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Assessment of Research and Development (R&D) Needs in LPG Safety and Environmental Control

May 1982

Prepared for the U.S. Department of Energy
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Pacific Northwest Laboratory
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ASSESSMENT OF RESEARCH AND DEVELOPMENT (R&D) NEEDS IN LPG SAFETY AND ENVIRONMENTAL CONTROL

PNL Project Coordinator

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Under Contract DE-AC06-76RLO 1830

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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 assessment of liquefied petroleum gas (LPG) safety and environmental control benefited from the technical direction and guidance provided by Dr. John M. Cece and Dr. Henry F. Walter of the ESED.

Completed effort in other tasks of the PNL project includes an overview study of release prevention and control systems in liquefied natural gas (LNG) facilities and an assessment of research and development (R&D) needs in the safety and environmental control of ammonia plants and systems. The results of these studies are reported in:

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

Work in progress includes more detailed studies of topics identified in the LNG scoping assessment as being worthy of further investigation. Other reports of this series are in preparation on the following subjects:

- Import Terminal Release Prevention Analysis
- Peakshaving Plant Release Prevention Analysis
- Storage Tank Analysis
- Fire Prevention and Control Assessment
- Human Factors in LNG Operations.

STUDY/REPORT CONTRIBUTORS

This project and final report have involved the efforts of PNL staff and three subcontractors. Battelle Columbus Laboratories (BCL) provided a review of LPG phenomenology and related R&D; systems descriptions of LPG transportation by pipeline, rail and truck; a description of consumer storage and use and of the consumer market and a compilation of the properties of LPG. BCL also contributed a state-of-the-art summary of LPG release prevention and control technology in the areas of pipeline, rail and consumer storage. The Institute of Gas Technology (IGT) contributed descriptions of production, import/export and peakshaving plants, as well as barge and ship transportation systems. Both subcontractors and PNL evaluated R&D needs and recommended R&D projects that address LPG safety and environmental concerns. The Applied Technology Corporation performed an independent review of the draft final report and provided additional input on many topics. PNL coordinated all efforts in this project, contributed information and data to supplement material supplied by the subcontractors, reevaluated controversial issues and prepared this final report.

Many individuals from PNL and the subcontractors assisted the preparation and publication of this report. The following listing acknowledges the contributions of principal authors and others involved in this effort.

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1.0 EXECUTIVE SUMMARY

This report documents the results of a research project undertaken by Pacific Northwest Laboratory (PNL) for the Environmental and Safety Engineering Division (ESED) in the Department of Energy (DOE) Office of the Assistant Secretary for Environmental Protection, Safety and Emergency Preparedness. The purpose of the project was to provide the ESED with background information and a data base on R&D needs in LPG safety and environmental control for use in planning and implementing LPG Subprogram activities in the DOE Liquefied Gaseous Fuels Safety Program. In addition, this assessment contains information and perspectives that may assist the decision-making and R&D planning activities of both the LPG industry and other government agencies. PNL was assisted in this assessment by the Applied Technology Corporation, Battelle Columbus Laboratories and the Institute of Gas Technology.

1.1 SCOPE

The technical content of this report includes a review of LPG characteristics, hazards and risks in general terms. This is followed by highlights of production, transportation, storage and consumer use of LPG. Two sections summarize the extent of available knowledge and capabilities that minimize the consequences of LPG hazards. The first of these reviews current knowledge of phenomena involved in LPG releases. The second summarizes the current state-of-the-art in LPG release prevention and control. The last section documents the conclusions resulting from PNL's assessment. These conclusions are presented in the form of specific recommendations for R&D projects that address knowledge gaps in LPG phenomenology and needed improvements in release prevention and control.

1.2 ASSESSMENT SUMMARY

LPG is the name given to mixtures of low molecular weight hydrocarbons which can be transported and stored in liquid form under relatively low pressures and normal temperatures. When released to atmospheric pressures, LPG vaporizes and can be used as a gas.

LP gases, if improperly handled or accidentally released, can be hazardous. In order to keep them in the liquefied state, they must be pressurized and/or refrigerated. Mishandling can result in damage from cryogenic and pressure effects. When released and converted to the gaseous state, they are denser than air and can form flammable vapor clouds. This type event can result in vapor cloud ignition or detonation, both of which can be very destructive. A special hazard is created by the potential for a boiling liquid expanding vapor explosion (BLEVE).

