Downcomers	Marginal	• Low P alarm on manifold.
Electric Power to Air Blower & Pumps	Failure	4×10^{-4}/hr	• Flameout. • See Burner & Downcomers.	Marginal	• Auxiliary power supply. • See Burner & Downcomers

TABLE A.2 Failure Modes & Effects Analysis Vaporizer System (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
STEADY STATE OPERATION					
LNG Sendout Pumps Suction Line	Break	$< 7 \times 10^{-10}$/ft-hr.	• Release rate depends on break, driven by hydrostatic head in tank.	Critical	• Low T alarm in spill basin. • ESD closes in-tank valve in tank outlet as well as valve in suction line. • Foam discharge.
LNG Sendout Pumps	Body failure from thermal shock or external cause.	$< 3 \times 10^{-5}$/hr 1×10^{-8}/hr	• Release rate depends on break, driven by hydrostatic head in tank.	Critical	• Low T alarm in spill basin. • Foam discharge. • ESD closes in-tank valve at tank outlet.
	Deadheaded (valve failure, clog or blunder)	$< 5 \times 10^{-8}$/hr for valve failure 3×10^{-4}/hr minor failure 5×10^{-5}/hr valve	• LNG vaporized in pump	Safe	• Vapor vent line from pump to tank. • Increased i on pump opens recycle line.
Transfer Line to Vaporizers	Break	$< 7 \times 10^{-10}$/ft hr $< 3.5 \times 10^{-7}$/hr	• Release rate depends on break, driven by pumps	Critical	• Low discharge P alarms on pumps. • ESD closes pump discharge valve.
Tank and Weir	Failure, complete or partial loss of water	$< 3 \times 10^{-8}$/hr	• Reduced heat transfer, LNG not vaporized, possible cryogenic failure downstream. • Direct firing of tubes leads to tube bundle failure.	Marginal	• Low T outlet activates ESD. • Low water level controller/alarm adds make-up water.

TABLE A.2 Failure Modes & Effects Analysis Vaporizer System (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Water Supply	Pressure loss.	4×10^{-4}/hr	• Loss of heat transfer, LNG not vaporized. • Direct firing of tubes leads to tube bundle failure. • Pump not able to supply burner jacket leading to failure of burner.	Marginal	• Low level alarm. • Low outlet T activates ESD and alarm. • High T alarm in stack.
Overflow Line	Back pressure due to plugging.	6×10^{-5}/hr	• Tank overflow. • Loss of circulation effect causing tube bundle overheating. • Extinguishing flame/ release of combustible gas. • Tank or weir failure.	Marginal	• High level alarm.
LNG Flow Control Valve	Fails open.	5×10^{-8}/hr 8.5×10^{-6}/ hr	• LNG not completely vaporized.		• Low outlet T activates ESD and alarm.
Burner Gas and Air Control Valves	Fails open.	5×10^{-5}/hr 8.5×10^{-6}/hr	• LNG overheated.		• High T stack alarm.
	Fails closed.	5×10^{-5}/hr 8.5×10^{-6}/hr	• Flameout/LNG not vaporized.		• Low outlet T activates ESD and alarm. • Burner UV Flame detector-alarm. • Thermal storage in water bath. • Auxiliary electric heater.

TABLE A.2 Failure Modes & Effects Analysis Vaporizer System (cont)

Component	Failure Mode	Failure Frequency	Effect	Class	Compensating Provisions
Vapor Outlet Line	Major Leak	3.3×10^{-6}/hr	• Release of natural gas. • Depressurization causing LNG surge through vaporizer. • LNG not vaporized - possible cryogenic failure.		• ESD manually activated. • Low outlet T activates ESD and alarm. • Low discharge P on sendout pumps.

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System

TRUCK LOADING

Subsystem or Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
LNG line from storage tank to loading station	Ruptures while loading	$1\text{x}10^{-10}$/section	Release of LNG	C	• Piping standards, operator inspection • DIKE[a], LAVB[b], WC[c], WM[d], ESD[e], TVM[f]
Expansion joints	Rupture while loading	$1\text{x}10^{-7}$	Release of LNG or LNG vapor	C	• Piping standards, operator inspection • DIKE, LAVB, WC, WM, ESD, TVM
Vapor return line from loading station to storage tank	Rupture while loading	$1\text{x}10^{-10}$/section	Release of LNG vapor	M	• Piping standards, operator inspection • TVM, ESD • Vapor return valves
Loading station fill valve	Rupture	$1\text{x}10^{-8}$	Release of LNG	C	• Valve maintenance & inspection of stream valve • LAVB and C[g], INC, WM
	Left open after last load out	$1\text{x}10^{-2}$	Release of LNG when upstream valve is opened	C	• Same as above
Transfer pump	Rupture	$1\text{x}10^{-8}$	Release of LNG	C	• Operator inspection, upstream valve • Indicator lights • LAVB, WC, WM, EXT[h]
Flexible loading hose	Rupture while loading	$1.7\text{x}10^{-6}$	Release of LNG	C	• High pressure hose standards • Preloading inspection/loading valve • ESD, LAVB and C, WC, WM
Grounding cables	Bad condition	$3\text{x}10^{-7}$	None unless spark causes explosion and fire	M	• Preloading inspection • WC, WM, EXT

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System (con't)

Subsystem or Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
	Failure to hook up	$1x10^{-2}$	None unless spark causes explosion and fire	M	• Same as above
Tank atmosphere	Oxygen present		Possible explosive mixture, explosion and fire	M	• Preloading inspection • WC, WM, EXT
Trailer road safety valve	Failure to close before loading	$1x10^{-2}$	Possible release of safety valve and LNG or natural gas	M	• Loading procedures/loading valves • ESD, LAVB and C, WC, WM
	Failure to open before release of truck after loading	$1x10^{-2}$	Possible over-pressurization of truck tank, possible rupture or leak	M	• Postloading procedure/pressure of gauge
Fill trycock valves	Failure to open as indicators while loading	$1x10^{-2}$	Possible overfilling of tank	M	• Loading procedures/scale readings or liquid level gauge
Vapor return valves on truck and at loading station	Rupture	$1x10^{-9}$	Release of natural gas	C	• Valve maintenance & inspection • ESD
	Leak	$1x10^{-8}$	Same as above	C	• Same as above
	Left closed during loading	$1x10^{-2}$	Pressurization of truck tank, possible leak or rupture	M	• Loading procedure/ESD, LAVB and C, WC, WM
Fill valve on truck	Rupture	$1x10^{-9}$	Release of LNG or natural gas	C	• Valve maintenance & inspection/ truck loading station fill valve • ESD, LAVC and C, WC, WM

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System (con't)

Subsystem or Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
	Leak	1×10^{-7}	Same as above	C	• Same as above
	Fails to close completely	5×10^{-5}	Leak of LNG when flexible hose is removed	M	• Valve maintenance & inspection/ sump
Drain valves on flexible hose	Rupture	1×10^{-9}	Release of LNG or natural gas	C	• Valve maintenance & inspection/ ESD, loading valves • LAVB and C, WC, WM
	Leak	1×10^{-8}	Same as above	C	• Same as above
	Left open after last loading	1×10^{-2}	Release of LNG or natural gas when loading	C	• Preloading inspection/ESD loading valves • LAVB and C, WC, WM
Sump	Leaks	1×10^{-7}	Release of LNG or natural gas	C	• Maintenance and inspection • LAVB and C, WC, WM
Relief valves	Fail to relieve pressure	5×10^{-5}	Overpressurization, possible rupture or leak	M	• Inspection and testing 1 ESD, loading valves • LAVB and C

(a) Dike - Diked impoundment area
(b) LAVB - Loading area vapor barrier
(c) INC - Water curtain
(d) WM - Water monitor
(e) ESD - Emergency shutdown system
(f) TVM - Closed circuit television monitor
(g) LAVB and C - Loading area vapor barrier and channels
(h) EXT - Fire extinguishers

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System (con't)

NORMAL TRANSPORTATION

Subsystem or Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
Inner tank shell	Metal fatigue from cooldown/heatup cycles resulting in weakened/cracked tank		Release of LNG	C	Inspection, use of appropriate metal
	Use for an incompatible commodity, resulting in tank corrosion and possible cracking		Release of LNG	C	DOT regulations, inspection, shipper license, special tank fittings
	Use for an incompatible commodity, possibly forming explosive mixture with LNG		Tank explosion release of LNG	C	DOT regulations, inspection, shipper license, special tank fittings
	Rollover (of LNG)		Rapid increase in vaporization of LNG, possible failure of inner tank	M	Relief valves, vent system
	Fails		Major release of LNG	C	Flatplate safety valves, outer shell slows release
Outer tankshell	Defective (corroded)		Lowered failure threshold. Potential heat leaks	M	Flatplate safety valves, inspection

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System (con't)

Subsystem of Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
	Fails		Loss of insulation. Increase in vaporization of LNG. Possible failure of inner tank	M	Relief valves, vent system
Main liquid line	Rupture	$1x10^{-10}$/section	Release of LNG	C	Manual & remote shut-off valves
Hose connection (main liquid line)	Defective (loose seat)	$1x10^{-8}$	Release of LNG	C	Manual shut-off valve Emergency shut-off valve
Manual shut-off valve (Main liquid line)	Rupture	$1x10^{-9}$	Release of LNG	C	Emergency shut-off valve
	Leak	$1x10^{-8}$	Release of LNG	C	Same as above
Emergency shut-off valve (main liquid line)	Rupture	$1x10^{-9}$	Release of LNG	C	Manual shut-off valve, line in tact
	Leak	$1x10^{-7}$	Release of LNG	C	same as above
Pressure build line	Rupture	$1x10^{-10}$/section	Release of LNG	C	Manual & remote shut-off valve
Manual shut-off valve	Rupture	$1x10^{-9}$	Release of LNG	C	Automatic shut-off valve
	Leak	$1x10^{-8}$	Release of LNG	C	Same as above
Automatic shut-off valve (pressure build line)	Rupture	$1x10^{-9}$	Release of LNG	C	Manual shut-off valve closed, line intact

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System (con't)

Subsystem of Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
	Leak	1×10^{-8}	Release of LNG	C	Same as above
Automatic pressure build regulator	Fails open, while isolation valve open	5×10^{-5}	Pressure increases in tank and over-pressure system	M	Relief valves, vent systems
	Fails closed, while isolation valve open	5×10^{-5}	Pressure drops in tank not compensated for. LNG heats up. Possible failure of inner tank.	M	Insulation, from pressure gage
Vent line	Rupture	1×10^{-10}/section	Release of LNG or natural gas	C	Inspection procedures
Main safety valve	Rupture	1×10^{-9}	Tank over pressured and fails, release of LNG	C	Burst disc safety valve (up to 105 psig)
	Leak	1×10^{-8}	Same as above	C	Same as above

TABLE A.3 Failure Modes & Effects Analysis, Transportation & Transfer System (con't)

TRANSPORTATION ACCIDENT

Subsystem or Component	Failure Mode	Failure Rate (Events/hr)	Effect on System	Class	Compensating Provisions
LNG semi-trailer	Accident	$5x10^{-7}$/mile(*)			Double-walled tank resistant to failure
	Overturn				Driver training, double-walled tank
Cargo tank	Accident fails both tanks catastrophically	0.01/accident(**)	Release of entire tank contents instantaneously	C	Double-walled tank resistant to failure
	Impact fails outer shell, loss of insulation		LNG heat-up, possible failure of inner tank	M	Relief valves, vent system
	Accident causes serious spill	0.05/accident(**)	Release of 10% of tank instantaneously or entire tank in 5 minutes	C	Double-walled tank
	Accident puntures shell or causes small valve leak	0.2/accident(**)	Maximum release of 10% of tank contents	M	Valve placement, double-walled tank
	Fire fails shell		LNG heats up and over-pressurizes tank	C	Heat resistance of insulation, relief valves/vent system
Driver	Untrained, unaware of high center of gravity. Mishandles and overturns truck		Possible tank failure and release of LNG	C	Training (DOT & company)
Valve controls	Rear impact damages valve controls & piping		Release of LNG	C	Judicious valve placement, enclosed in compartment

* Failure rate given in accidents per mile.
** Probability of a spill given an accident has occurred (A. D. Little, Inc. 1978)(conditional spill probability)

APPENDIX B

FAULT TREE ANALYSIS

AND SUPPORTING CALCULATIONS

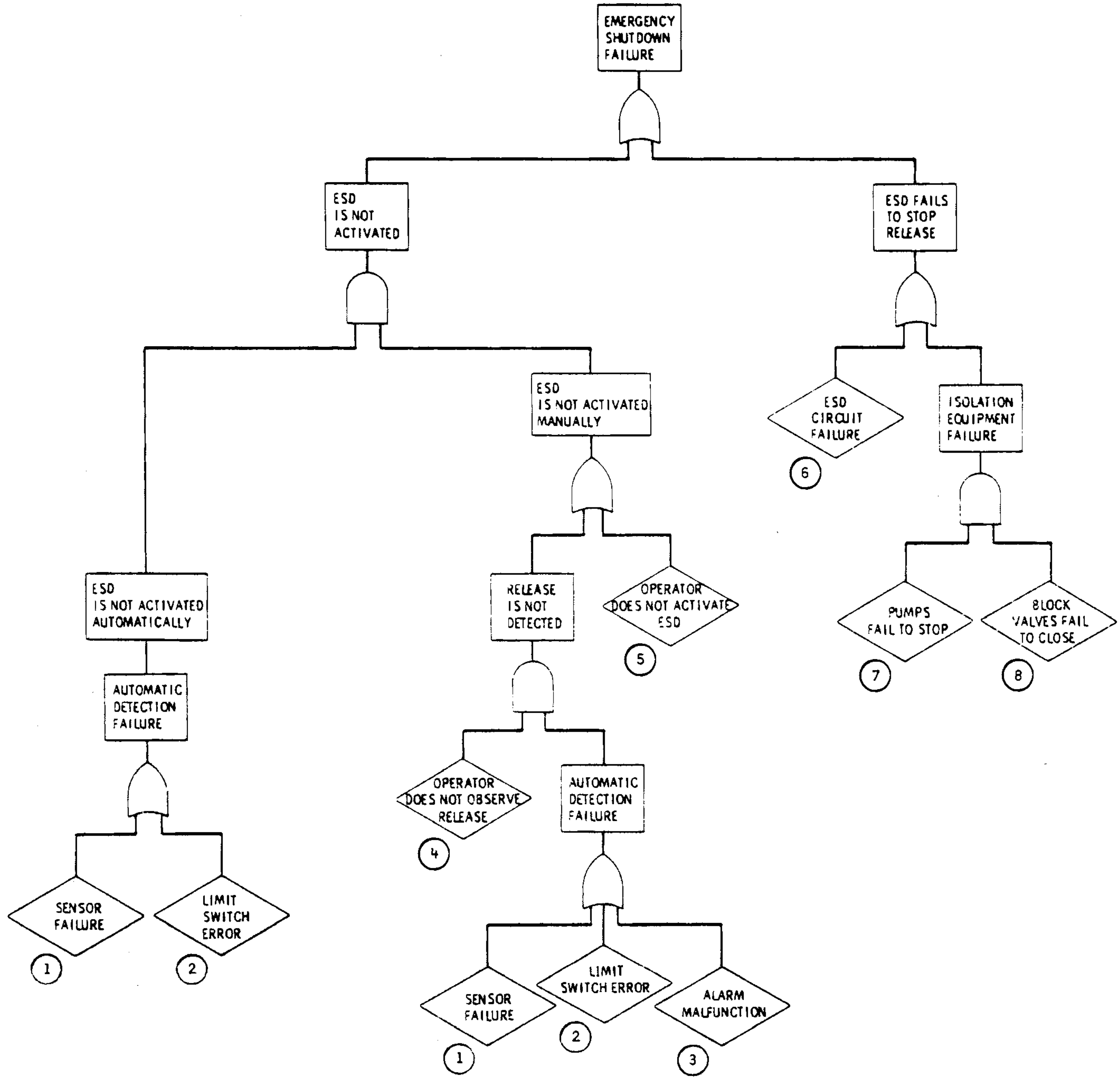


FIGURE B.1 Fault Tree Failure of a Fully Automatic Emergency Shutdown (Supporting Calculations in Table B.1)

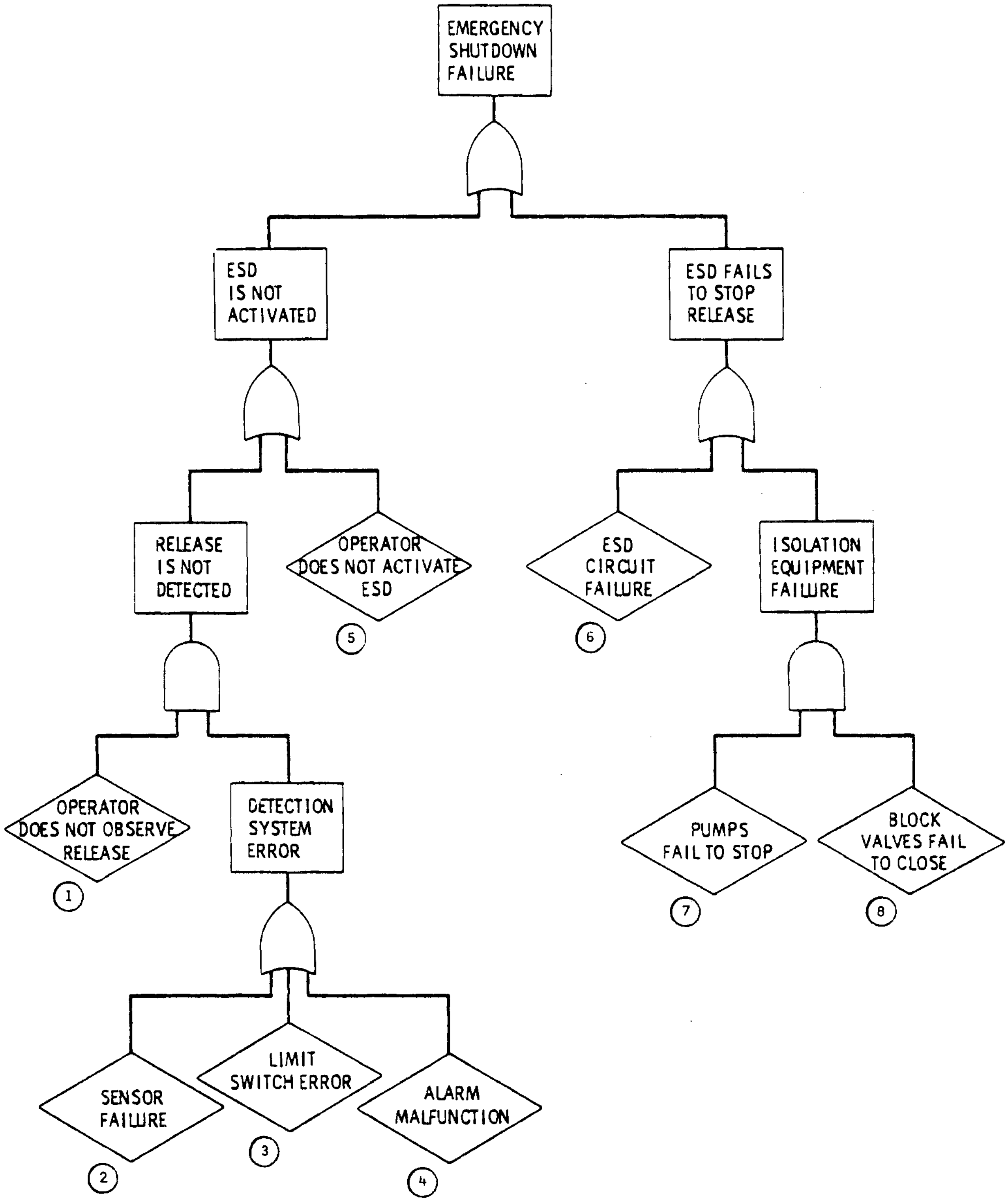


FIGURE B.2 Fault Tree for Failure of an Automatic or Manual Detection, Manual Activation Emergency Shutdown (Supporting Calculations in Table B.2)

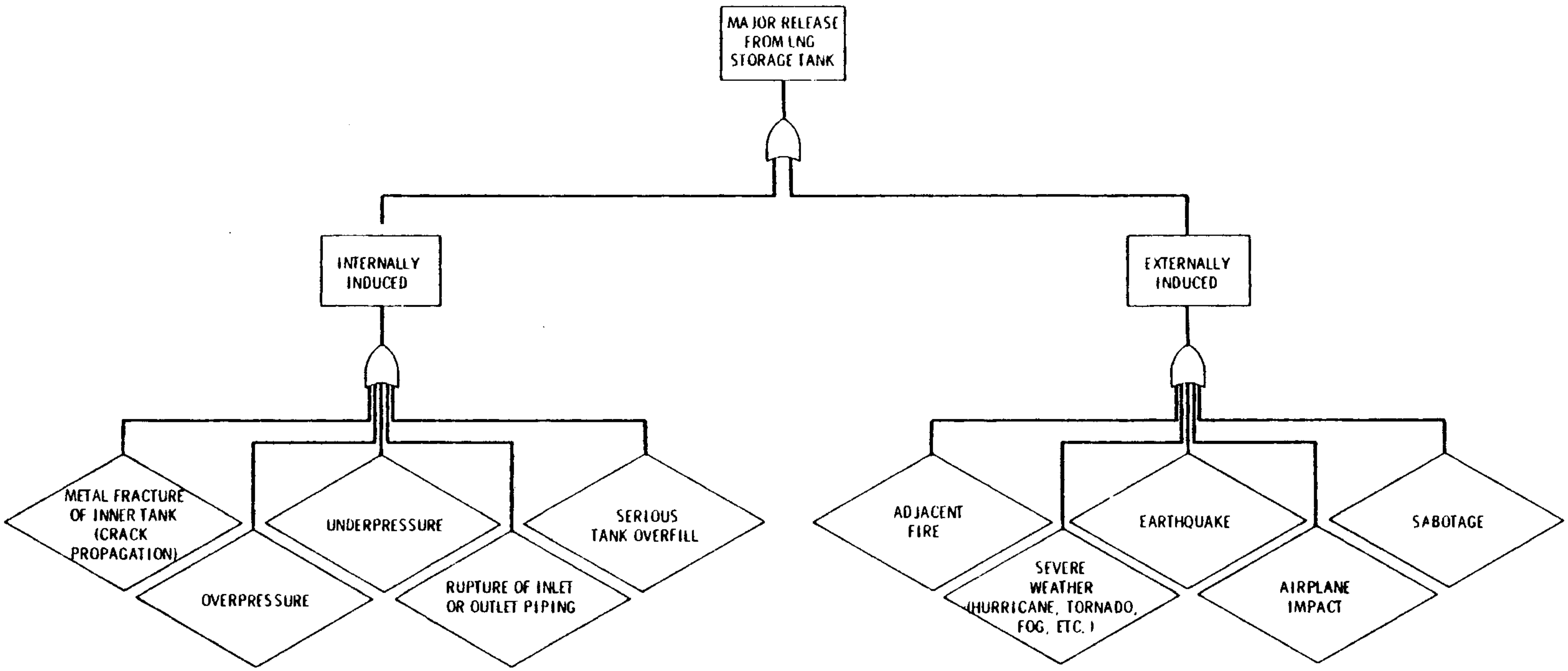


FIGURE B.3 Fault Tree for Large Release from an LNG Storage Tank

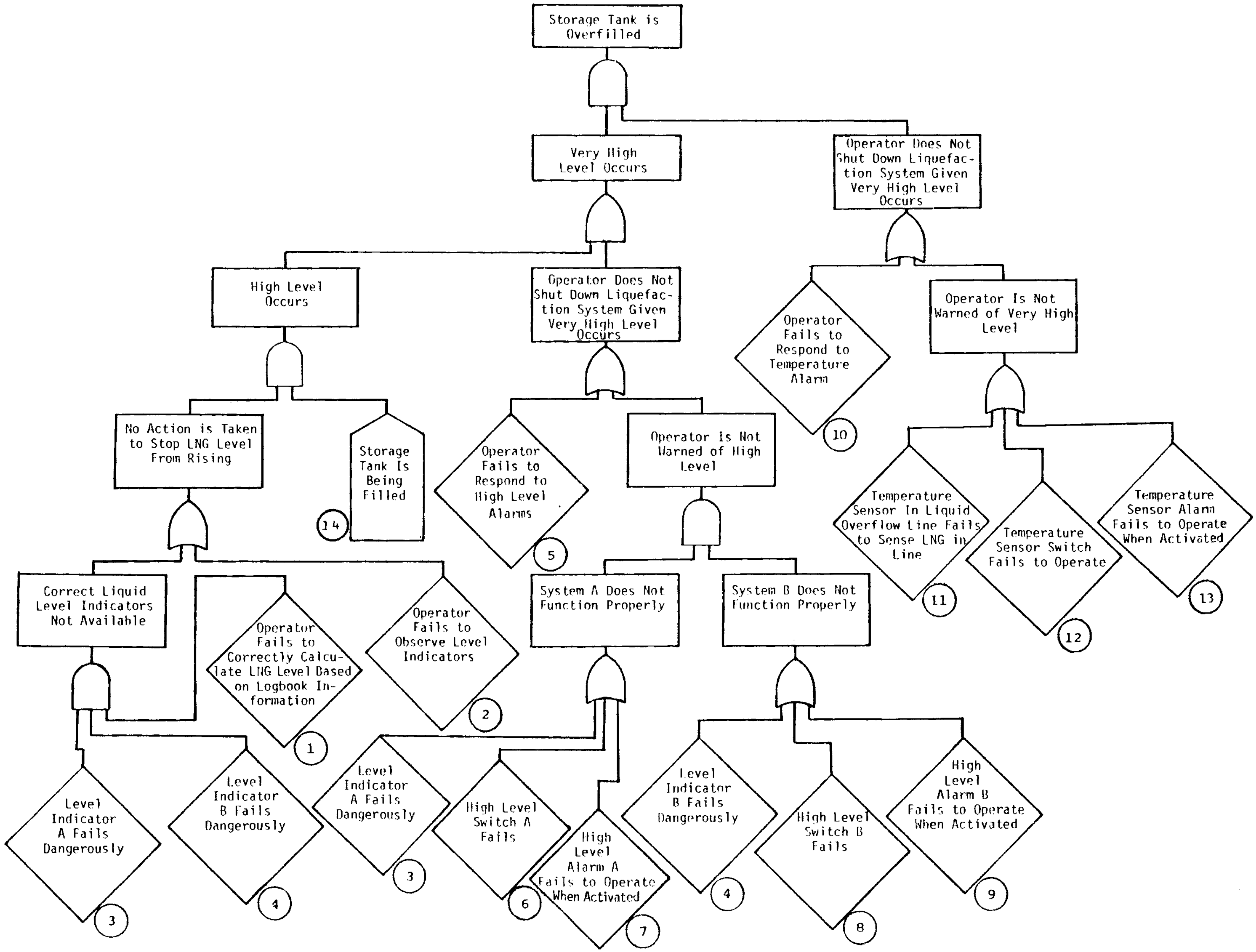


FIGURE B.4 Fault Tree for Overfilling an LNG Storage Tank (Supporting Calculations in Table B.3)

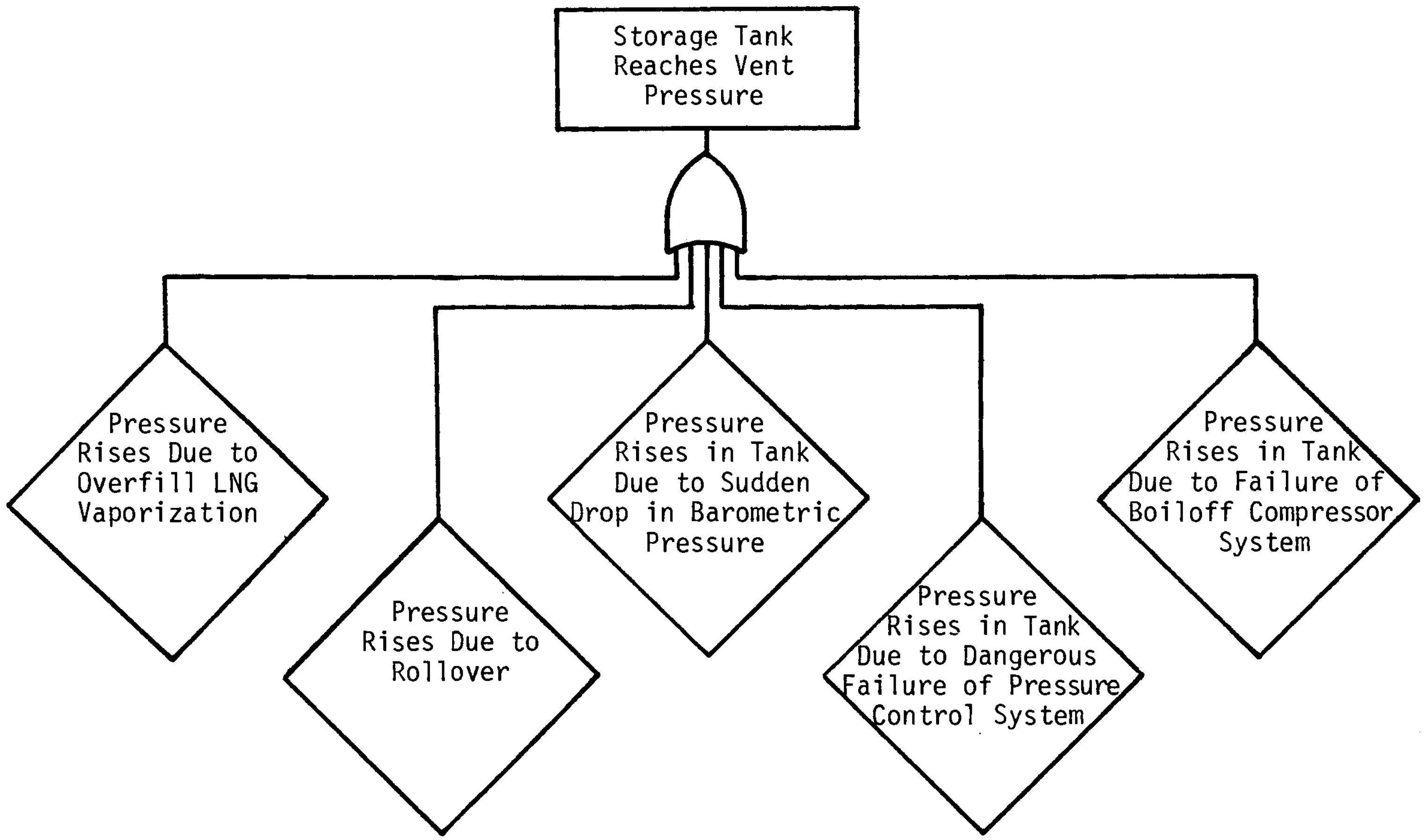


FIGURE B.5 Fault Tree for an LNG Storage Tank Reaches Vent Pressure (Supporting Calculations in Table B.4)

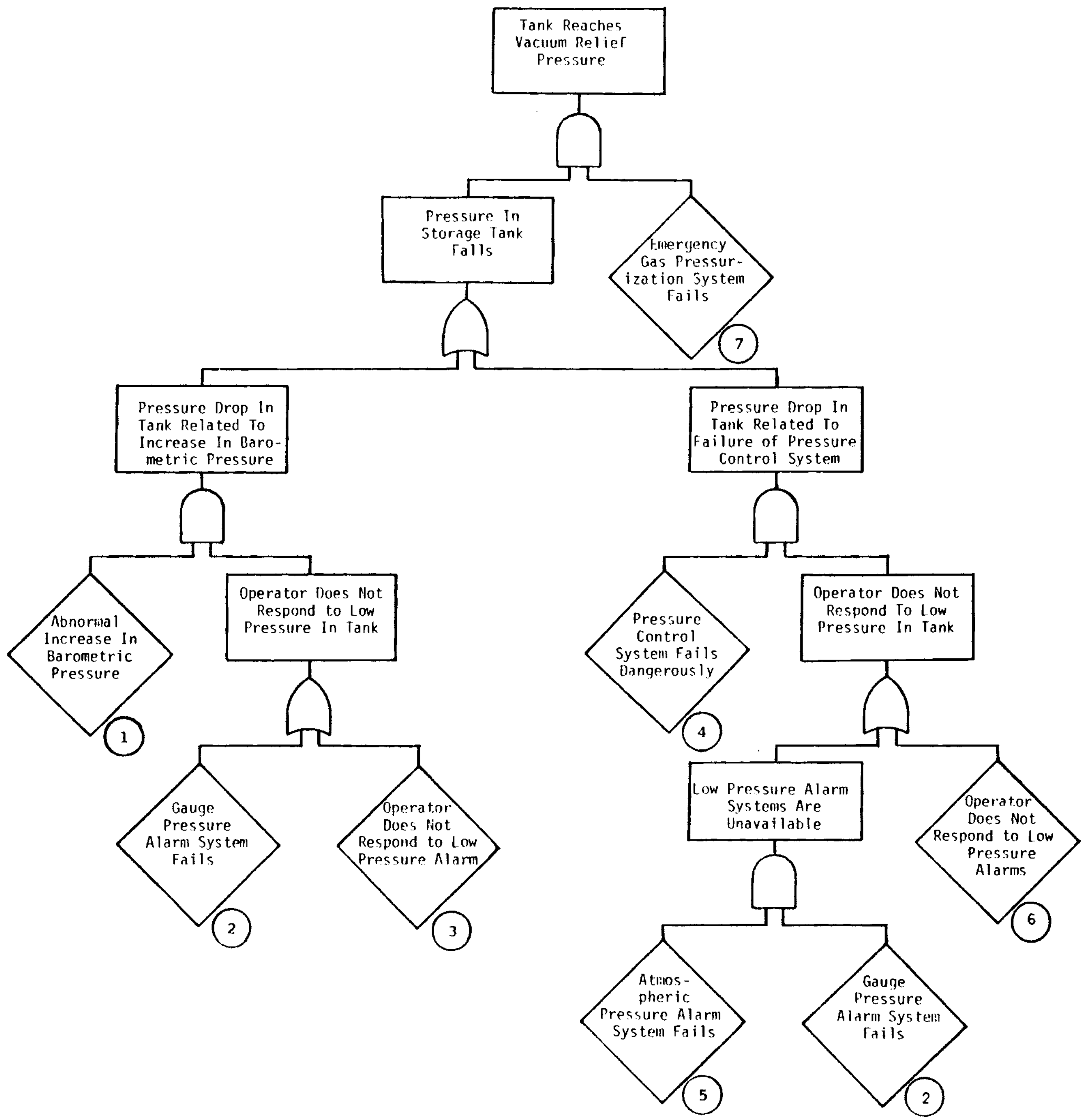


FIGURE B.6 Fault Tree for an LNG Storage Tank Reaches Vacuum Relief Pressure (Supporting Calculations in Table B.5)

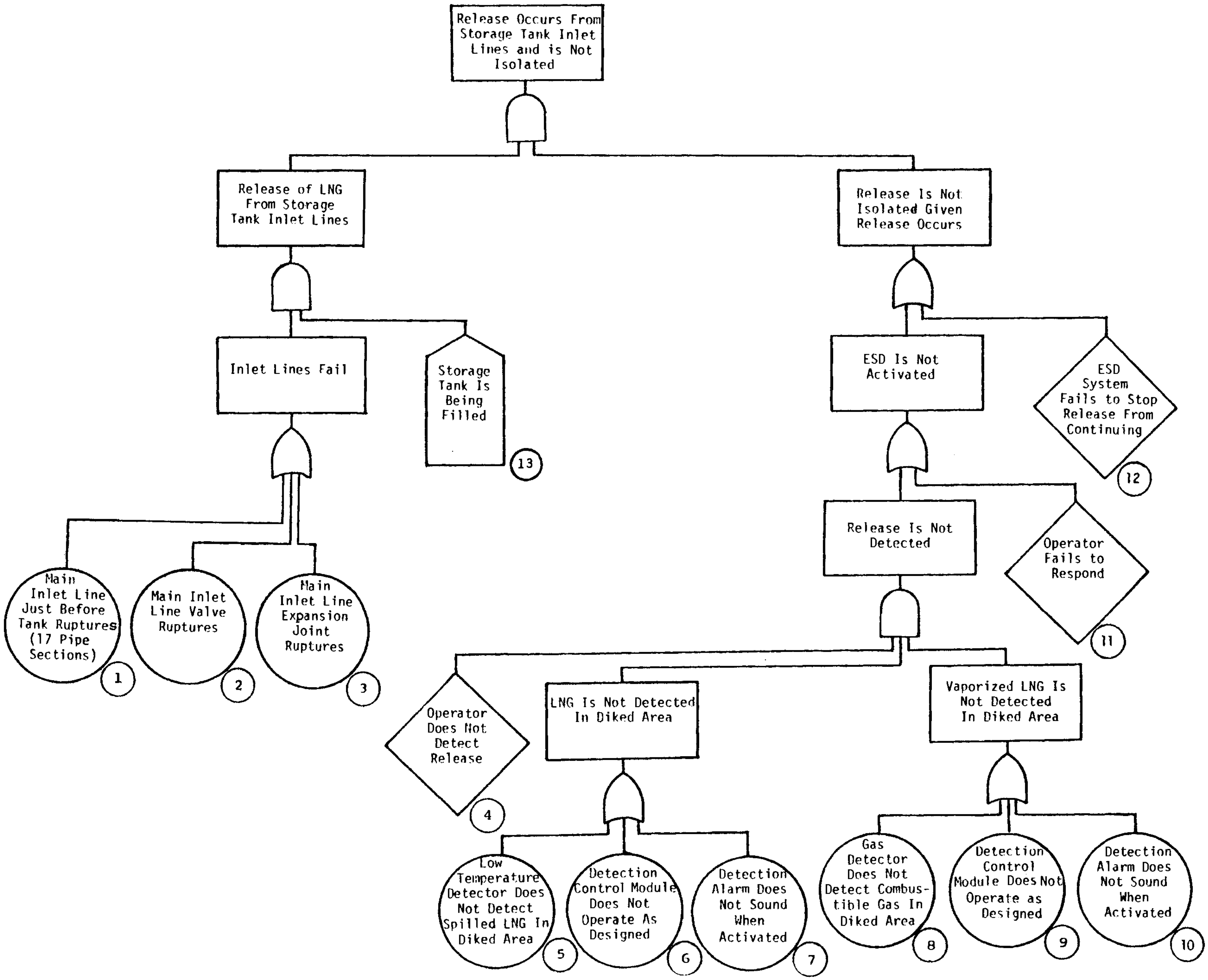


FIGURE B.7 Fault Tree for Large Release from Storage Tank Inlet Line (Supporting Calculations in Table B.6)