On a statistical basis, the risks associated with LPG production and consumption are less than many voluntary risks taken by the public such as smoking, driving and exposure to medical x-rays. This assessment developed an estimate of the annual individual risk of death resulting from LPG accidents in the range of 2×10^{-7} to 5×10^{-7}. Risks at the consumer level are the dominant contribution. LPG transportation and storage risks appear to be comparable to the risk of a person on the ground being killed by an airplane crash. Risks associated with consumer use of LPG appear similar to the risk of fatalities resulting from lightning and tornados.

The two principal sources of LPG are natural gas from oil or gas wells and crude oil. Rather substantial amounts of liquids are associated with natural gas at the well-head. These liquids, which are primarily propane, butanes, and varying amounts of other hydrocarbons, are removed at extraction facilities before the natural gas is compressed and transported. The extracted hydrocarbon fraction is marketed as LPG. If a mixture of natural gas and oil are produced by a well, the associated gas contains large concentrations of heavier hydrocarbons (i.e., propane, butanes, ethane, and smaller amounts of other hydrocarbons) which are separated from the gas and sold as LPG. LPG produced from crude oil refining processes is either found occurring naturally in small amounts or is produced as a by-product of catalytic cracking and reforming processes.

From production facilities, LPG is usually transported via pipelines, tanker trucks, rail tankers, and barge/marine tankers to storage facilities for eventual distribution to consumers. In some cases LPG is transported directly from production facilities to consumers.

LPG storage can be divided into four major categories: 1) primary storage, 2) peakshaving storage, 3) distributor storage, and 4) consumer storage. Primary storage facilities provide storage for the transportation, distribution, and utilization segments of the LPG industry. Peakshaving storage is used to accommodate fluctuations in consumer demand while maintaining a constant nominal supply of LPG. Distributor storage involves principally LPG stored in tanks owned by retail distributors that supply individual customers. Consumer storage covers a range of quantities from a few pounds to in excess of 2000 gallons. This category includes the bottled gas cylinders and containers used domestically and in other applications where portability is required.

LPG is used in numerous ways by residential, commercial/industrial, and agricultural customers. The compressibility of the gas and its availability in liquid form, stored in portable containers make it an ideal fuel for residential uses such as space heating and cooling, water heating and refrigeration. Transportability via road and rail tankers, barges and ships, and pipelines enhances LPG's use as a fuel in commercial/industrial applications such as space heating, cooking and baking, and process steam and heat production (e.g., steam boilers, heat treating furnaces, drying and firing of ceramic products). LPG is increasingly used as a feedstock for the petrochemical industry. In the agricultural sector, LPG is used as a fuel for residential and farm building heating and farm equipment.

If safety devices or precautions fail, LPG may be released and dispersed into the environment as liquid or vapor. Development of effective release control equipment and practices depend on knowledge of release progressions and consequences. Unfortunately, information about LPG release phenomenology is not as available as for other liquefied gaseous fuels such as liquefied natural gas (LNG). For this reason, much of the evidence presented is based on knowledge and experience gained with LNG. Enough similarity exists between LNG and LPG to make this information useful. The broad topics of LPG vapor generation and dispersion, pool fires, and behavior of combustible LPG mixtures are reviewed, and under each of these topics, discussions are presented for specific areas of interest such as physical description of important associated phenomena, existing analytical and experimental studies with LPG and other dense gases, and related work undertaken by the DOE and other researchers.

Release prevention and control encompass all activities that tend to prevent the occurrence of uncontrolled LPG spills or, when a release does occur, the application of techniques and action designed to minimize the consequences of the spill. Optimizing release prevention and control efforts involves the activities of designers, constructors, installers, owner/operators, regulators, consumers, and the general public. This report reviews the state-of-the-art in release prevention and control as currently practiced in LPG production, transportation and consumer storage.

1.3 RECOMMENDATIONS

Based on the assessments contained in preceding sections, Section 10 presents specific recommendations for R&D projects that address knowledge gaps in LPG phenomenology and needed improvements in release prevention and control. The recommended projects are grouped into the following five topical areas:

- LPG release phenomena
- accident data collection and analysis
- assessment and analysis supporting regulatory decision making
- system and device improvements
- procedure development and human factors.

Within these five topical areas, a total of 33 distinct projects are recommended. These recommendations are listed in Table 1.1 and are briefly summarized below.