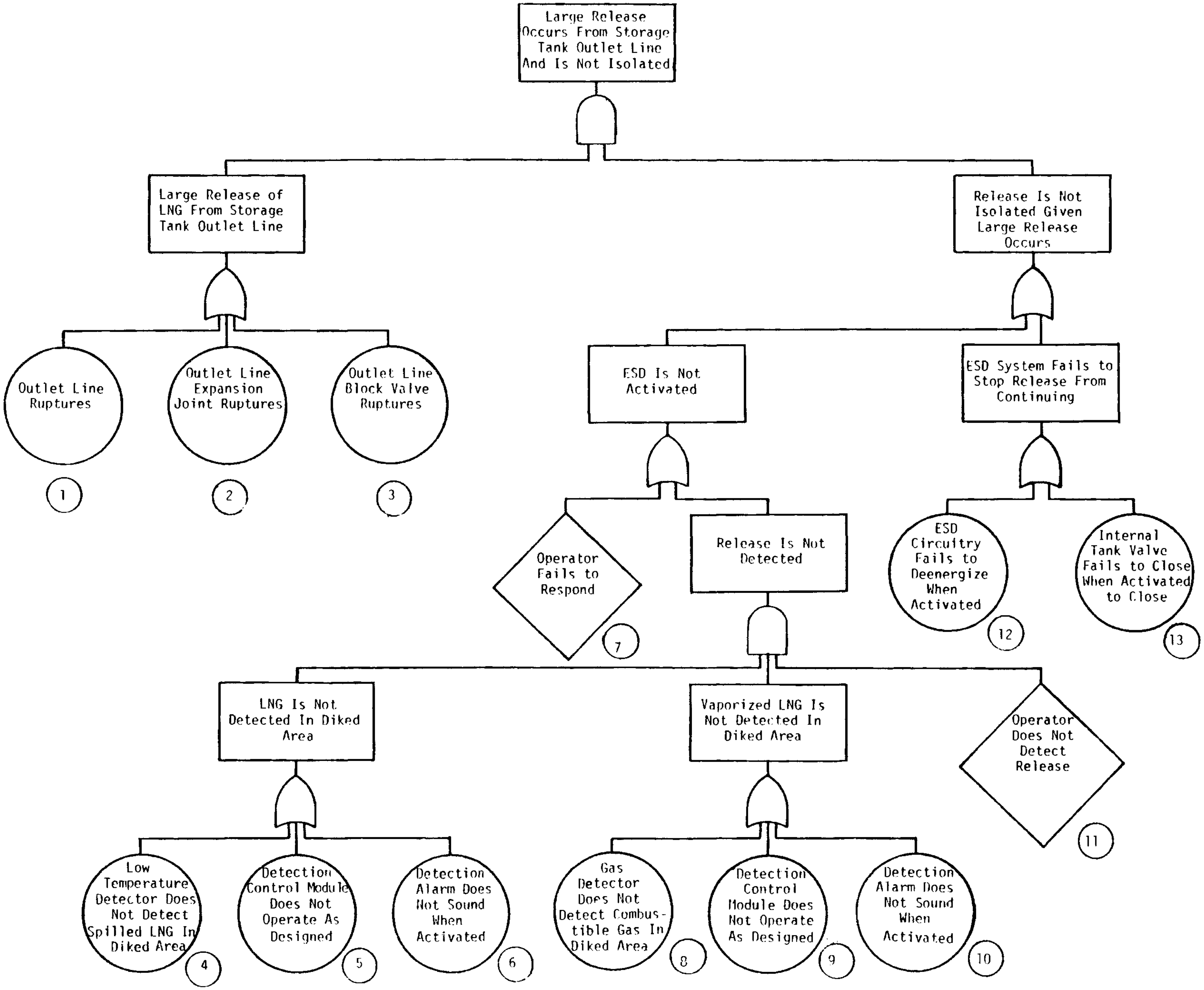


FIGURE B.8 Fault Tree for a Large Release from the Storage Tank Outlet Line (Supporting Calculations in Table B.7)

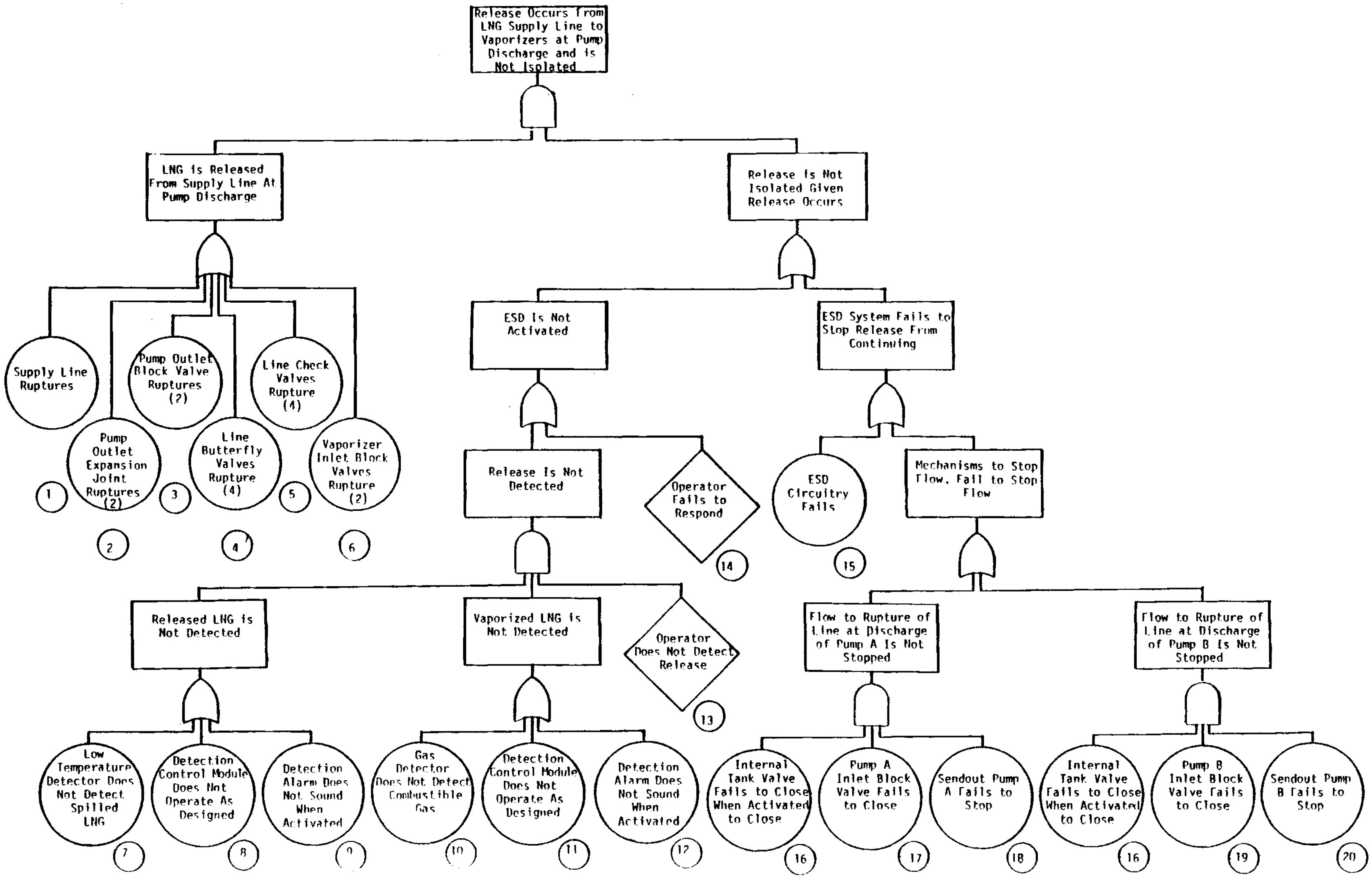


FIGURE B.9 Fault Tree for a Large Release from the Pump Discharge Line (Supporting Calculations in Table B.8)

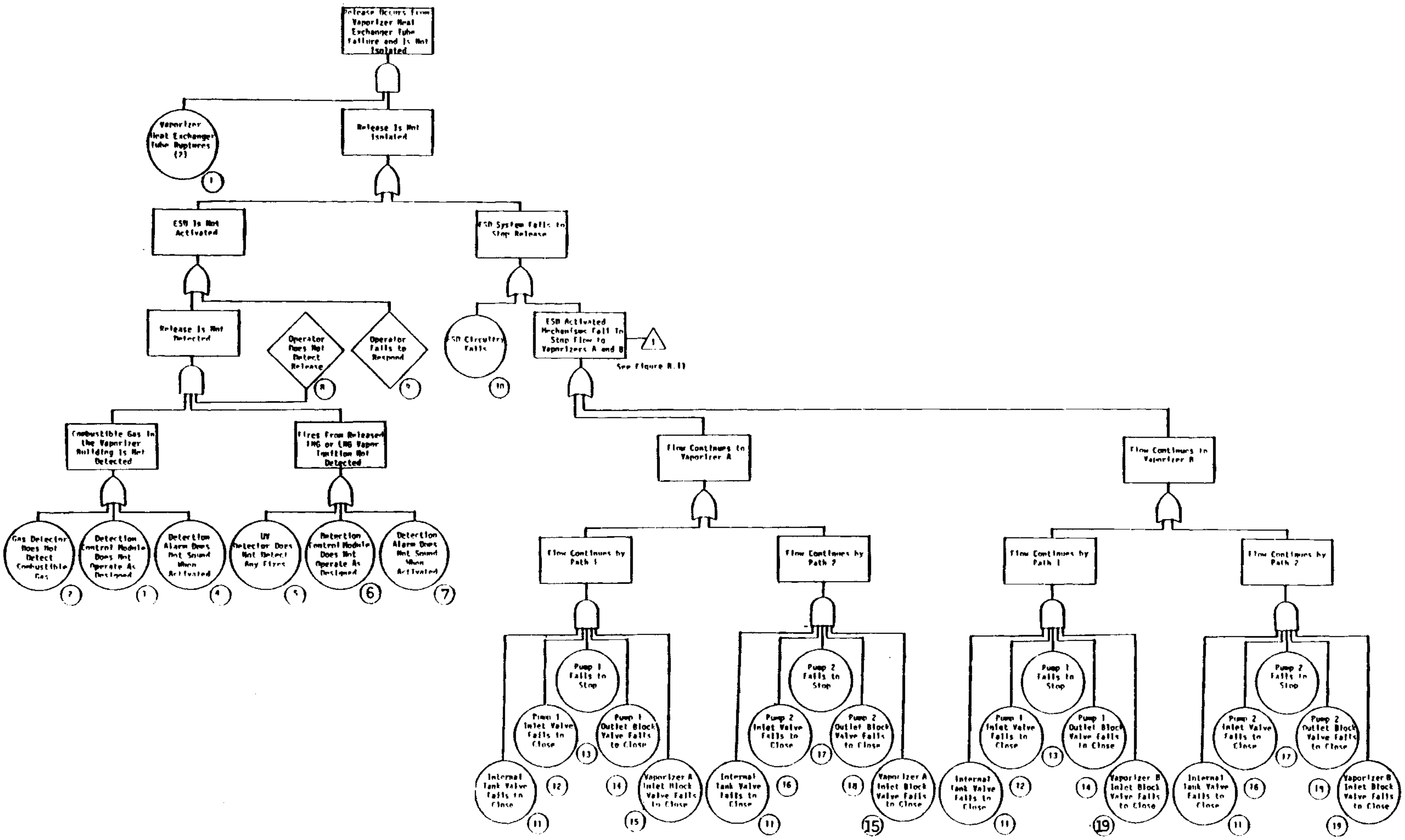


Figure B.10 Fault Tree for a Large Release from a Vaporizer Heat Exchanger Tube (Supporting Calculations in Table B.9)

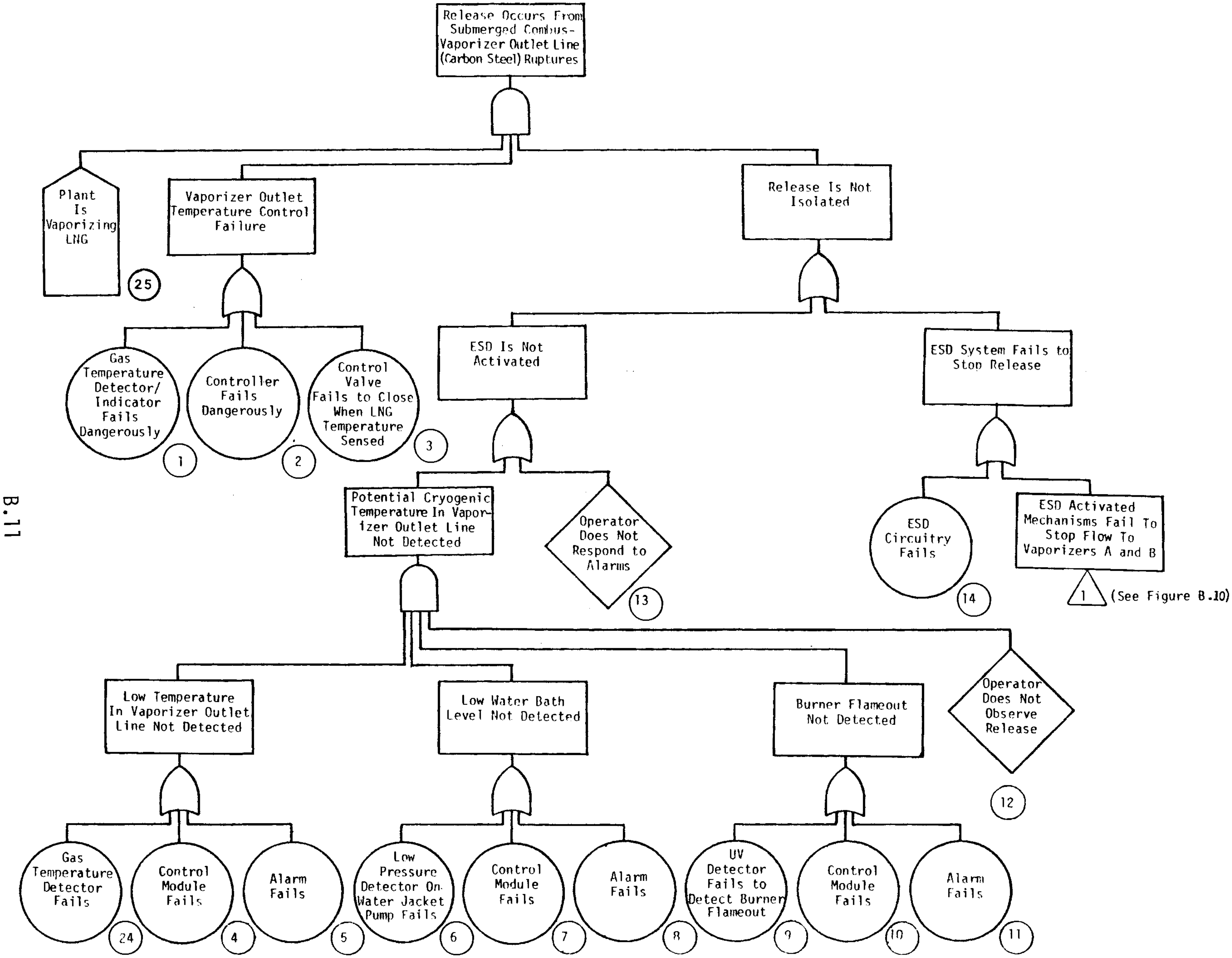


FIGURE B.11 Fault Tree for a Large Release from the Vaporizer Outlet Line (Supporting Calculations in Table B.10)