TABLE 1.1. Summary of R&D Needs in LPG Safety and Environmental Control

TOPICAL AREA	SPECIFIC AREA FOR STUDY
LPG Release Phenomena	• Planning relevant to information needs
	• Near-field flow and dispersion
	• Far-field dispersion
	• BLEVE initiation phenomena
	• Fireball phenomena
	• Pool fire radiation
	• Burning LPG vapor cloud deflagration rates and pressure effects
	• LPG vapor cloud detonation initiation
Accident Data Collection and Analysis	• Reporting requirements for LPG releases in production facilities
	• Develop improved DOT/OPSR report form for pipeline incidents and expand OPSR program to include computerization of accident data and analysis of accident causes
	• LPG truck accident and damage data
	• Survey design features of LPG import/export terminals and collect information on unreported accidents
	• Survey accident experience in peakshaving facilities
	• Collect and analyze data on accidents involving consumer storage and appliances
	• Collect and analyze information on accidents which have occurred at temporary or emergency LPG installations
Assessment and Analysis Supporting Regulatory Decision Making	• General safety design and practice at LPG facilities
	• Assess the need for upgrading LPG bulk storage practices at refineries, gas extraction plants and terminals
	• Review hazards, accident experience, and existing practices for storage of LPG at peakshaving plants to determine if requirements need to be changed
	• Review and analyze recent railroad accident data to determine incremental value of tank car safety appliances
	• Analyze cost/benefit ratios of improvements to LPG ship steering gear and control systems
	• Conduct a non-site-specific risk assessment for LPG maritime transportation (both ship and barge) to assist in decision- and rule-making
	• Review available information and determine actual incidence history and potential for BLEVE in consumer storage installations
System and Device Improvements	• Conduct state-of-the-art study to determine an optimal integrity assessment procedure for LPG pipelines
	• Investigate the extent and nature of soil movement in the vicinity of LPG storage tanks and associated piping at refineries, gas plants and similar facilities
	• Investigate possible improvements in LPG tanker truck appurtances
	• Investigate possible improvements to reduce LPG tank truck fuel systems vulnerability to accident damage
	• Investigate the severity of the "sloshing" problem in partially-filled LPG truck tanks and determine mitigation measures
	• Develop an economical and effective LPG leak detector/alarm for consumer use
	• Review and recommend possible changes in requirements for relief-valve pressure settings on consumer storage tanks
Procedure Development and Human Factors	• Review operating practices in peakshaving plants and determine if improved procedures are needed
	• Develop a broad-based safety education and training program for LPG truck drivers, management and emergency response personnel
	• Determine most effective and safe methods for handling derailed LPG cars
	• Investigate the need for improved procedures for handling LPG spills

1.3.1 LPG Release Phenomena

Possibly as much as 80% of our current knowledge of LPG release behavior is based on the extrapolation of LNG information. The following recommendations complement relevant research on LNG that is expected to continue. The development of a "relevance matrix" methodology is also recommended for planning and establishing the priority of R&D on LPG release behavior.

A model is needed for the gravity- and wind-induced motion of LPG vapor to account for near-field flow and dispersion. The model should specifically incorporate, as far as possible, factors relating to the difference in density between the atmosphere and the vapor cloud, and should account for weather factors, topography, obstacles, and vapor generation rates. An improved description of vapor cloud drift and flow could help in defining safe distances between possible LPG accident sites (e.g., transfer and storage sites, railroad grade crossings, etc.) and ignition sources in the neighborhood.

Models used for LPG far-field dispersion have been derived from LNG and other gases. Assumptions in some of these models, such as neutral buoyancy, do not apply to LPG. Current studies on LPG should provide far-field dispersion models that can be modified for LPG use to give concentrations as accurately as may be required for plant siting and safety considerations.

BLEVEs are perhaps the most serious manifestation of failure in release prevention and control systems. The rapid vaporization of LPG involves two-phase fluid mechanics and is a very complex phenomenon. In general, work to reduce the occurrence of BLEVEs must first consider changes in tank construction materials, insulation materials and safety valve capacity. These approaches do not require extensive knowledge of LPG behavior. However, critical two-phase flow of a flashing liquid is still poorly understood. The study of two-phase flow in safety valves under BLEVE-initiating conditions could lead to desirable improvements in safety valve design.

Extensive literature already exists on the subject of radiation from LPG pool fires, and fireball formation mechanisms are fairly well understood. While current knowledge appears adequate for establishing isolation distances for structures and materials that may be exposed to LPG fires, existing analytical models do not completely describe the complex phenomenon of a free-burning LPG fire. There are additional areas where new knowledge may lead to improvements in fire, injury and damage control strategies. The mechanics of soot production and consumption should be investigated together with the effects of soot on radiation characteristics. Further pool fire tests on water should be performed for comparison with experiments on land. Radiation should be measured in several directions as a function of time. Other measurements would include fuel consumption rate, vapor and liquid composition and weather conditions. These and other relevant data available in the literature should be combined to verify and enhance existing analytical models and used to model large-scale fires. Additional work is needed to determine if available knowledge on radiation characteristics of test fires is scalable to fires that actually occur. News media films are possible sources of data that have not yet been fully utilized for this purpose.

The aerothermochemistry and dynamics of LPG combustion under controlled conditions are well understood and documented. However, before adequate predictions can be made of flame propagation rates and pressures that might result from the ignition of LPG vapor clouds, further experimental data are needed under conditions simulating possible or likely spill scenarios. As with LNG, these experiments should be conducted in semi-confined geometries to simulate actual spill conditions. Further investigations are required to extend recently performed work on flame propagation in channels. Flame speed and pressure rise should be determined as a function of fuel type, fuel/air gradient and external wind direction and velocity.

It is known that the energy release requirements to initiate detonation in propane-air mixtures are less than those required to produce detonation in methane-air mixtures. However, there are few specific data on the energy required for stoichiometric mixtures of propane or butane in air, and data for oxygen-nitrogen mixtures probably cannot be reliably extrapolated. Also, no direct data are available on the conditions necessary to detonate LPG-air mixtures exiting from a duct or in an open system. It is also debatable whether a detonation once started will be always self-sustaining. Finally, there is considerable evidence that a disproportionate lowering of the critical limit of a pure fuel occurs when a small amount of more easily detonated fuel is added. Experimental approaches are required to investigate these areas of uncertainty to guide the development of LPG fire and damage control strategies.

1.3.2 Accident Data Collection and Analysis

The collection and analysis of accident data are essential first steps in the development of a sound basis for planning R&D activities, developing codes and standards and making regulatory decisions. Experience in other areas has demonstrated that accident records provide a valuable means of identifying hazardous conditions and accident causes. Unfortunately, there is a general lack of LPG accident information available as a coherent and consistent body of data. With some notable exceptions such as LPG transportation, much of the existing accident information fails to meet many of the basic criteria of usefulness and is not readily accessible. This lack precludes the objective assessment of safety-related R&D needs and, in some cases, even a determination of whether R&D effort is necessary. The situation arises in the U.S. principally because no single agency has overall responsibility for LPG safety and environmental control. There is a need for additional accident data and analysis in each major area of LPG operations, including production facilities, import/export terminals, peakshaving plants, all modes of transportation, consumer storage and appliances, and temporary installations.

1.3.3 Assessment and Analysis Supporting Regulatory Decision-Making

For an established organization such as the LPG industry, the purpose of further regulations should be to minimize, at reasonable cost, the frequency and consequences of accidental LPG releases. As technology improves and situations develop or change, it appears prudent for regulatory agencies to review the efficacy of existing regulations before considering additional rules. The following assessments are recommended to provide information that may assist future decisions relating to the development of both industry standards and new regulations.

Present safety standards and regulations for LNG are far more restrictive than those pertaining to LPG. Yet there are indications that handling LPG involves greater risks. In the U.S. a large amount of work has been done in recent years establishing the need for and implementing 49 CFR Part 193 - Liquid Natural Gas Facilities: New Federal Safety Regulations. No comparable standards have been prepared for LPG. Much of what has been done in examining LNG safety is a useful basis for planning improvements in LPG standards and regulations.

Safety standards for LPG tanks located in large tank farms such as refineries, gas extraction plants and terminals often follow oil tank safety standards more closely than LNG tank requirements. These standards do not appear to adequately account for the hazards of potential LPG releases. LNG tank design, diking requirements, spacing and isolation requirements, as well as potential risks of current LPG practice, should be reviewed to determine a basis for upgrading LPG bulk storage practices and requirements. Similar consideration should be given to storage tanks at peakshaving plants.