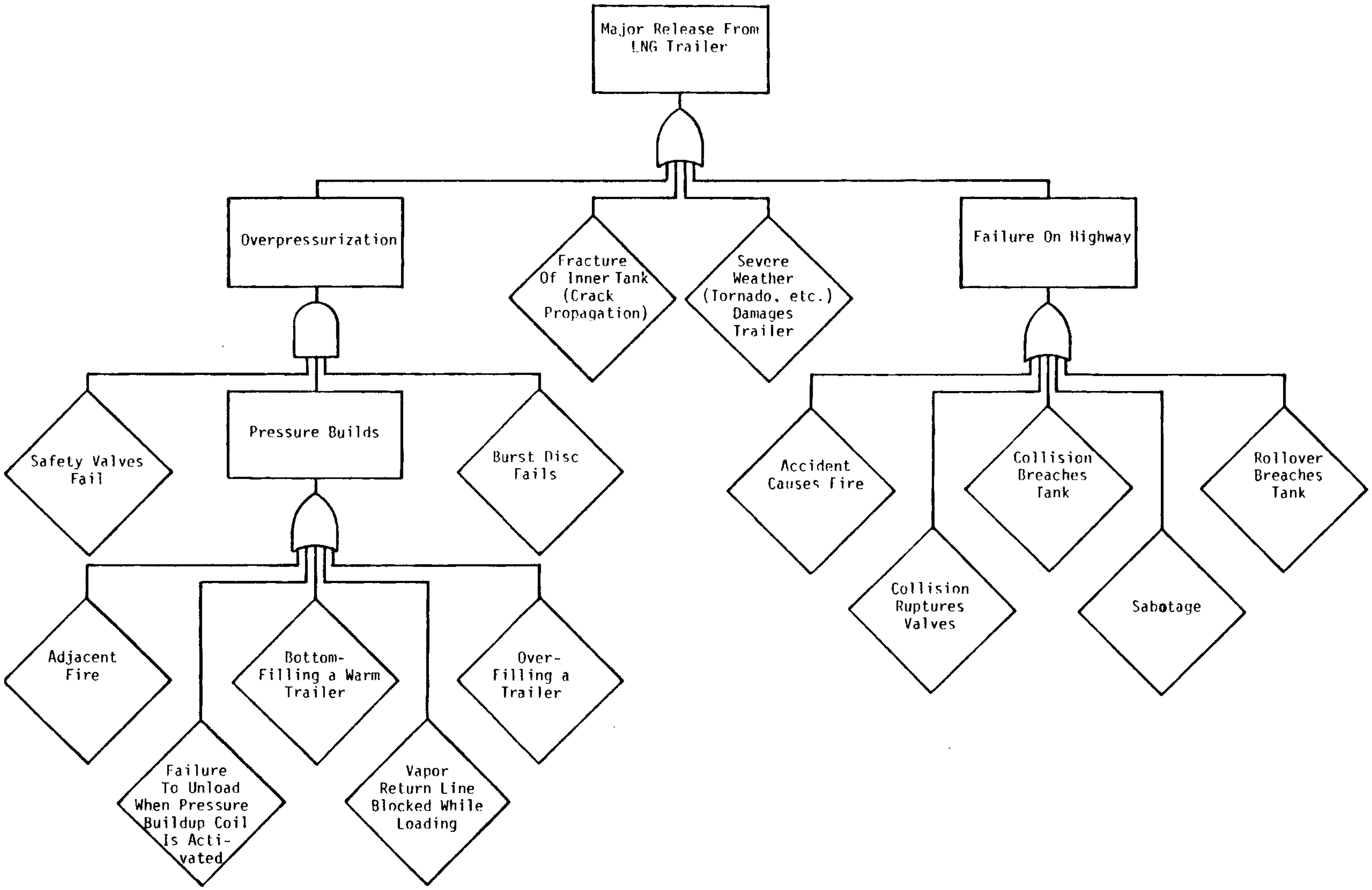


FIGURE B.12 Fault Tree for Complete Failure of an LNG Semi-Trailer

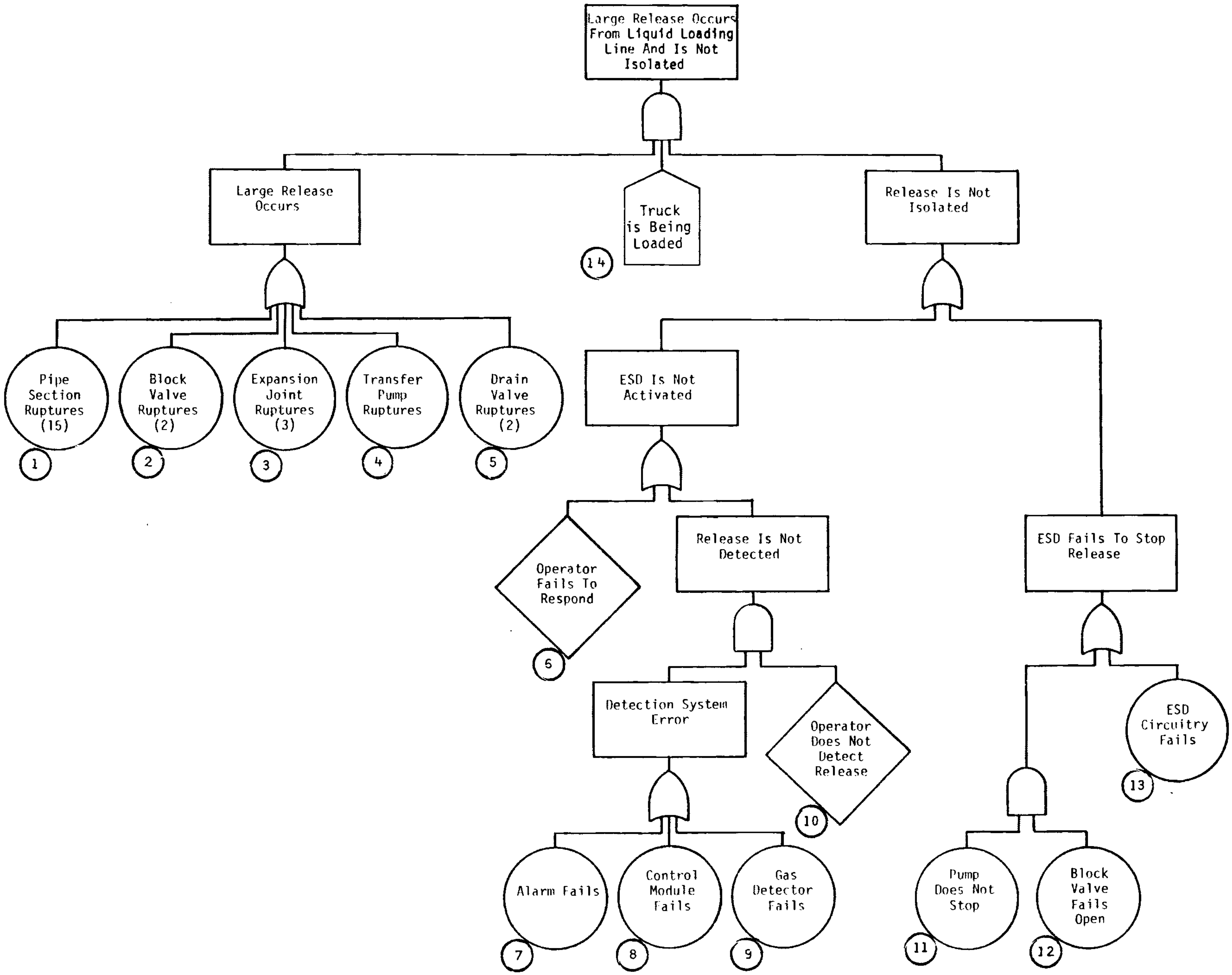


FIGURE B.13 Fault Tree for Failure of the Truck Loading Line (Supporting Calculations in Table B.11)

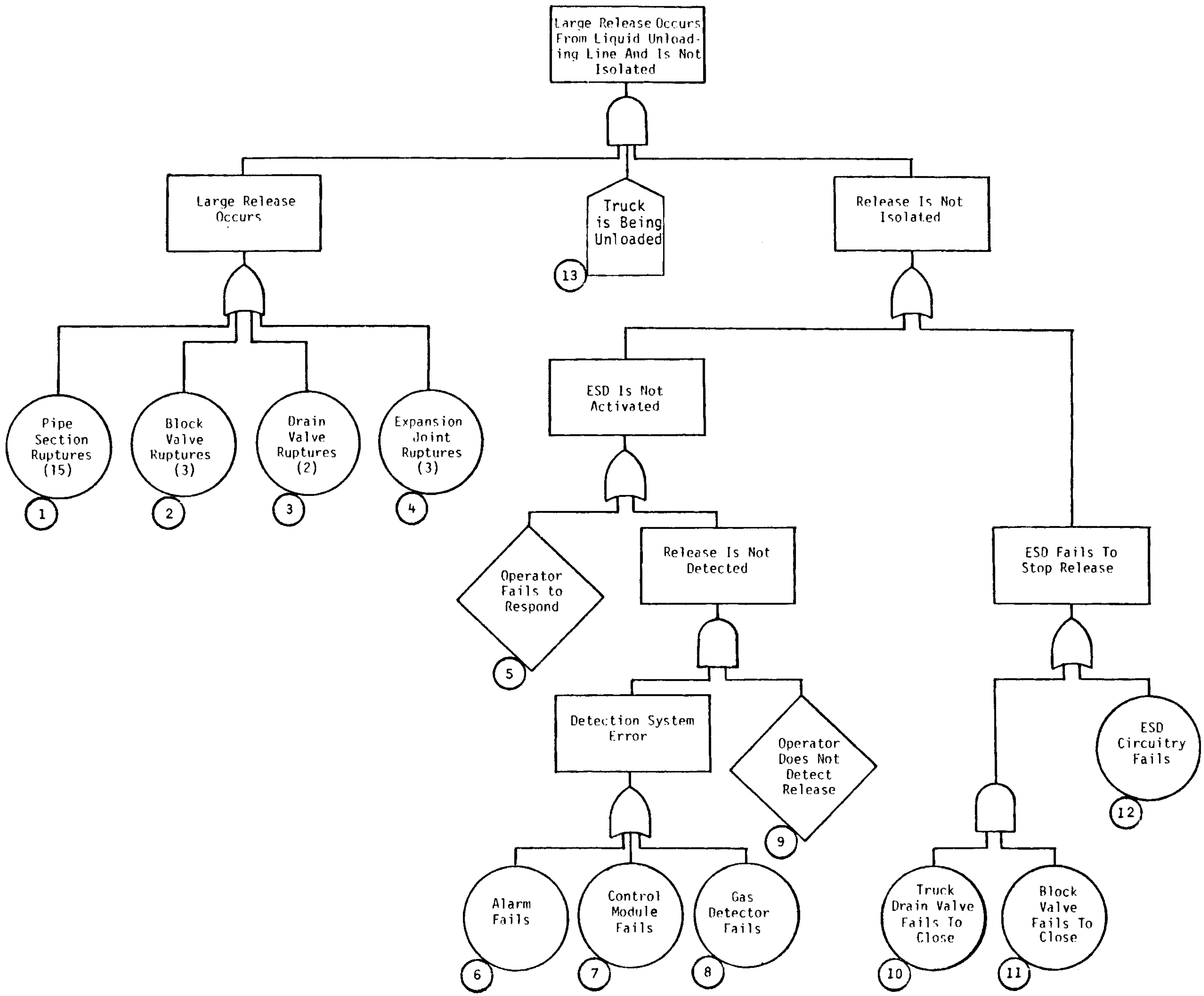


FIGURE B.14 Fault Tree for Failure of the Truck Unloading Line. (Supporting Calculations in Table B.12)

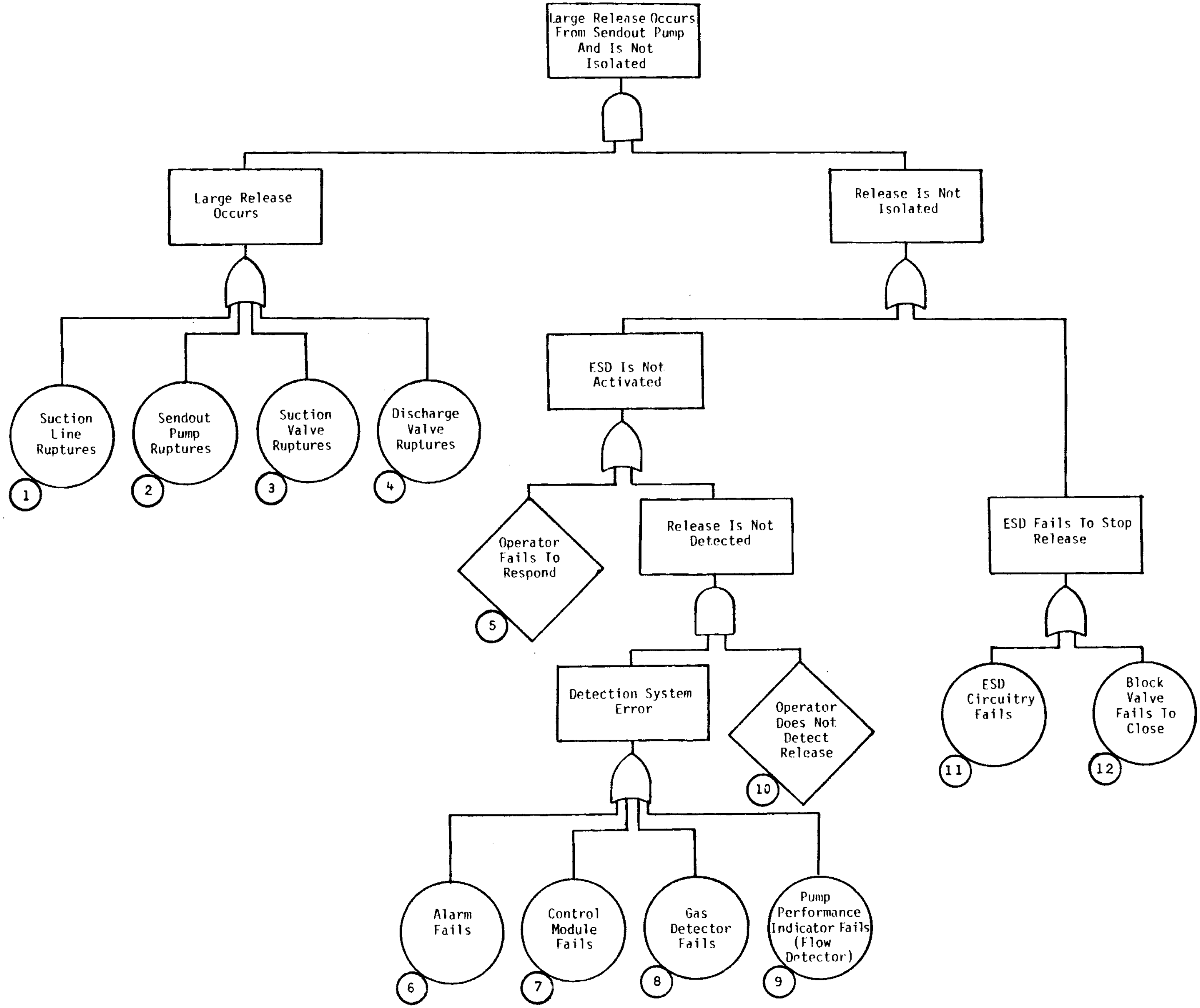


FIGURE B.15 Fault Tree for Failure of the LNG Sendout Pump. (Supporting Calculations in Table B.13)

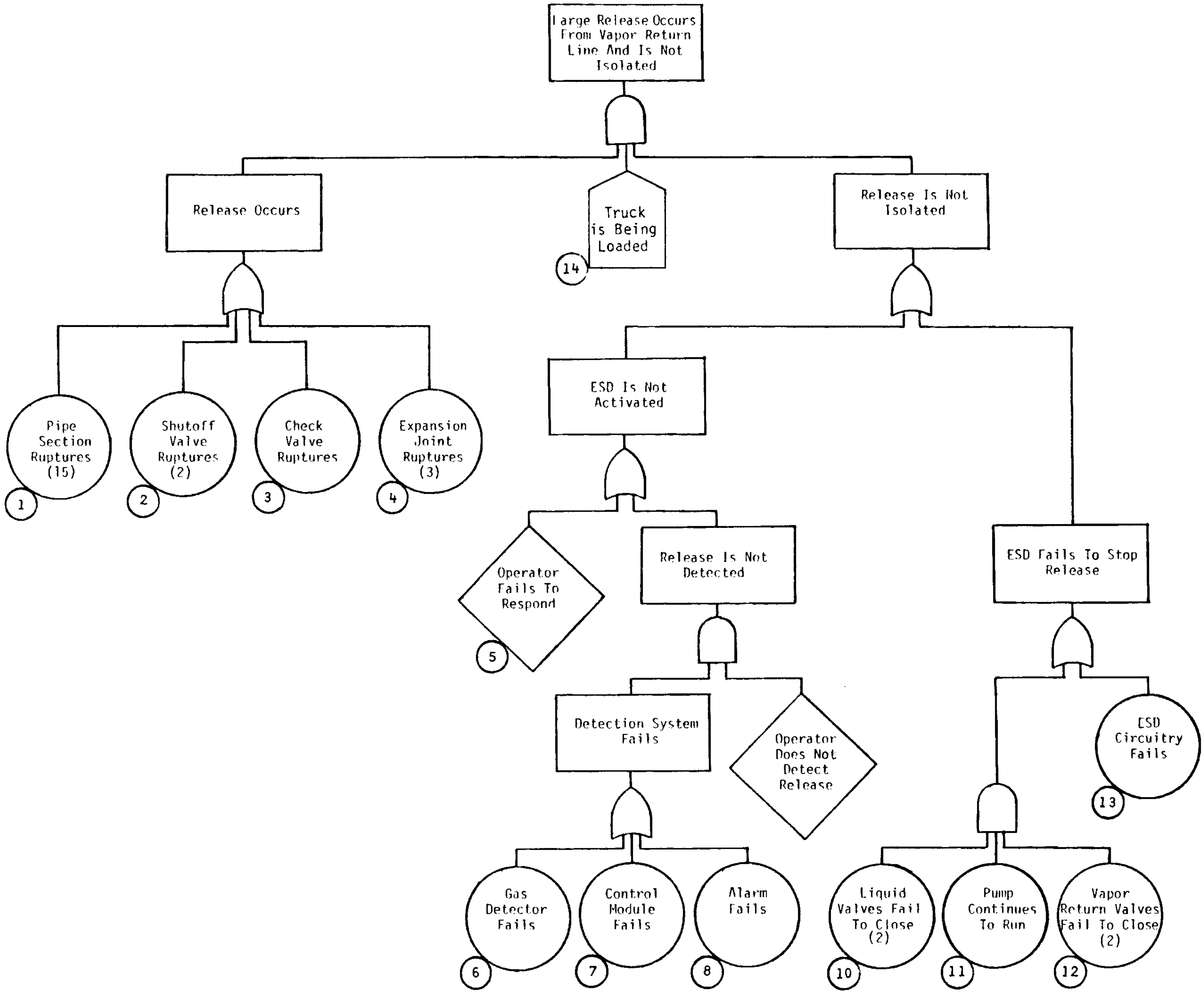


FIGURE B.16 Fault Tree for a Large Release Occurring from the Vapor Return Line. (Supporting Calculations in Table B.14)

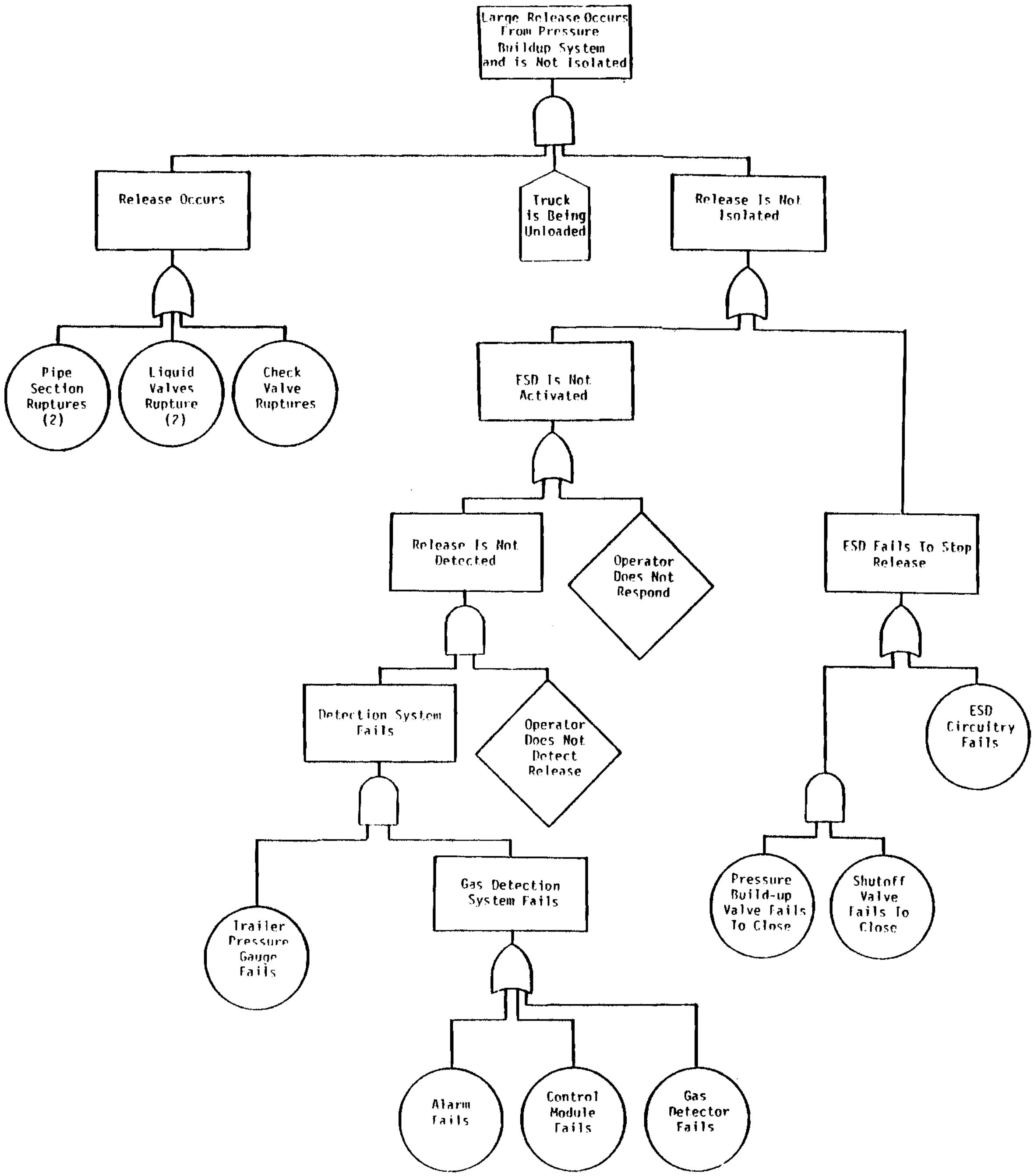


FIGURE B.17 Fault Tree for a Large Release from the Pressure Buildup System. (Supporting Calculations in Table B.15)

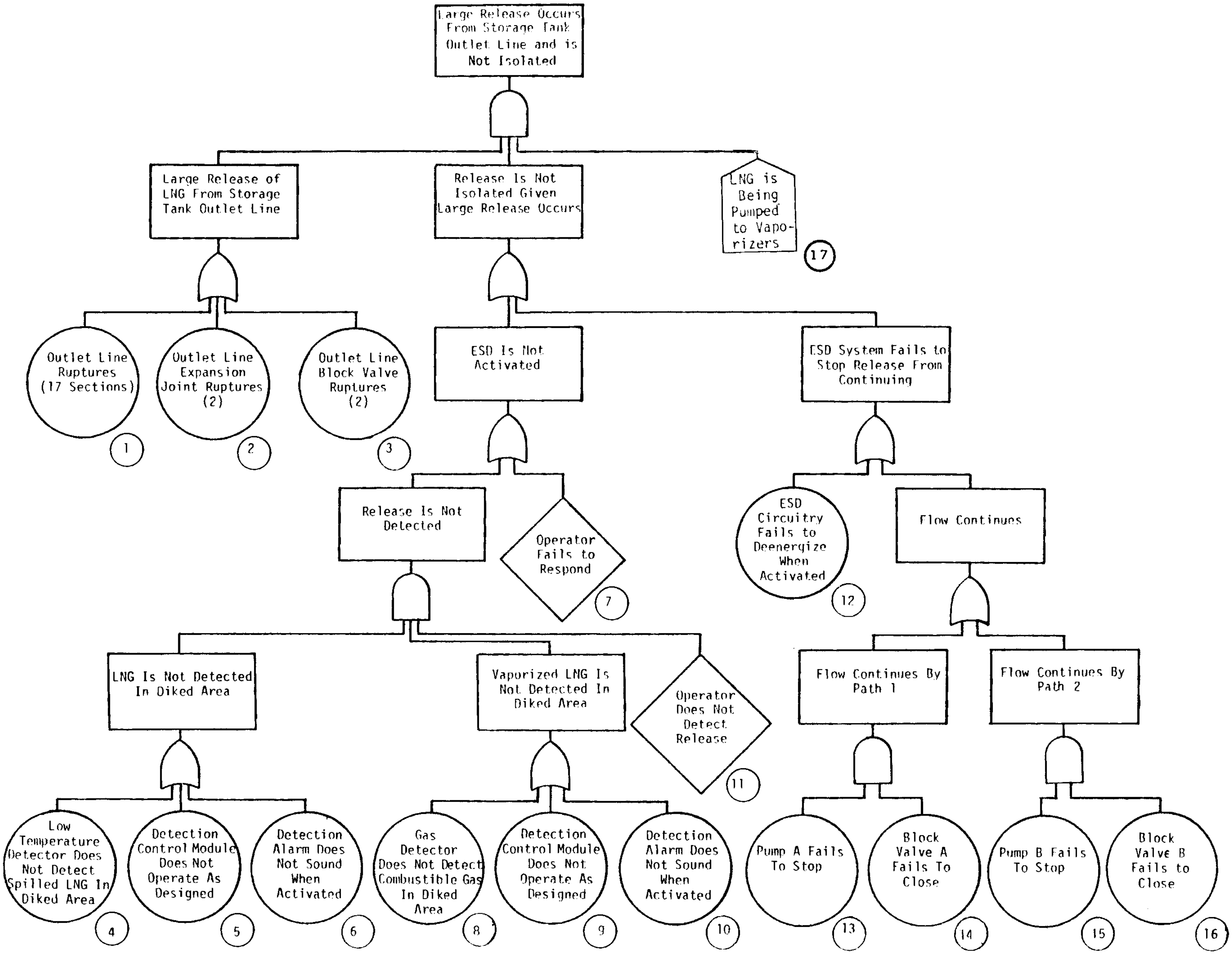


FIGURE B.18 Fault Tree for a Large Release from the Storage Tank Outlet Line Utilizing Intank Pumps. (Supporting Calculations in Table B.16)

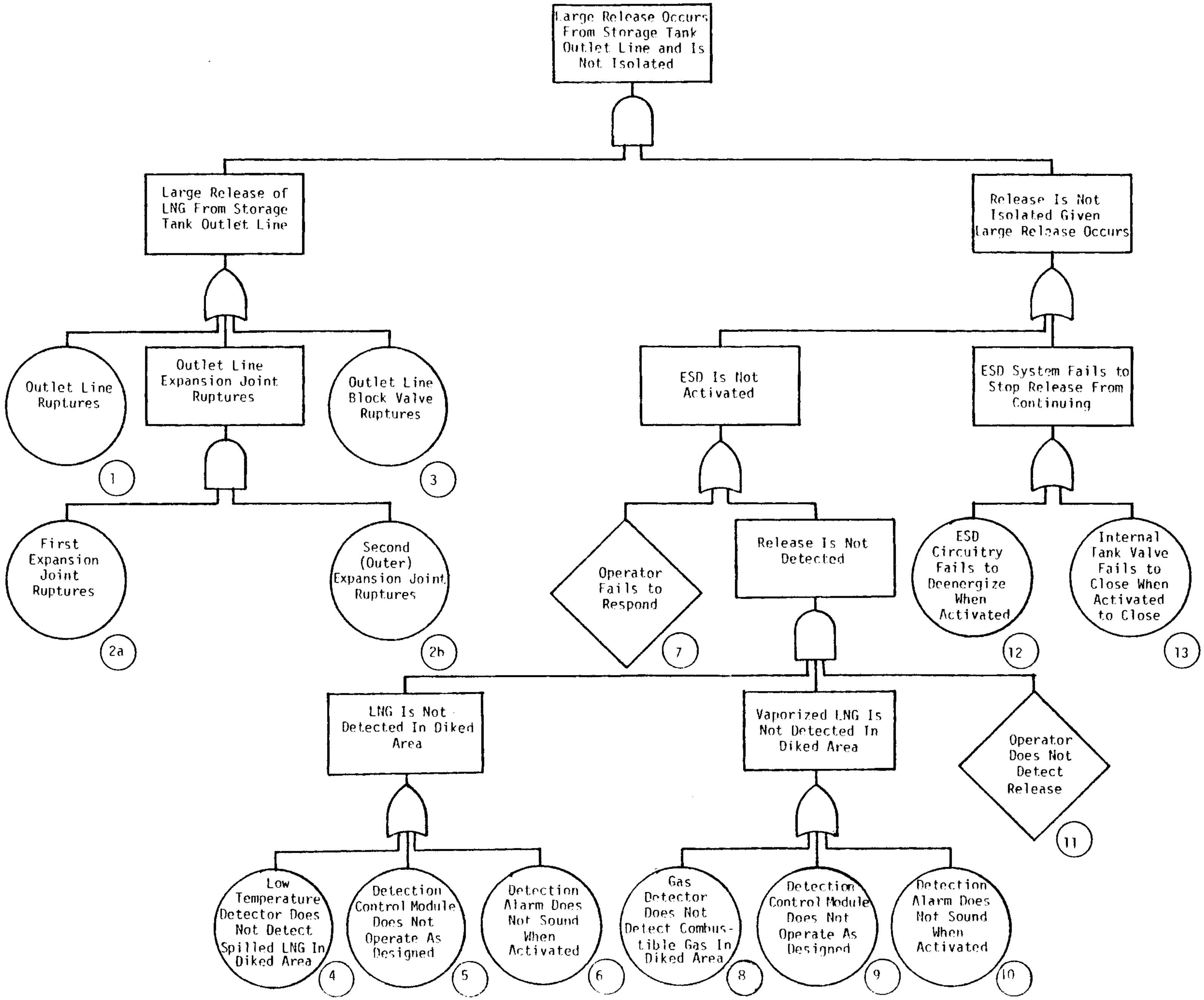


FIGURE B.19 Fault Tree for a Large Release from the Storage Tank Outlet Line Utilizing Double-Ply Expansion Joints. (Supporting Calculations in Table B.17)

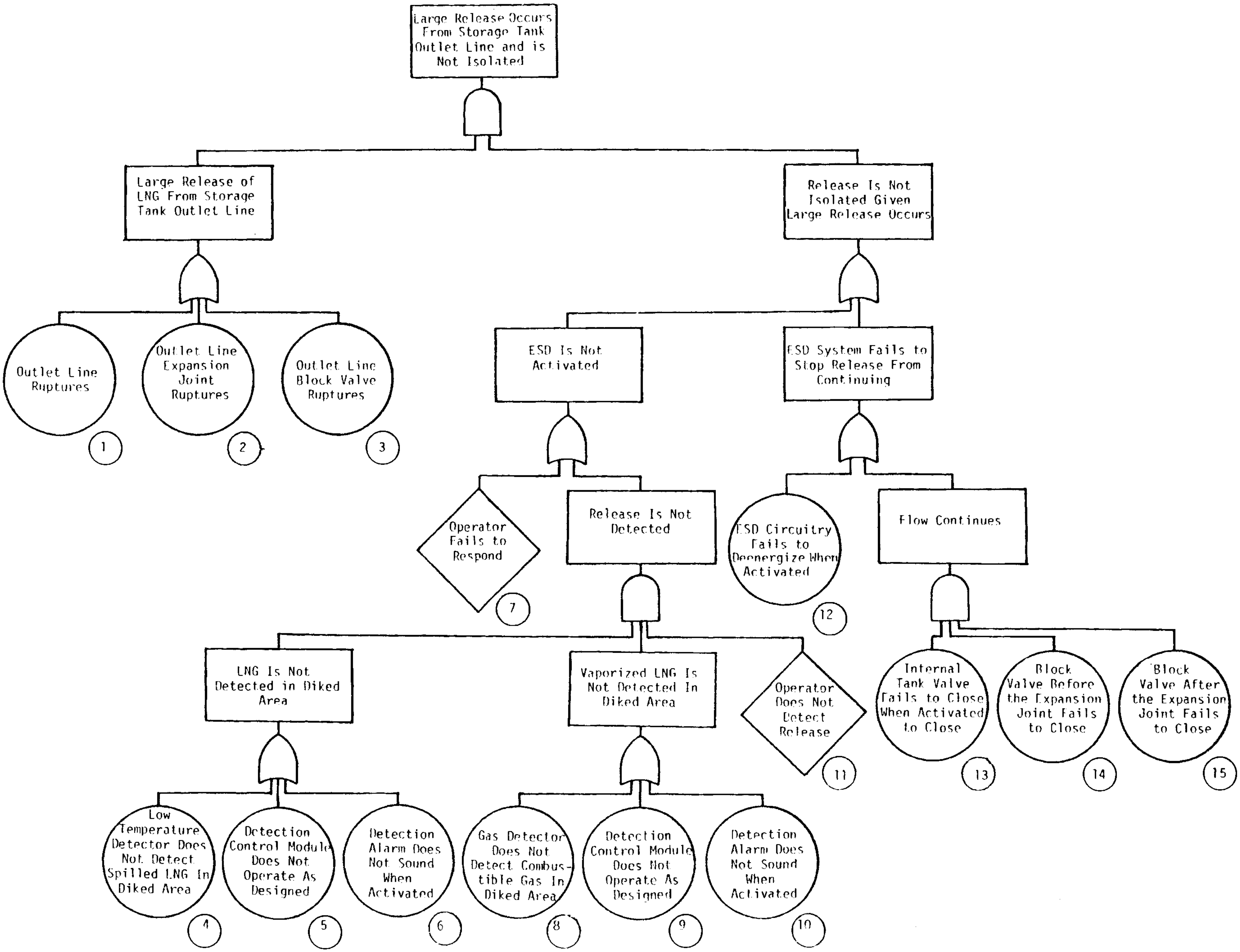


FIGURE B.20 Fault Tree for a Large Release from the Storage Tank Outlet Line Utilizing a Block Valve Upstream of the Expansion Joints. (Supporting Calculations in Table B.18)

TABLE B.1. Supporting Calculations for Failure of a Fully Automatic Emergency Shutdown System (See Figure B.1)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Sensor failure[b]	1×10^{-4}	1000	1×10^{-1}
2	Limit switch error[b]	3×10^{-5}	1000	3×10^{-2}
3	Alarm malfunction[c]	5×10^{-5}	80	4×10^{-3}
4	Operator does not observe release[a]			1×10^{-1}
5	Operator does not activate ESD[a]			1×10^{-2}
6	ESD circuit failure[b]	1×10^{-7}	1000	1×10^{-4}
7	Pumps fail to stop[b]	1×10^{-7}	1000	1×10^{-4}
8	Block valves fail to close[b]	5×10^{-5}	1000	5×10^{-2}

Minimum Cut Sets:[d]

$6 = 1 \times 10^{-4}$

$1 * 4 = 1 \times 10^{-2}$

$1 * 5 = 1 \times 10^{-3}$

$2 * 4 = 3 \times 10^{-3}$

$2 * 5 = 3 \times 10^{-4}$

$7 * 8 = \underline{5 \times 10^{-6}}$

1.4×10^{-2}/demand

(a) Faults per demand.
(b) Sensors, limit switches, ESD circuitry, and hardware are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and thus a mean dead time of 1000 hours.
(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and thus a mean dead time of 80 hours.
(d) Minimum cut sets are based on the assumption that the sensors and limit switches shown in Figure B.1 are <u>not</u> redundant but are the same in each branch of the fault tree that they occur.

TABLE B.2. Supporting Calculations for Failure of an Automatic or Manual Detection, Manual Activation Emergency Shutdown System (See Figure B.2)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Operator does not observe release[a]			1×10^{-1}
2	Sensor failure[b]	1×10^{-4}	1000	1×10^{-1}
3	Limit switch error[b]	3×10^{-5}	1000	3×10^{-2}
4	Alarm malfunction[c]	5×10^{-5}	80	4×10^{-3}
5	Operator does not activate ESD[a]			1×10^{-2}
6	ESD circuit failure[b]	1×10^{-7}	1000	1×10^{-4}
7	Pumps fail to stop[b]	1×10^{-7}	1000	1×10^{-4}
8	Block valves fail to close[b]	5×10^{-5}	1000	5×10^{-2}

9 Dectection System Error

$2 + 3 + 4 = 1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3} = 1.3 \times 10^{-1}$

10 Release is not Detected

$1 * 9 = 1 \times 10^{-1} * 1.3 \times 10^{-1} = 1.3 \times 10^{-2}$

11 ESD is not Activated

$10 + 15 = 1.3 \times 10^{-2} + 1 \times 10^{-2} = 2.3 \times 10^{-2}$

12 Isolation Equipment Failure

$7 * 8 = 1 \times 10^{-4} * 5 \times 10^{-2} = 5 \times 10^{-6}$

13 ESD Fails to Stop Release

$6 + 12 = 1 \times 10^{-4} + 5 \times 10^{-6} = 1 \times 10^{-4}$

14 Emergency Shutdown Failure

$11 + 13 = 2.3 \times 10^{-2} + 1 \times 10^{-4} = 2.3 \times 10^{-2}$/demand

(a) Faults per demand.
(b) Sensors, limit switches, ESD circuitry, and hardware are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and thus a mean dead time of 1000 hours.
(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and thus a mean dead time of 80 hours.

TABLE B.3. Supporting Calculations for LNG Storage Tank Overfill Fault Tree (See Figure B.4)

Assumptions:

1. Liquid level indicator A is a servo-powered, displacer-type device connected to a high level alarm that is set for a 92.5 ft height in the storage tank.
2. Liquid level indicator B is a differential pressure gauge connected to a high level alarm that is set for a 92.5 ft height in the storage tank.
3. The overflow line containing the low temperature sensor is attached to the storage tank at the 95 ft height in the storage tank. The low temperature sensor is connected to an alarm.
4. The storage tank is filled at a rate of 1710 bbl LNG/day for 200 days/year.
5. The operators calculate the total volume of LNG in the storage tank, based on the liquefaction rate, at the end of each liquefaction shift.
6. Operator failing to observe liquid level indicators high readings is assumed to be 3×10^{-1}/demand.