Recent efforts to improve the safety of LPG tank cars have mandated the use of "shelf couplers," head shields and thermal insulation. However, there is continued public pressure to improve the safety of all transportation systems. In this environment of increasing safety consciousness, there is a risk that new safety appliances for rail tank cars will be developed without the benefit of risk assessment to evaluate their incremental worth. It is thus suggested that risk assessment techniques be applied to evaluate the current worth of safety appliances and potentially new safety device developments that are contemplated. Cost estimates could also be made for implementing new safety requirements. This study will provide up-to-date risk assessment-based insight on the incremental value of safety appliances on railroad tank cars and the cost of implementing these safety-related changes. This information would provide guidance for establishing the nature and priority of future safety appliance development.

With projected increases in LPG traffic, higher standards may be required for certain conventional ship systems to minimize the possibility of ship collisions and groundings. Stricter regulations on navigation equipment, collision avoidance systems, steering apparatus, and continuity of electric and propulsive power appear to be needed. Expected increases in the numbers and movements of LPG ships and barges combined with construction of new terminals and the expansion of existing facilities suggest that greater emphasis needs to be placed on risk assessment. The majority of risk analyses done so far have been made in connection with LNG rather than LPG facilities and operations. Many of these studies have been site-specific and have been quite inconsistent in their methodology. An LPG maritime risk assessment covering a range of site-dependent parameters is needed to update previous efforts and provide a future tool for regulatory decision-making in this area.

The BLEVE phenomenon typical of LPG pressure vessels exposed to fire can cause casualties comparable to those of vapor cloud explosions. There is no information available on how severe this problem is with consumer-type installations. The frequency of BLEVE occurrence in consumer storage should be established. Results of such a study should indicate whether changes are appropriate in related codes, regulations or inspections and enforcement procedures.

1.3.4 System and Device Improvements

At the present time, structural integrity cannot be determined with high confidence for either a liquid or gas pipeline. Currently, there are several methods in use to assess the condition of operating pipelines that have various limitations or place higher than desirable stresses on the system. It is recommended that a state-of-the-art study be conducted to determine an optimal integrity assessment procedure for LPG pipelines. Appropriate laboratory and field R&D should be conducted to confirm the results of this assessment.

Tank and equipment foundations are not always designed with adequate considerations for soil movement. While tanks and connected components may be built on independent but adequate foundations, soil movements between the foundations may strain the connected piping and devices. Industry experience in various climates and soil conditions and with various industrial piping systems should be examined to determine the extent and nature of movement over time that has occurred in existing refineries, gas plants, and similar facilities. The nature and extent of movement should be examined and techniques for preventing strain in piping and components should be reported. This information would be a basis for recommendations to be incorporated in facility design standards and practices, as required.

Available statistics for LPG tank truck accidents indicate a large incidence of tank leakage after overturns. To reduce the incidence of tank leakage in overturn accidents, R&D efforts should be initiated to improve the design of tank appurtenances. The objectives of this work would be to reduce the leakage potential of fittings, vents and valves and develop a standard overturn qualification test for the whole system. One condition contributing to overturn accidents is sloshing of the LPG in partially filled trailers that interfered with driver control. The severity of this problem should be investigated and anti-sloshing measures developed.

In a highway accident, there is a potential for fire resulting from failure of the truck fuel system even though the LPG tank and trailer are initially undamaged. Fuel system design requirements for LPG trucks should be reviewed with the objective of developing improved systems, particularly in more vulnerable areas such as the crossover line.

A number of LPG explosions have been caused by the ignition of LPG leaking from tanks and fuel lines. Studies of safety problems associated with LPG tanks indicate that this is the primary hazard associated with LPG containers used for consumer storage. Currently an odorant is used for leak detection. This, however, is ineffective in many cases. A project should be initiated to develop an economic LPG leak and alarm detector for consumer use. Some types of natural gas alarms are available and should be investigated to determine their adaptability for this purpose. Also, a study to review and recommend possible changes in requirements for relief valve pressure settings on consumer storage tanks is needed.

1.3.5 Procedure Development and Human Factors

In all aspects of handling LPG, as with other hazardous materials, human factors effects represent a generally underexplored area in which the application of established human factors engineering (HFE) principles can further and profitably minimize risk. Recommendations in this section address needs for procedure development and HFE in peakshaving operations, transportation and accident response procedures.

Peakshaving plants may be left unattended during most of the year except during the time LPG is delivered. Many are operated only ten to twenty days per year during extremely cold weather. Deficiencies discovered during plant operations are apt to be tolerated for the short operating period rather than being serviced immediately under possibly severe weather conditions. Service personnel may be required on short notice to leave other work and assist in repairs. This can result in the use of probably less experienced personnel working unsupervised on potentially hazardous assignments. A review of the frequency of these practices may identify the need for improved safety equipment, operating procedures and training of personnel.

Errors by tank truck drivers are frequent causes of accidents having the potential of releasing LPG in the public environment. While some educational efforts are under way, these activities do not appear adequate as an organized approach for reaching all persons responsible for safety in LPG truck transportation. Projects should be initiated to develop a broad-based safety education and training and qualification program for LPG truck drivers, maintenance, management and emergency response personnel.

Pressurized tank cars damaged in derailments may not fail immediately, but can subsequently rupture as a result of mechanical stresses imposed during the accident cleanup process or in later operations. In view of the fact that the failure pressures of damaged tanks cannot be accurately estimated and that the measurement of controlling parameters is difficult, it is recommended that research be conducted on methods for handling derailed LPG tank cars and training emergency response personnel.

Spilled LPG is often removed by flushing with water as in the case of gasoline spills. While this method is effective at the scene of the release, there is the risk that LPG may reach ignition sources after flowing through drains and sewers. The potential hazard of ignition of spilled LPG is a serious problem, particularly since it may expose third parties not involved in or even aware of the spill. Available accident reports should be reviewed to determine the frequency of these ignitions and the time and distances involved in the migration of spilled vapors. Both gasoline and LPG spills should be reviewed to make comparisons of the relative hazards. Based on this comparison, the need for improved practices should be evaluated, including spill containment and treatment methods and emergency response training and procedure development.

1.4 OVERVIEW

The recommendations included in this report represent a measured approach to filling knowledge gaps in LPG release phenomenology and improving release prevention and control practices. A finite number of problems associated with LPG safety and environmental control appear to be worthy areas for further R&D. New projects are recommended only when there is a need to extend and complement existing knowledge and proven practices. A coordinated effort by elements of the LPG industry in collaboration with appropriate agencies of government is a reasonable approach for addressing these problems.

2.0 INTRODUCTION

Liquefied petroleum gas (LPG) is a clean-burning residential and industrial fuel and a feedstock for many chemical processes. The commonly used abbreviation LPG includes propane, butane and various mixtures of propane and higher hydrocarbons. The production and use of LPG provide direct and indirect benefits that improve the quality of life for individuals in our society. These benefits are accompanied by some level of risk because LPG is potentially hazardous when accidentally released or otherwise improperly handled. LPG continues to be involved in sometimes fatal, destructive and spectacular accidents. Some of these accidents, especially those associated with transportation, occur in public places. News media accounts of these incidents tend to dwell on the hazard potential with little or no consideration of the large volume of LPG that is handled and consumed safely. It is, therefore, understandable that public pressure for more protection against the potential consequences of LPG accidents has grown in recent years. While much has already been done to minimize these risks, there is a consensus in the industry and in government that more safety-related research and development (R&D) should be undertaken. The U.S. Department of Energy (DOE) is one of the government agencies working on LPG safety-related problems.

The DOE Office of the Assistant Secretary for Environmental Protection, Safety and Emergency Preparedness (EP) has the responsibility for identifying, characterizing and mitigating environmental, safety and health issues associated with the commercial use of specific energy materials. The EP Environmental and Safety Engineering Division (ESED) is responsible for assessing some of these materials, including liquefied gaseous fuels. To fulfill this responsibility, the ESED is conducting an R&D program on the safety and environmental control of liquefied natural gas (LNG), LPG, ammonia and hydrogen. The objectives are to gather, analyze and disseminate technical information that will aid future decisions made by industry, regulatory agencies and the general public on facility siting, system operations, and accident prevention and mitigation. This effort is known as the DOE Liquefied Gaseous Fuels (LGF) Safety and Environmental Control Assessment Program.[a] The LGF Program is coordinated with the related efforts of other agencies. Program and subprogram R&D is being conducted by national laboratories, universities, technical institutions and industrial research contractors.

As a contribution to the LGF Program, the ESED requested Pacific Northwest Laboratory (PNL) to prepare this report on LPG safety and environmental control issues and R&D needs. This PNL project was planned to provide the ESED with an information and data base that would assist, as needed, the planning and conduct of safety-related research in the LPG Subprogram. In addition, this assessment offers perspectives on LPG safety issues that may aid decision-making and R&D planning activities of both the LPG industry and other government agencies.