   (Note: At the import terminal during filling, the operators standard procedure is to observe the level indicators so the failure rate was reduced by a factor of 10 to 3×10^{-2}/demand.)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Operator fails to correctly calculate LNG level based on logbook information[(a)]			5×10^{-2}
2	Operator fails to observe level indicators[(a)]			3×10^{-1}
3	Level indicator A fails dangerously[(b)]	2×10^{-4}	100	2×10^{-2}

TABLE B.3. Contd

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
Basic Events:				
4	Level indicator B fails dangerously[b]	2×10^{-4}	100	2×10^{-2}
5	Operator fails to respond to high level alarms[a]			1×10^{-2}
6	High level switch A fails[c]	3×10^{-5}	1000	3×10^{-2}
7	High level alarm A fails to operate when activated[d]	5×10^{-5}	80	4×10^{-3}
8	High level switch B fails[c]	3×10^{-5}	1000	3×10^{-2}
9	High level alarm B fails to operate when activated[d]	5×10^{-5}	80	4×10^{-3}
10	Operator fails to respond to temperature alarm[a]			1×10^{-2}
11	Temperature sensor in liquid overflow line fails to sense LNG in line[c]	1×10^{-4}	1000	1×10^{-1}
12	Temperature sensor switch fails to operate[c]	3×10^{-5}	1000	3×10^{-2}
13	Temperature sensor alarm fails to operate when activated[d]	5×10^{-5}	80	4×10^{-3}
14	Storage tank is being filled 55% of the year.			

Minimum Cut Sets:

$$X = (10 + 11 + 12 + 13) * 14 = 7.9 \times 10^{-2}/\text{year}$$

$$2 * 5 * X = 3 \times 10^{-3} * X = 2.4 \times 10^{-4}/\text{year}$$

$$1 * 3 * 4 * X = 2 \times 10^{-5} * X = 1.6 \times 10^{-6}/\text{year}$$

$$2 * 3 * 4 * X = 1.2 \times 10^{-4} * X = 9.5 \times 10^{-6}/\text{year}$$

$$2 * 3 * 8 * X = 1.8 \times 10^{-4} * X = 1.4 \times 10^{-5}/\text{year}$$

$$2 * 3 * 9 * X = 2.4 \times 10^{-5} * X = 1.9 \times 10^{-6}/\text{year}$$

$$2 * 4 * 6 * X = 1.8 \times 10^{-4} * X = 1.4 \times 10^{-5}/\text{year}$$

TABLE B.3. Contd

Minimum Cut Sets:

$(2) * (4) * (7) * X = 2.4 \times 10^{-5} * X = 1.9 \times 10^{-6}$/year
$(2) * (6) * (8) * X = 2.7 \times 10^{-4} * X = 2.1 \times 10^{-5}$/year
$(2) * (6) * (9) * X = 3.6 \times 10^{-5} * X = 2.8 \times 10^{-6}$/year
$(2) * (7) * (8) * X = 3.6 \times 10^{-5} * X = 2.8 \times 10^{-6}$/year
$(2) * (7) * (9) * X = 4.8 \times 10^{-6} * X = \underline{3.8 \times 10^{-7}\text{/year}}$
3.1×10^{-4}/year

(a) Faults per demand.
(b) Liquid level indicator failures are assumed to have combined detection and repair time of 100 hours and thus a mean dead time of 100 hours also.
(c) Sensors, limit switches, ESD circuitry, and hardware are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and thus a mean dead time of 1000 hours.
(d) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and thus a mean dead time of 80 hours.

Release Calculations for Overfilling Tank

During liquefaction the tank is being filled at 6×10^6 scfd or 50 equivalent gallons of LNG per minute.

Assuming an overfill of 1 minute, then 50 gallons of LNG will spill into the annulus or vaporize.

Assuming an overfill lasting 10 minutes, then 50 gallons of LNG will spill into the annulus or vaporize.

TABLE B.4. Supporting Calculations for Storage Tank Reaches Vent Pressure (See Figure B.5)

Assumption:

1. The overall probability of peakshaving storage tank reaching vent pressure is 1/yr based on Reed Welker's information.

Release Calculations for LNG Vapor Venting from the Storage Tank

Assuming the maximum venting capacity of the two 12" relief valves is 283 MMscfd or ∿2400 equivalent gallons LNG per minute.

Assuming the venting for 1 minute

- At normal boiloff rates of 0.6×10^6 scfd or 417 scfm or 5 equivalent gallons per minute:

  417 scfm x 1 minute = 417 scf or 5 equivalent gallons LNG

- At maximum venting rates:

  $283 \times 10^6 \frac{\text{scf}}{\text{day}} \times \frac{\text{day}}{24 \text{ hrs}} \times \frac{\text{hr}}{60 \text{ min}} \times 1 \text{ min} = 197{,}000$ scf or 2350 equivalent gallons LNG.

Assuming venting for 10 minutes

- At normal boiloff rates:

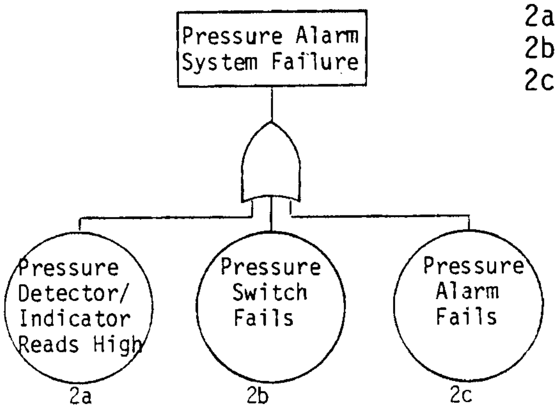
  417 scfm x 10 minute = 4170 scf or 50 equivalent gallons LNG

- At maximum venting rates:

  197,000 scfm x 10 minutes = 1.97×10^6 scf or 23,500 equivalent gallons LNG.

TABLE B.5. Supporting Calculations for LNG Storage Tank Reaches Vacuum Relief Pressure (See Figure B.6)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Abnormal increase in barometric pressure[(a)]	1×10^{-4}		
2	Gauge pressure alarm system fails			3.7×10^{-2}
	2a	1×10^{-4}	30[(b)]	3×10^{-3}
	2b	3×10^{-5}	1000[(c)]	3×10^{-2}
	2c	5×10^{-5}	80[(d)]	4×10^{-3}



		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
3	Operator does not respond to low pressure alarm			1×10^{-2}
4	Pressure control system fails dangerously	1.8×10^{-4}		
	4a	1×10^{-4}		
	4b	3×10^{-5}		
	4c	5×10^{-5}		

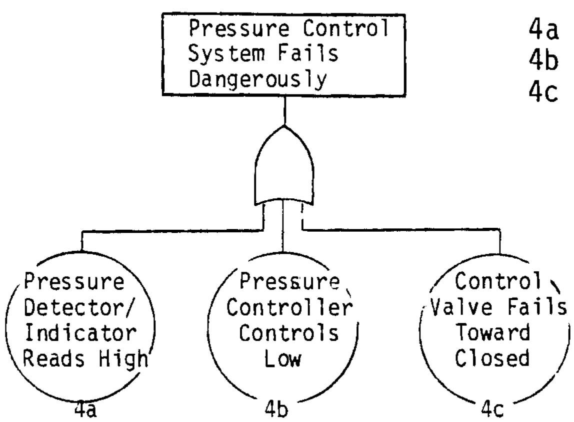


TABLE B.5. Contd

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Faults)
Basic Events:				
5	Atmospheric pressure alarm system fails (see details for 2)			3.7×10^{-2}
6	Operator does not respond to low pressure alarms			1×10^{-2}
7	Emergency gas pressurization system fails			8.3×10^{-2}
	7a	1×10^{-4}	30[b]	3×10^{-3}
	7b	3×10^{-5}	1000[c]	3×10^{-2}
	7c	5×10^{-5}	1000[c]	5×10^{-2}

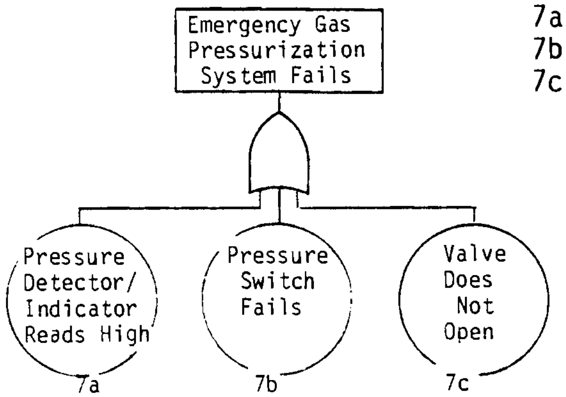


TABLE B.5. Contd

Cut Set Calculations:

1 · 2 · 7 = $(1 \times 10^{-4}/hr)\ (3.7 \times 10^{-2})\ (8.3 \times 10^{-2}) = 3.1 \times 10^{-7}/hr$

1 · 3 · 7 = $(1 \times 10^{-4}/hr)\ (1 \times 10^{-2})\ (8.3 \times 10^{-2}) = 8.3 \times 10^{-8}/hr$

4 · 6 · 7 = $(1.8 \times 10^{-4}/hr)\ (1 \times 10^{-2})\ (8.3 \times 10^{-2}) = 1.5 \times 10^{-7}/hr$

2 · 4 · 5 · 7 = $(3.7 \times 10^{-2})\ (1.8 \times 10^{-4}/hr)\ (3.7 \times 10^{-2})$

$(8.3 \times 10^{-2}) = 2.0 \times 10^{-8}/hr$

1 · 2 · 7 + 1 · 3 · 7 + 4 · 6 · 7 + 2 · 4 · 5 · 7
$= 5.6 \times 10^{-7}/hr$

$5.6 \times 10^{-7}/hr \times 8760/hr/year = 4.9 \times 10^{-3}/year$

(a) Assumed hazard rate.
(b) Pressure detector/indicators are assumed to have a mean dead time of 30 hours.
(c) Pressure switches and valves are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and thus a mean dead time of 1000 hours.
(d) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and thus a mean dead time of 80 hours.

TABLE B.6. Supporting Calculations for a Release from Storage Tank Inlet Line (See Figure B.7)

Assumptions:

1. The storage tank is assumed to be filled 200 days/year or 200/365 = .55 fraction of year.
2. The inlet line is taken from the plant descriptions as being 325' long. Therefore, there would be ∿17 - 20 ft pipe sections from cold box to storage tank.

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Main inlet line just before tank ruptures (17 pipe sections)[a]	$17(1 \times 10^{-10})$	t = 8760	1.5×10^{-5}
2	Main inlet line valve ruptures[a]	1×10^{-9}	t = 8760	8.8×10^{-6}
3	Main inlet line expansion joint ruptures[a]	1×10^{-7}	t = 8760	8.8×10^{-4}
4	Operator does not detect release[d] 1×10^{-1}			
5	Low temperature detector does not detect spilled LNG in diked area[b] 1×10^{-1}	1×10^{-4}	1000	1×10^{-1}
6	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
7	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
8	Gas detector does not detect combustible gas in diked area[b]	4×10^{-5}	1000	4×10^{-2}
9	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
10	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
11	Operator fails to respond[d]			1×10^{-2}

TABLE B.6. Contd

	λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:			
12 ESD system fails to stop release from continuing[b]	1×10^{-7}	1000	1×10^{-4}
13 Storage tank is being filled - exists .55/year			

Calculations:

14 ENF(expected number of failures) = 13 (1 + 2 + 3)
= (.55/year) $(1.5 \times 10^{-5} + 8.8 \times 10^{-6} + 8.8 \times 10^{-4})$
= 50×10^{-4}/year

15 Release is not isolated = [4 * (5 + 6 + 7) (8 + 9 + 10) + 11 + 12] = $[1 \times 10^{-1} * (1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3}) (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3}) + 1 \times 10^{-2} + 1 \times 10^{-4}]$
= 1.1×10^{-2}/demand

TOP EVENT = 14 * 15 = 5.5×10^{6}/year

(a) t = mission time of 1 year = 8760 hours
(b) Detectors, control modules, and ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and thus a mean dead time of 1000 hours.
(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and thus a mean dead time of 80 hours.
(d) Faults per demand.

Release Calculation for Tank Inlet Line Rupture

Based on a liquefaction rate of 6×10^{6} scfd or 50 gallons LNG/minute.

- Assuming a 1 minute release
  50 gallons LNG are released.
- Assuming a 10 minute release
  500 gallons LNG are released.

TABLE B.7. Supporting Calculations for Large Release from Storage Tank Outlet Line (See Figure B.8)

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Outlet line ruptures (5 sections)[a]	$5\ (1 \times 10^{-10})$	t = 480	2.4×10^{-7}
2	Outlet line expansion joint ruptures[a]	1×10^{-7}	t = 480	4.8×10^{-5}
3	Outlet line block valve ruptures[a]	1×10^{-9}	t = 480	4.8×10^{-7}
4	Low temperature detector does not detect spilled LNG in diked area[b]	1×10^{-4}	1000	1×10^{-1}
5	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
6	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
7	Operator fails to respond[d]			1×10^{-2}
8	Gas detector does not detect combustible gas in diked area[b]	4×10^{-5}	1000	4×10^{-2}
9	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
10	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
11	Operator does not detect release[d]			1×10^{-1}
12	ESD circuitry fails to deenergize when activated[b]	1×10^{-7}	1000	1×10^{-4}
13	Internal tank valve fails to close when activated to close[e]	5×10^{-5}	4000	2×10^{-1}

TABLE B.7. Contd

Calculations:

14 ENF(expected number of failures) = 1 + 2 + 3

$= (2.4 \times 10^{-7}/yr + 4.8 \times 10^{-5}/yr + 4.8 \times 10^{-7}/yr) = 4.9 \times 10^{-5}/yr$

15 Release is not isolated = [(4 + 5 + 6) (8 + 9 + 10) (11) + 7 + 12 + 13] $= [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3}) (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3}) (1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4} + 2 \times 10^{-1}] = 2.1 \times 10^{-1}$/demand

TOP EVENT = 14 * 15 $= 1.0 \times 10^{-5}$/year

(a) t = mission time of 20 days/year = 480 hours/year

(b) Detectors, control modules, and ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.

(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.

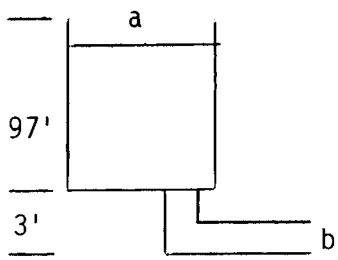
(d) Faults per demand.

(e) Internal tank valve is tested once a year for an assumed combined detection time and repair time of 8000 hours and a mean dead time of 4000 hours.

Rupture of an outlet LNG line will occur about 5×10^{-5} times per year. If the sendout system is shut down and the release isolated in 1 minute, 28,000 gallons would be spilled. If the system is not shut down for 10 minutes, 280,000 gallons would be released. This will occur 2×10^{-1} per demand, resulting in a probability of about 1×10^{-5} per year for this scenario. If an outlet line ruptures, gas detectors and LTDs 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. A fault tree for the rupture of a storage outlet line is shown in Figure C.8. If the internal valve did not close, the contents of the whole tank would be released up to 348,000 bbls.

LNG Outlet Line from the Storage Tank Fails:

The maximum height of LNG in the storage tank is 97'. The out line from the storage tank is assumed to exist 3' below the bottom of the inner tank. Velocity at a is assumed negligible. Assume a guillotine break.



$$\frac{gZa}{gc} + \cancel{\frac{U_a^2}{2gc}}^{0} = \cancel{\frac{gZb}{gc}}^{0} + \frac{U_b^2}{2gc}$$

$$\frac{gZa}{gc} = \frac{U_b^2}{2gc}$$

$$\frac{U_b^2}{2gc} = \frac{100g}{gc}$$

$$U_b^2 = 2 \times 100 \times 32.17 = 6434$$

$$U_b = 80.21 \text{ t/sec}$$

The pipe the LNG is released from is 12" pipe.

$$\text{Release rate} = 80.21 \frac{\text{ft}}{\text{sec}} \times \frac{(1 \text{ ft})^2 \pi}{4} \times \frac{7.48 \text{ gal}}{\text{ft}^3} = 471 \frac{\text{gal}}{\text{sec}}$$

$\therefore$ for 1 minute release, 471 gal/sec x 60 sec = 28,260 gal.

$\therefore$ for 10 minute release, 471 gal/sec x 600 sec = 282,600 gal.

Release Calculations Tank Outlet Line Rupture

Based on a hydrostatic pressure due to 100' of LNG (from top of inner tank to level of break) causing flow speed of 80 ft/sec or flow through 12" pipe of 470 gal/sec:

- Assuming a one minute release:
  28,000 gallons of LNG released.
- Assuming a 10 minute release:
  280,000 gallons of LNG released.

TABLE B.8. Supporting Calculations for Release from LNG Supply Line to Vaporizers at Pump Discharge (See Figure B.9)

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Supply line ruptures (27 pipe sections)[a]	$(1 \times 10^{-10})(27)$	t = 8760	2.4×10^{-5}
2	Pump outlet expansion joint ruptures (2 joints)[a]	$(1 \times 10^{-7})(2)$	t = 8760	1.8×10^{-3}
3	Pump outlet block valve ruptures (2 valves)[a]	$(1 \times 10^{-9})(2)$	t = 8760	1.8×10^{-5}
4	Line butterfly valves rupture (4 valves)[a]	$(1 \times 10^{-9})(4)$	t = 8760	3.5×10^{-5}
5	Line check valves rupture (4 valves)[a]	$(1 \times 10^{-9})(4)$	t = 8760	3.5×10^{-5}
6	Vaporizer inlet block valves rupture (2 valves)[a]	$(1 \times 10^{-9})(2)$	t = 8760	1.8×10^{-5}
7	Low temperature detector does not detect spilled LNG[b]	1×10^{-4}	1000	1×10^{-1}
8	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
9	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
10	Gas detector does not detect combustible gas[b]	4×10^{-5}	1000	4×10^{-2}
11	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
12	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
13	Operator does not detect release[d]			1×10^{-1}
14	Operator fails to respond[d]			1×10^{-2}

TABLE B.8. Contd

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:				
15	ESD circuitry fails[b]	1×10^{-7}	1000	1×10^{-4}
16	Internal tank valve fails to close when activated to close[e]	5×10^{-5}	4000	2×10^{-1}
17	Pump A inlet block valve fails to close[b]	5×10^{-5}	1000	5×10^{-2}
18	Sendout pump A fails to stop[f]	1×10^{-7}	4000	4×10^{-4}
19	Pump B inlet block valve fails to close[b]	5×10^{-5}	1000	5×10^{-2}
20	Sendout pump B fails to stop[f]	1×10^{-7}	4000	4×10^{-4}

Calculations:

21 ENF(expected number of failures) = 1 + 2 + 3 + 4 + 5 + 6 $= 2.4 \times 10^{-5} + 1.8 \times 10^{-3} + 1.8 \times 10^{-5} + 3.5 \times 10^{-5} + 3.5 \times 10^{-5} + 1.8 \times 10^{-5}) = 1.9 \times 10^{-3}$/year

22 Release is not isolated given Release Occurs = [(7 + 8 + 9) (10 + 11 + 12) (13) + 14 + 15 + 16 · 17 · 18 + 16 · 19 · 20] $= [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3})(4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})(1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4} + (2 \times 10^{-1} * 5 \times 10^{-2} * 4 \times 10^{-4}) + (2 \times 10^{-1} * 5 \times 10^{-2} * 4 \times 10^{-4})] = 1.1 \times 10^{-2}$/demand

TABLE B.8. Contd

Calculations:

TOP EVENT = 21 * 22 = 1.9×10^{-3}/year * 1.1×10^{-2}/demand
= 2.1×10^{-5}/year

(a) t = mission time of 8760 hours/year
(b) Low temperature detectors, gas detectors, detection control modules, ESD circuitry, and pump inlet block valves are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and thus a mean dead time of 80 hours.
(d) Faults per demand.
(e) Internal tank valve is tested once a year for an assumed combined detection time plus repair time of 8000 hours and a mean dead time is assumed to be 4000 hours.
(f) LNG pumps are assumed to have combined detection and repair time of 8000 hours and a man dead time of 4000 hours.