2.1 PROJECT OBJECTIVES AND SCOPE

Four major objectives were established to fulfill the requirements of this project. These objectives are:

1) characterize the LPG industry covering all operations from production to end use,
2) review current knowledge of LPG release phenomenology,
3) summarize the status of current LPG release prevention and control (RP&C) methodology, and
4) identify remaining safety and environmental problems and recommend R&D strategies that may mitigate these problems.

2.1.1 Industry Characterization

The industry description encompasses processing plants, major storage facilities, such as peakshaving plants; and pipeline, truck, rail and water transportation. These are characterized in terms of their number, location,

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[a] Throughout this report, referred to as the LGF Program or the DOE Program.

process variations, general design and operation. Intermediate and small storage installations and containers utilized by wholesale outlets, retailers, industrial users and domestic consumers are described, together with typical consumer applications.

2.1.2 Knowledge of LPG Release Phenomenology

Current knowledge of LPG release progressions and consequences is divided into three major categories: (1) vapor generation and dispersion, (2) pool fire phenomena, and (3) behavior of combustible mixtures. This review includes a description of LPG spread, vaporization and dispersion. The discussion of pool fires covers vaporization, mixing with air, combustion and radiation effects. The review of vapor cloud combustion includes deflagration, detonation, fireballs and flameless explosions. This assessment summarizes both analytical and experimental work conducted to date in these areas.

Because LPG safety issues have not been researched as thoroughly as those of other LGFs, much of the evidence presented in this section is actually based on knowledge and experience gained with LNG.

2.1.3 Release Prevention and Control State-of-the-Art

This assessment focuses on the approaches currently used to protect the public from accidental LPG releases. The emphasis of this effort is in areas where LPG is in the public domain. A summary of release prevention and control methodology is provided for pipelines, production facilities, railroad and truck transportation and consumer use. This summary does not consider RP&C for LPG in the marine environment, because of the substantial coverage of this topic by the Applied Technology Corporation (ATC) in other work contributing to the DOE Program (Martinsen et al. 1980).

2.1.4 Identified Problems and Recommended R&D

Specific problems relating to LPG safety and environmental control are identified and established on the basis of existing conditions and the effectiveness of current RP&C practices. Each problem assessment includes the recommendation of specific R&D project activities that are anticipated to reduce the severity of the problem.

2.2 PROJECT CONTRIBUTORS

This project and final report have involved the efforts of PNL staff and three subcontractors.[a] Battelle Columbus Laboratories provided a review of LPG phenomenology and related R&D; system descriptions of LPG transportation by pipeline, rail and truck; a description of consumer storage and use and of the consumer market and a compilation of the properties of LPG. BCL also contributed a state-of-the-art summary of LPG release prevention and control technology in the areas of pipeline, rail and consumer storage. The Institute of Gas Technology contributed descriptions of production, import/export and peakshaving plants, as well as barge and ship transportation systems. Both subcontractors and PNL evaluated R&D needs and recommended R&D projects that address LPG safety and environmental concerns. The Applied Technology Corporation (ATC) performed an independent review of the draft final report and provided Appendix I and additional input on many topics. PNL coordinated all efforts in this project, contributed information and data to supplement material supplied by the subcontractors, reevaluated controversial issues and prepared this final report.

2.3 ORGANIZATION OF THIS REPORT

The intrinsic nature of this project resulted in a copious collection of data and descriptive material. This lengthy report still represents a substantial condensation of the collected information. The report organization is intended to offer the reader answers to the following questions:

- What is the nature of LPG and why should we be concerned about its potential hazards?
- How is the LPG production and distribution industry organized?
- How much has been already accomplished in understanding and controling the hazard characteristics of LPG?

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[a] Contributions to each section by individuals in these organizations is acknowledged on Pages v, vi, vii and viii.

- Are there safety and environmental control problems or knowledge gaps that remain to be addressed?
- What additional R&D is recommended to address these problems?

The balance of this report addresses these questions taken in order.

Section 3 contains a brief overview of LPG characteristics, hazards and risks. The industry description occupies Sections 4 through 7 covering LPG production, transportation, storage and end use, respectively. A summary of current knowledge and ongoing effort in the area of LPG release phenomenology is presented in Section 8. The state-of-the-art in release prevention and control is reviewed in Section 9.