Release Calculations for Pump Outlet Line Rupture

Assuming the pumps do not overspeed with zero backpressure, then with one pump running at 625 gallons/minute

- Assuming a one minute release:
  625 gallons of LNG released.
- Assuming a 10 minute release:
  6250 gallons of LNG released.

TABLE B.9. Supporting Calculations for Release from Vaporizer Heat Exchanger Tube Failure (See Figure B.10)

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:				
1	Vaporizer heat exchanger tube ruptures (2 vaporizers)[a]	$(1 \times 10^{-4})(2)$	t = 480	9.6×10^{-2}
2	Gas detector does not detect combustible gas[b]	4×10^{-5}	1000	4×10^{-2}
3	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
4	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
5	UV detector does not detect any fires[b]	1×10^{-4}	1000	1×10^{-1}
6	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
7	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
8	Operator does not detect release[d]			1×10^{-1}
9	Operator fails to respond[d]			1×10^{-2}
10	ESD circuitry fails[b]	1×10^{-7}	1000	1×10^{-4}
11	Internal tank valve fails to close[e]	5×10^{-5}	4000	2×10^{-1}
12	Pump 1 inlet block valve fails to close[b]	5×10^{-5}	1000	5×10^{-2}
13	Pump 1 fails to stop[f]	1×10^{-7}	4000	4×10^{-4}
14	Pump 1 outlet block valve fails to close[b]	5×10^{-5}	1000	5×10^{-2}
15	Vaporizer A inlet block valve fails to close[b]	5×10^{-5}	1000	5×10^{-2}

TABLE B.9. Contd

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Faults)
Basic Events:				
16	Pump 2 inlet valve fails to close(b)	5×10^{-5}	1000	5×10^{-2}
17	Pump 2 fails to stop(f)	1×10^{-7}	4000	4×10^{-4}
18	Pump 2 outlet block valve fails to close(b)	5×10^{-5}	1000	5×10^{-2}
19	Vaporizer B inlet block valve fails to close(b)	5×10^{-8}	1000	5×10^{-2}

Calculations:

20 ENF(expected number of failures) = (1) = 9.6×10^{-2}/year

Minimal Cut Set Calculation

$$A = (11 * 12 * 13 * 14 * 15 * 19) + 2(11 * 12 * 13 * 14 * 15 * 16 * 17 * 18 * 19) + (11 * 15 * 16 * 17 * 18 * 19) = 1.0 \times 10^{-9}$$

21 Release is not isolated = $(2 + 3 + 4)(5 + 6 + 7)(8) + 9 + 10 + A = 1.1 \times 10^{-2}$/demand

TOP EVENT = $20 * 21 = 9.6 \times 10^{-2}$/year $* 1.1 \times 10^{-2}$/demand
$= 1.1 \times 10^{-3}$/year

(a) t = mission time of 480 hours/year
(b) Gas detector, UV detector, detection control modules, ESD circuitry, pump inlet and outlet block valves and vaporizer inlet block valves are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(d) Faults per demand.
(e) Internal tank valve is tested once a year for an assumed combined detection plus repair time of 8000 hours and a mean dead time is assumed to be 4000 hours.
(f) LNG pumps are assumed to have combined detection and repair time of 8000 hours and a man dead time of 4000 hours.

Release Calculations for Vaporizer Heat Exchanger Tube Rupture

Assuming one tube is completely ruptured:

- Assuming a one minute release:
  625 gallons/minute x 2 pumps x 1 minute = 1,250 gallons.
- Assuming a 10 minute release:
  625 gallons/minute x 2 pumps x 10 minute = 12,500 gallons.

TABLE B.10. Supporting Calculations for Release from Submerged Compustion Vaporizer Outlet Line Rupture (See Figure B.11)

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Gas temperature detector/ indicator fails dangerously	1×10^{-4}		
2	Controller fails dangerously	3×10^{-5}		
3	Control valve fails to close when LNG temperature is sensed	5×10^{-5}		
4	Control module fails[a]	3×10^{-5}	1000	3×10^{-2}
5	Alarm fails	5×10^{-5}	80	4×10^{-3}
6	Low pressure detector on water jacket pump fails[a]	1×10^{-4}	1000	1×10^{-1}
7	Control module fails[a]	3×10^{-5}	1000	3×10^{-2}
8	Alarm fails[b]	5×10^{-5}	80	4×10^{-3}
9	UV detector fails to detect burner flame out[a]	1×10^{-4}	1000	1×10^{-1}
10	Control module fails[a]	3×10^{-5}	1000	3×10^{-2}
11	Alarm fails[b]	5×10^{-5}	80	4×10^{-3}
12	Operator does not observe release[c]			1×10^{-1}
13	Operator does not respond to alarms[c]			1×10^{-2}
14	ESD circuitry fails[a]	1×10^{-7}	1000	1×10^{-4}
15	Internal tank valve fails to close[d]	5×10^{-5}	4000	2×10^{-1}

TABLE B.10. Contd

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Probability)
Basic Events:				
16	Pump 1 inlet block valve fails to close[a]	5×10^{-5}	1000	5×10^{-2}
17	Pump 1 fails to stop[e]	1×10^{-7}	4000	4×10^{-4}
18	Pump 1 outlet block valve fails to close[a]	5×10^{-5}	1000	5×10^{-2}
19	Vaporizer A inlet block valve fails to close[a]	5×10^{-5}	1000	5×10^{-2}
20	Pump 2 inlet valve fails to close[a]	5×10^{-5}	1000	5×10^{-2}
21	Pump 2 fails to stop[e]	1×10^{-7}	4000	4×10^{-4}
22	Pump 2 outlet block valve fails to close[a]	5×10^{-5}	1000	5×10^{-2}
23	Vaporizer B inlet block valve fails to close[a]	5×10^{-5}	1000	5×10^{-2}
24	Gas temperature detector fails[a]	1×10^{-4}	1000	1×10^{-1}
25	Plant is vaporizing LNG - 20 days/365 days = 5.5×10^{-2}			

Calculations:

26 ENF(expected number of failures) = 25 (1 + 2 + 3) = (5.5×10^{-2}) $(1 \times 10^{-4}/\text{hr} + 3 \times 10^{-5}/\text{hr} + 5 \times 10^{-5}/\text{hr})$ (8760 hr/yr) = 8.7×10^{-2}/yr

Minimal Cut Set Calculation

A = (15 * 16 * 17 * 18 * 19 * 23) + 2(15 * 16 * 17 * 18 * 19 * 20 * 21 * 22 * 23) + (15 * 19 * 20 * 21 * 22 * 23) = 1.0×10^{-9}

TABLE B.10. Contd

Calculations:

27 Release is not isolated = (24 + 4 + 5) (6 + 7 + 8) (9 + 10 + 11) (12) + 13 + 14 + A = 1.0×10^{-2}/demand

TOP EVENT = 26 * 27 = 8.7×10^{-2}/year * 1.0×10^{-2}/demand
= 8.7×10^{-4}/year

(a) Gas temperature detector, low pressure detector, UV detector, control modules, ESD circuitry, pump inlet and outlet valves, and vaporizer inlet valves are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(b) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Faults per demand.
(d) Internal tank valve is tested once a year for an assumed combined detection plus repair time of 8000 hours and a mean dead time is assumed to be 4000 hours.
(e) LNG pumps are assumed to have combined detection and repair time of 8000 hours and a man dead time of 4000 hours.

Release Calculations for Vaporizer Outlet Line Rupture

Based on two pumps running and assuming all flow will exit through the rupture:

- Assuming a one minute release:
  625 gallons/minute x 2 pumps x 1 minute = 1,250 gallons.
- Assuming a 10 minute release:
  625 gallons/minute x 2 pumps x 10 minute = 12,500 gallons.

TABLE B.11. Supporting Calculations--Liquid Loading Line (See Figure B.13)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Pipe section ruptures (15 sections)	$(15)(1 \times 10^{-10})$		
2	Block valve ruptures (2 valves)	$(2)(1 \times 10^{-9})$		
3	Expansion joint ruptures (3 joints)	$(3)(1 \times 10^{-7})$		
4	Transfer pump ruptures	(1×10^{-8})		
5	Drain valve ruptures (2 valves)	$(2)(1 \times 10^{-9})$		
6	Operator fails to respond[a]			1×10^{-2}
7	Alarm fails[b]	5×10^{-5}	80	4×10^{-3}
8	Control module fails[c]	3×10^{-5}	1000	3×10^{-2}
9	Gas detector fails[c]	4×10^{-5}	1000	4×10^{-2}
10	Operator does not detect release[a]			1×10^{-1}
11	Pump does not stop[c]	1×10^{-7}	1000	1×10^{-4}
12	Block valve fails to open[c]	5×10^{-5}	1000	5×10^{-2}
13	ESD circuitry fails[c]	1×10^{-7}	1000	1×10^{-4}
14	Truck is being loaded[d]	340 hr/8760 hr = 3.9×10^{-2}		

TABLE B.11. Contd

Calculations:

15 ENF(expected number of failures) = 14 (1 + 2 + 3 + 4 + 5)
$= (3.9 \times 10^{-2})\ (1.5 \times 10^{-9}/hr + 2 \times 10^{-9}/hr + 3 \times 10^{-7}/hr$
$+ 1 \times 10^{-8}/hr + 2 \times 10^{-9}/hr)\ (8760\ hr/yr)^{(e)} = 1.1 \times 10^{-4}/yr$

16 Release is not isolated = 6 + (7 + 8 + 9) (10) + (11 * 12)
+ 13 $= 1 \times 10^{-2} + (4 \times 10^{-3} + 3 \times 10^{-2} + 4 \times 10^{-2})\ (1 \times 10^{-1})$
$+ (1 \times 10^{-4} * 5 \times 10^{-2}) + 1 \times 10^{-4} = 1.8 \times 10^{-2}/demand$

TOP EVENT = 15 * 16 $= (1.1 \times 10^{-4}/year)\ (1.8 \times 10^{-2})$
$= 2.0 \times 10^{-6}/year$

(a) Faults per demand.
(b) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Gas detector, control module, pump, block valve, and ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(d) The loading time of 340 hours per year is based on assumed 400 trucks loaded out per year. Of these 400 loads, 90% were assumed to be 'cold' when loading started so only took one-half hour to load and 10% were assumed to be 'warm' when loading started so took 4 hours to load.

400 x .90 x 1/2 hour + 400 x .10 x 4 hours = 340 hours

(e) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

TABLE B.12. Supporting Calculations--Liquid Loading Line (See Figure B.14)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Pipe section ruptures (15 sections)	$(15)(1 \times 10^{-10})$		
2	Block valve ruptures (3 valves)	$(3)(1 \times 10^{-9})$		
3	Drain valve ruptures (2 valves)	$(2)(1 \times 10^{-9})$		
4	Expansion joint ruptures (3 joints)	$(3)(1 \times 10^{-7})$		
5	Operator fails to respond[a]			1×10^{-2}
6	Alarm fails[b]	5×10^{-5}	80	4×10^{-3}
7	Control module fails[c]	3×10^{-5}	1000	3×10^{-2}
8	Gas detector fails[c]	4×10^{-5}	1000	4×10^{-2}
9	Operator does not detect release[a]			1×10^{-1}
10	Truck drain valve fails to close[b]	5×10^{-5}	80	4×10^{-3}
11	Block valve fails to close[b]	5×10^{-5}	80	4×10^{-3}
12	ESD circuitry fails[c]	1×10^{-7}	1000	1×10^{-4}
13	Truck is being unloaded[d]	200 hr/8760 hr = 2.3×10^{-2}		

TABLE B.12. Contd

Calculations:

14 ENF(expected number of failures) = (13) ((1) + (2) + (3) + (4))
$= 2.3 \times 10^{-2}(1.5 \times 10^{-9}/hr + 3 \times 10^{-9}/hr + 2 \times 10^{-9}/hr$
$+ 3 \times 10^{-7}/hr)\ (8760\ hr/yr)^{(e)} = 6.2 \times 10^{-5}/yr$

15 Release is not isolated = (5) + ((6) + (7) + (8)) ((9)) + ((10) * (11))
$+ (12) = 1 \times 10^{-2} + (4 \times 10^{-3} + 3 \times 10^{-2} + 4 \times 10^{-2})\ (1 \times 10^{-1})$
$+ (4 \times 10^{-3} * 4 \times 10^{-3}) + 1 \times 10^{-4} = 1.8 \times 10^{-2}$/demand

TOP EVENT = (14) * (15) $= (6.2 \times 10^{-5}/year)\ (1.8 \times 10^{-2})$
$= 1.1 \times 10^{-6}$/year

(a) Faults per demand.
(b) Alarms, truck drain valve, and block valve are assumed to be on a weekly test interval for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Gas detector, control module, and ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(d) The loading time of 200 hours per year is based on an assumed 400 trucks unloaded per year for an average of one-half hour per truck.
(e) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

TABLE B.13. Supporting Calculations--Sendout Pump (See Figure B.15)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Suction line ruptures	1×10^{-10}		
2	Sendout pump ruptures	1×10^{-8}		
3	Suction valve ruptures	1×10^{-9}		
4	Discharge valve ruptures	1×10^{-9}		
5	Operator fails to respond[a]			1×10^{-2}
6	Alarm fails[b]	5×10^{-5}	80	4×10^{-3}
7	Control module fails[c]	3×10^{-5}	1000	3×10^{-2}
8	Gas detector fails[c]	4×10^{-5}	1000	4×10^{-2}
9	Pump performance indicator fails (flow detector)[c]	2×10^{-4}	1000	2×10^{-1}
10	Operator does not detect release[a]			1×10^{-1}
11	ESD circuitry[c]	1×10^{-7}	1000	1×10^{-4}
13	Block valve fails to close[c]	5×10^{-5}	1000	5×10^{-4}

TABLE B.13. Contd

Calculations:

13 ENF(expected number of failures) = (1 + 2 + 3 + 4)
= $(1 \times 10^{-10}/\text{hr} + 1 \times 10^{-8}/\text{hr} + 1 \times 10^{-9}/\text{hr} + 1 \times 10^{-9}/\text{hr})$
$(8760\ \text{hr/yr})^{(d)} = 1.1 \times 10^{-4}/\text{yr}$

14 Release is not isolated = 5 + (6 + 7 + 8 + 9) (10) + 11
+ 12 $= 1 \times 10^{-2} + (4 \times 10^{-3} + 3 \times 10^{-2} + 4 \times 10^{-2} + 2 \times 10^{-1})$
$(1 \times 10^{-1}) + 1 \times 10^{-4} + 5 \times 10^{-2} = 8.8 \times 10^{-2}/\text{demand}$

TOP EVENT = 13 * 14 = $(1.1 \times 10^{-4}/\text{year})\ (8.8 \times 10^{-2}/\text{demand})$
$= 9.7 \times 10^{-6}/\text{year}$

(a) Faults per demand.
(b) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Gas detector, control module, flow detector, ESD circuitry, and block valve are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(d) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

TABLE B.14. Supporting Calculations--Vapor Return Line (See Figure B.16)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Pipe section ruptures (15 sections)	$(15)(1 \times 10^{-10})$		
2	Shutoff valve ruptures (2 valves)	$(2)(1 \times 10^{-9})$		
3	Check valve ruptures	(1×10^{-9})		
4	Expansion joint ruptures (3 joints)	$(3)(1 \times 10^{-7})$		
5	Operator fails to respond[(a)]			1×10^{-2}
6	Gas detector fails[(c)]	4×10^{-5}	1000	4×10^{-2}
7	Control module fails[(c)]	3×10^{-5}	1000	3×10^{-2}
8	Alarm fails[(b)]	5×10^{-5}	80	4×10^{-3}
9	Operator does not detect release[(a)]			1×10^{-1}
10	Liquid valves fail to close (2 valves)[(b)]	$(2)(5 \times 10^{-5})$	80	8×10^{-3}
11	Pump continues to run[(c)]	(1×10^{-7})	1000	1×10^{-4}
12	Vapor return valves fail to close (2 valves)[(b)]	$(2)(5 \times 10^{-5})$	80	8×10^{-3}
13	ESD circuitry fails[(c)]	1×10^{-7}	1000	1×10^{-4}
14	Truck is being loaded[(d)]	340 hr/8760 hr = 3.9×10^{-2}		

TABLE B.14. Contd

Calculations:

15 ENF(expected number of failures) = 14 (1 + 2 + 3 + 4)
$= 3.9 \times 10^{-2}(1.5 \times 10^{-9}/hr + 2 \times 10^{-9}/hr + 1 \times 10^{-9}/hr + 3 \times 10^{-7}/hr)\ (8760\ hr/yr)^{(e)} = 1.0 \times 10^{-4}/yr$

16 Release is not isolated = 5 + (6 + 7 + 8) (9) + (10 * 11 * 12) + 13 $= 1 \times 10^{-2} + (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})(1 \times 10^{-1}) + (8 \times 10^{-3} * 1 \times 10^{-4} * 8 \times 10^{-3}) + 1 \times 10^{-4}$
$= 1.8 \times 10^{-2}$/demand

TOP EVENT = 15 * 16 $= (1.0 \times 10^{-4}/year)\ (1.8 \times 10^{-2}/demand)$
$= 1.8 \times 10^{-6}/year$

(a) Faults per demand.
(b) Alarms, liquid valves, and vapor return valves are all assumed to be on a weekly test interval for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Gas detector, control module, pump, and ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(d) Since the vapor return lines are only used when trucks are being loaded, the loading time of 340 hours per year is based on an assumed 400 trucks loaded out per year. Of these 400 loads, 90% were assumed to be 'cold' when loading started so only took one-half hour to load and 10% were assumed to be 'warm' when loading started so took 4 hours to load.

400 * .90 * 1/2 hour + 400 * .10 x 4 hours = 340 hours

(e) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

TABLE B.15. Supporting Calculations--Pressure Build-up System (See Figure B.17)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Pipe section ruptures (2 sections)	$(2)(1 \times 10^{-10})$		
2	Liquid valves rupture (2 valves)	$(2)(1 \times 10^{-9})$		
3	Check valve ruptures	1×10^{-9}		
4	Operator does not respond[a]			1×10^{-2}
5	Trailer pressure gauge fails[b]	1×10^{-4}	80	8×10^{-3}
6	Alarm fails[b]	5×10^{-5}	80	4×10^{-3}
7	Control module fails[c]	3×10^{-5}	1000	3×10^{-2}
8.	Gas detector fails[c]	4×10^{-5}	1000	4×10^{-2}
9	Operator does not detect release[a]			1×10^{-1}
10	Pressure build-up valve fails to close[b]	5×10^{-5}	80	4×10^{-3}
11	Shutoff valve fails to close[b]	5×10^{-5}	80	4×10^{-3}
12	ESD circuitry fails[c]	1×10^{-7}	1000	1×10^{-4}
13	Truck is being unloaded[d]	200 hr/8760 hr = 2.3×10^{-2}		

TABLE B.15. Contd

Calculations:

14 ENF(expected number of failures) = 13 (1 + 2 + 3) = 2.3 $\times 10^{-2}(2 \times 10^{-10}/hr + 2 \times 10^{-9}/hr + 1 \times 10^{-9}/hr)$
$= 7.4 \times 10^{-11}/hr$ (8760 hr/yr)[e] $= 6.4 \times 10^{-7}/yr$

15 Release is not isolated = 4 + [(5) (6 + 7 + 8)] [9] + (10 * 11) + 12 $= 1 \times 10^{-2} + [(8 \times 10^{-3})(4 \times 10^{-3} + 3 \times 10^{-2} + 4 \times 10^{-2})][1 \times 10^{-1}] + (4 \times 10^{-3} * 4 \times 10^{-3}) + 1 \times 10^{-4} = 1.0 \times 10^{-2}$/demand

TOP EVENT = 14 * 15 $= (6.4 \times 10^{-7}/year)(1.0 \times 10^{-2}/demand)$
$= 6.4 \times 10^{-9}/year$

(a) Faults per demand.
(b) Alarms, trailer pressure gauge, shutoff valve and pressure buildup valve are all assumed to be on a weekly test interval for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Gas detector, control module, and the ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(d) Since the pressure buildup system is only used when unloading the truck, the unloading time of 200 hours per year is based on an assumed 400 trucks unloaded out per year at one-half hour per truck to unload.
(e) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

TABLE B.16. Supporting Calculations for Release from Storage Tank Outlet Line with Alternative Design of In-Tank Pumps (See Figure B.18)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Outlet line ruptures (17 sections)	$(17)(1 \times 10^{-10})$		
2	Outlet line expansion joint ruptures (2 joints)	$(2)(1 \times 10^{-7})$		
3	Outlet line block valve ruptures (2 valves)	$(2)(1 \times 10^{-9})$		
4	Low temperature detector does not detect spilled LNG in diked area[a]	1×10^{-4}	1000	1×10^{-1}
5	Detection control module does not operate as designed[a]	3×10^{-5}	1000	3×10^{-2}
6	Detection alarm does not sound when activated[b]	5×10^{-5}	80	4×10^{-3}
7	Operator fails to respond[c]			1×10^{-2}
8	Gas detector does not detect combustible gas in diked area[a]	4×10^{-5}	1000	4×10^{-2}
9	Detection control module does not operate as designed[a]	3×10^{-5}	1000	3×10^{-2}
10	Detection alarm does not sound when activated[b]	5×10^{-5}	80	4×10^{-3}
11	Operator does not detect release[c]			1×10^{-1}
12	ESD circuitry fails to deenergize when activated[a]	1×10^{-7}	1000	1×10^{-4}
13	Pump A fails to stop[e]	1×10^{-7}	4000	4×10^{-4}
14	Block valve A fails to close[a]	5×10^{-5}	1000	5×10^{-2}
15	Pump B fails to stop[e]	1×10^{-7}	4000	4×10^{-4}
16	Block valve B fails to close[a]	5×10^{-5}	1000	5×10^{-2}

TABLE B.16. Contd

	λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)

Basic Events:

17 LNG is being pumped to the vaporizers[d] 480 hr/8760 hr = 5.5×10^{-2}

Calculations:

18 Minimum Cut Set for Outlet Line Rupture

This assumes that an outlet line rupture is downstream of the pumps and block valves therefore a release will continue if both pumps and block valves fail to stop or close respectively.

$$(1) [((4) + (5) + (6)) ((8) + (9) + (10)) ((11)) + (7) + (12) + ((13) * (14)) + ((15) * (16))] (17) = (1.7 \times 10^{-9}/\text{hr}) (8760 \text{ hrs/yr})^{(f)} [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3}) (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3}) (1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4} + (4 \times 10^{-4} * 5 \times 10^{-2}) + (4 \times 10^{-4} * 5 \times 10^{-2})] (5.5 \times 10^{-2}) = 9.1 \times 10^{-9}/\text{yr}$$

19 Minimum Cut Set for Outlet Line Expansion Joint Rupture

(2A) = Hazard rate expansion Joint A = 1×10^{-7}/hr
(2B) = Hazard rate expansion Joint B = 1×10^{-7}/hr

This assumes that the expansion joints are upstream of the block valves therefore is expansion joint A ruptures block valve A won't be involved in stopping flow by path 1 but it will be possible to stop flow by path 2. Vice-versa for expansion joint B.

$$(17) \left\{ (2) [((4) + (5) + (6)) ((8) + (9) + (10)) ((11)) + (7) + (12)] + (2A) [(13) + (14) (15) (16)] + (2B) [(15) + (13) (14) (16)] \right\} = (5.5 \times 10^{-2}) \left\{ (2 \times 10^{-7}/\text{hr}) (8760 \text{ hr/yr})^{(f)} [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3}) * (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3}) (1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4}] + (1 \times 10^{-7}/\text{hr}) (8760 \text{ hr/yr})^{(f)} [4 \times 10^{-4} + (5 \times 10^{-2} * 4 \times 10^{-4} * 5 \times 10^{-2})] + (1 \times 10^{-7}/\text{hr}) (8760 \text{ hr/yr})^{(f)} [4 \times 10^{-4} + (4 \times 10^{-4} * 5 \times 10^{-2} * 5 \times 10^{-2})] \right\} = 1.1 \times 10^{-6}/\text{yr}$$

TABLE B.16. Contd

Calculations:

20 Minimum Cut Set for Outlet Line Block Valve Rupture

3A = Hazard rate block valve A rupture = 1×10^{-9}/hr

3B = Hazard rate block valve B rupture = 1×10^{-9}/hr

This calculation is based on the fact that if block valve A ruptures then block valve A won't be involved in ESD shutdown by either path 1 or 2. Vice-versa for block valve B.

$$17 \{ 3 [(4 + 5 + 6)(8 + 9 + 10)(11) + 7 + 12] + 3A [13 + 15 * 16] + 3B [15 + 13 * 14] \}$$

$$= (5.5 \times 10^{-2}) \{(2 \times 10^{-9}/\text{hr})(8760\ \text{hr/yr})^{(f)} [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3}) * (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})(1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4}] + (1 \times 10^{-9}/\text{hr})(8760\ \text{hr/yr})^{(f)} [4 \times 10^{-4} + (4 \times 10^{-4})(5 \times 10^{-2})] + (1 \times 10^{-9}/\text{hr})(8760\ \text{hr/yr})^{(f)} [4 \times 10^{-4} + (4 \times 10^{-4})(5 \times 10^{-2})]\} = 1.1 \times 10^{-8}/\text{yr}$$

$$\text{TOP EVENT} = 18 + 19 + 20 = 9.1 \times 10^{-9}/\text{yr} + 1.1 \times 10^{-6}/\text{yr} + 1.1 \times 10^{-8}/\text{yr} = 1.1 \times 10^{-6}/\text{yr}$$

(a) Detectors, control modules, ESD Circuitry, and block valves are tested on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(b) Alarms are on a weekly test interval for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Faults per demand.
(d) LNG is being pumped to vaporizers 20 days/year or 480 hours/year.
(e) Pumps are assumed to be tested once a year for an assumed combined detection plus repair time of 8000 hours and a mean dead time of 4000 hours.
(f) Multiplication by 8760 hours per year is to convert the λ = faults/hour to λ = faults/year.

Release Calculations for Tank Outlet Line Rupture with Alternative Design of In-Tank Pumps

Based on an assumed pumping rate for each pump of 625 gallons/minute:

- Assuming a one minute release:

$$625 \text{ gallons/minute} \times 2 \text{ pumps} \times 1 \text{ minute} + \frac{\pi(8")^2}{4 \times 144\frac{in^2}{ft^2}}$$

$$\times\ 17 \text{ pipe sections} \times \frac{20 \text{ ft}}{\text{section}} \times \frac{7.48 \text{ gal}}{ft^3} = 2{,}138 \text{ gallons}$$

- Assuming a ten-minute release:

$$625 \text{ gallons/minute} \times 2 \text{ pumps} \times 10 \text{ minutes} + \frac{\pi(8")^2}{4 \times 144\frac{in^2}{ft^2}}$$

$$\times\ 17 \text{ pipe sections} \times \frac{20 \text{ ft}}{\text{section}} \times \frac{7.48 \text{ gal}}{ft^3} = 13{,}388 \text{ gallons}$$

TABLE B.17. Supporting Calculations for Release from Storage Tank Outlet Line with Alternative Design of a Double Ply Expansion Joint (See Figure B.19)

		λ (Faults/Hr)	τ or t (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Outlet line ruptures (5 sections)	$(5)(1 \times 10^{-10})$		
2a	First expansion joint ruptures	1×10^{-7}		
2b	Second (outer) expansion joint ruptures[a]	1×10^{-7}	t = 8760	8.8×10^{-4}
3	Outlet line block valve ruptures	1×10^{-9}		
4	Low temperature detector does not detect spilled LNG in diked area[b]	1×10^{-4}	1000	1×10^{-1}
5	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
6	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
7	Operator fails to respond[d]			1×10^{-2}
8	Gas detector does not detect combustible gas in diked area[b]	4×10^{-5}	1000	4×10^{-2}
9	Detection control module does not operate as designed[b]	3×10^{-5}	1000	3×10^{-2}
10	Detection alarm does not sound when activated[c]	5×10^{-5}	80	4×10^{-3}
11	Operator does not detect release[d]			1×10^{-1}
12	ESD circuitry fails to deenergize when activated[b]	1×10^{-7}	1000	1×10^{-4}
13	Internal tank valve fails to close when activated to close[e]	5×10^{-5}	4000	2×10^{-1}

TABLE B.17. Contd

Calculations:

14 ENF(expected number of failures) = 1 + (2a * 2b) + 3
$= (8760\ hr/yr)^{(f)} [5 \times 10^{-10}/hr + (1 \times 10^{-7}/hr * 8.8 \times 10^{-4}) + 1 \times 10^{-9}/hr] = 1.4 \times 10^{-5}/yr$

15 Release is not isolated given Large Release Occurs = [(4 + 5 + 6) (8 + 9 + 10) (11) + 7 + 12 + 13] = $[(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3}) (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3}) (1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4} + 2 \times 10^{-1}] = 2.1 \times 10^{-1}/demand)$

TOP EVENT = 14 * 15 = $(1.4 \times 10^{-5}/year) (2.1 \times 10^{-1}/demand)$
$= 2.9 \times 10^{-6}/year$

(a) The mission time for the second (outer) expansion joint is assumed to be one year = 8760 hours.
(b) Detectors, control modules, and ESD circuitry are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(c) Alarms are tested weekly for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(d) Faults per demand.
(e) Internal tank valve is tested once a year for an assumed combined detection plus repair time of 8000 hours and a mean dead time of 4000 hours.
(f) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

Release Calculations Tank Outlet Line Rupture

Based on a hydrostatic pressure due to 100' of LNG (from top of inner tank to level of break) causing flow speed of 80 ft/sec or flow through 12" pipe of 470 gallons/second:

- Assuming a one minute release:
  28,000 gallons of LNG released.
- Assuming a 10 minute release:
  280,000 gallons of LNG released.

TABLE B.18. Supporting Calculations for Release from Storage Tank Outline Line (Bottom Withdrawal) with Alternative Design of a Block Valve Upstream as well as Downstream of the Expansion Joint (See Figure B.20)

		λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:				
1	Outlet line ruptures (5 sections)	$(5)(1 \times 10^{-10})$		
2a	Outlet line expansion joint ruptures	1×10^{-7}		
3a	Outlet line upstream block valve ruptures	1×10^{-9}		
3b	Outlet line downstream block valve ruptures	1×10^{-9}		
4	Low temperature detector does not detect spilled LNG in diked area[a]	1×10^{-4}	1000	1×10^{-1}
5	Detection control module does not operate as designed[a]	3×10^{-5}	1000	3×10^{-2}
6	Detection alarm does not sound when activated[b]	5×10^{-5}	80	4×10^{-3}
7	Operator fails to respond[c]			1×10^{-2}
8	Gas detector does not detect combustible gas in diked area[a]	4×10^{-5}	1000	4×10^{-2}
9	Detection control module does not operate as designed[a]	3×10^{-5}	1000	3×10^{-2}
10	Detection alarm does not sound when activated[b]	5×10^{-5}	80	4×10^{-3}
11	Operator does not detect release[c]			1×10^{-1}
12	ESD circuitry fails to deenergize when activated[ ]	1×10^{-7}	1000	1×10^{-4}
13	Internal tank valve fails to close when activated to close[d]	5×10^{-5}	4000	2×10^{-1}

TABLE B.18. Contd

	λ (Faults/Hr)	τ (Hr)	$\bar{a}$ (Probability)
Basic Events:			
14 Block valve before the expansion joint fails to close[a]	5×10^{-5}	1000	5×10^{-2}
15 Block valve after the expansion joint fails to close[e]	5×10^{-5}	500	2.5×10^{-2}

Calculations:

16 Cut Set for Outlet Line Rupture

This assumes that an outlet line rupture is downstream of all the expansion joint and block valves.

$(1)\ [((4) + (5) + (6))\ ((8) + (9) + (10))\ ((11)) + (7) + (12) + ((13) * (14) * (15))] = (5 \times 10^{-10}/\text{hr})\ (8760\ \text{hr/yr})\ [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3})\ (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})\ (1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4} + (2 \times 10^{-1} * 5 \times 10^{-2} * 2.5 \times 10^{-2})] = 5.0 \times 10^{-8}/\text{yr}$

17 Cut Set for Expansion Joint Failure

$(2)\ [((4) + (5) + (6))\ ((8) + (9) + (10))\ ((11)) + (7) + (12) + ((13) * (14))] = (1 \times 10^{-7}/\text{hr})\ (8760\ \text{hr/yr})\ [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3})\ (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})\ (1 \times 10^{-1}) + 1 \times 10^{-2} + 1 \times 10^{-4} + (2 \times 10^{-1} * 5 \times 10^{-2})] = 1.8 \times 10^{-5}/\text{yr}$

TABLE B.17. Contd

Calculations:

18 Cut Set for Upstream Block Valve Rupture

This assumes that once a block valve ruptures the fact that it could still close has no effect on system.

$$3a\ [(4 + 5 + 6)(8 + 9 + 10)(11) + 7 + 12 + 13)]$$
$$= (1 \times 10^{-9}/hr)\ (8760\ hr/yr)\ [(1 \times 10^{-1} + 3 \times 10^{-2} + 4 \times 10^{-3})$$
$$(4 \times 10^{-3})\ (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})\ (1 \times 10^{-1})$$
$$+ 1 \times 10^{-2} + 1 \times 10^{-4} + 2 \times 10^{-1}] = 1.8 \times 10^{-6}/yr$$

19 Cut Set for Downstream Block Valve Rupture

This assumes that once a block valve rupture the fact that it could still close has no effect on the system.

$$3b\ [(4 + 5 + 6)(8 + 9 + 10)(11) + 7 + 12 + 13$$
$$* 14)] = (1 \times 10^{-9}/hr)\ (8760\ hr/yr)\ [(1 \times 10^{-1} + 3 \times 10^{-2}$$
$$+ 4 \times 10^{-3})\ (4 \times 10^{-2} + 3 \times 10^{-2} + 4 \times 10^{-3})\ (1 \times 10^{-1})$$
$$+ 1 \times 10^{-2} + (2 \times 10^{-1} * 5 \times 10^{-2})] = 1.8 \times 10^{-7}/yr$$

$$\text{TOP EVENT} = 16 + 17 + 18 + 19 = 5.0 \times 10^{-8}/yr + 1.8 \times 10^{-5}/yr$$
$$+ 1.8 \times 10^{-6}/yr + 1.8 \times 10^{-7}/yr = 2.0 \times 10^{-5}/yr$$

(a) Detectors, control modules, ESD circuitry, and the upstream block valve are on a three month test interval for an assumed combined detection plus repair time of 2000 hours and a mean dead time of 1000 hours.
(b) Alarms are on a weekly test interval for an assumed combined detection plus repair time of 160 hours and a mean dead time of 80 hours.
(c) Faults per demand.
(d) The internal tank valve is tested once a year for an assumed combined detection and repair time of 8000 hours and a mean dead time of 4000 hours.
(e) The block valve downstream of the expansion joint is assumed to be always closed except for the ∿20 days/year when vaporizaton is assumed to occur. Thus the block valve has a mission time of ∿20 days or an assumed time of 500 hours to develop a fault not to close.
(f) Multiplication by 8760 hours per year is to convert the λ = faults/hr to λ = faults/yr.

Release Calculations Tank Outlet Line Rupture

Based on a hydrostatic pressure due to 100' of LNG (from top of inner tank to level of break) causing flow speed of 80 ft/sec or flow through 12" pipe of 470 gallons/second:

- Assuming a one minute release:
  28,000 gallons of LNG released.
- Assuming a 10 minute release:
  280,000 gallons of LNG released.

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16. Abstract

The purpose of this study is to provide an analysis of release prevention systems for a reference LNG peakshaving facility. An overview assessment of the reference peakshaving facility, which preceeded this effort, identified 14 release scenarios which are typical of the potential hazards involved in the operation of LNG peakshaving facilities. 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 the reference peakshaving facility. In addition, the effectiveness of release prevention, release detection, and release control systems were evaluated.

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