The crux of this report is Section 10 which recommends and justifies 33 specific projects that address knowledge gaps in LPG phenomenology and R&D needs in release prevention and control. Information sources used in this study are identified in the Reference Section. Tabular data and statistics and representative LPG accident case histories are included in Appendices A through H. Finally, Appendix I provides a preliminary analysis of LPG pool fire data, contributed by Welker, Johnson and Cavin of the ATC. The effort was performed under subcontract to assist the PNL assessment and to extend previously published ATC analyses of LPG fire phenomena.

3.0 LPG CHARACTERISTICS, HAZARDS AND RISKS

Liquefied petroleum (LP) gases exist as colorless flammable gases at atmospheric pressure and normal ambient temperature. LP gases primarily include propane, butane, ethane, propylene, butylene and butane-propane mixtures. These gases are utilized both as fuels and as feedstocks for petrochemical plants. Although safe for use when handled properly, the chemical properties of LP gases can result in hazardous conditions if the gases are accidentally released or are handled without proper safety controls. This section provides a brief summary of the chemical and physical properties of LPG as they relate to the potential hazards and risks involved in handling LPG.

3.1 LPG CHARACTERISTICS AND HAZARDS

Propane and butane are the LP gas products most used in the U.S. Average chemical and physical properties of these two substances are listed in Table 3.1. As discussed below, these properties are of practical significance in the safe handling and utilization of LP gases.

3.1.1 Vapor Pressure Temperature and Latent Heat

Vapor pressure of a product is important for container design and release consequence analysis. The vapor pressure is directly related to the stresses that a container undergoes during transport of the material. The vapor pressure is also a factor in determining the rate of vaporization of an LPG spill and can indicate the extent of the vapor cloud that may be formed. Because their boiling points are relatively close to ambient temperatures, LP gases can be stored in liquid form either by pressurizing them at ambient temperatures or cooling them at ambient pressures.

When stored in refrigerated form at ambient pressures, LPG temperatures are around the normal boiling point (-44°F). Liquids at this temperature can cause severe burns to human skin. Therefore, protective clothing (gloves, facemasks, etc.) should be worn by workers handling refrigerated LPG. Even though LPG at the consumer level is almost always used as a compressed gas at ambient temperatures, this type of hazard is not entirely eliminated.

TABLE 3.1. Characteristics of Commercial LP Gases

	Commercial Propane	Commercial Butane*
Chemical Formula	C_2H_8	C_4H_{10}
Boiling Point (°F at atmospheric pressure)	-44	28
Vapor Pressure (specified maximum pressure in psi)		
-40°F	20.0	--
32°F	80.0	27.9
100°F	224.7	84.9
Relative Density		
Compared to H_2O at 60°F	0.50-0.51	0.57-0.58
Compared to air at 60°F and atmospheric pressure	1.40-1.55	1.90-2.10
Ratio of Gas Volume to Liquid Volume at 60°F and Atmospheric Pressure	274	233
Latent Heat of Vaporization (BTU/lb)		
-20°F	176.0	--
0°F	171.0	--
20°F	165.0	156.0-167.0
35°F	160.0	152.0-164.0
45°F	--	150.0-162.0
Heating Value		
BTU/lb	20,700	20,450
BTU/ft^3 (dry)	2,405	3,150
Air Required for Combustion (ft^3 to burn 1 ft^3 of gas)	847	1,095
Minimum Ignition Temperature in Air (°F)**	860-1076	770-1067
Maximum Measures Flame Temperature in Air (°F)	3,500	3,450
Approximate Limits of Flammability (percent by volume of gas in air to form a combustible mixture)		
Lower Limit	2.2	1.8
Upper Limit	10.0	9.0

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*Usually a mixture of iso-butane and normal butane.

**The minimum ignition temperature is dependent on the air/fuel ratio and the pressure of the gas/air mixture.

The latent heat of vaporization represents the heat required to change the fuel from a liquid to a gaseous state (figures are given in Table 3.1 for various temperatures). When liquid LPG vaporizes, the heat required is first taken from the liquid itself. If LPG is vaporized under control or accidentally released, this causes a drop in the temperature of the liquid. The phenomenon, known as "auto-refrigeration," can bring the temperature of the liquid down to or below LPG boiling temperatures. Because it is not always realized by workers handling LPG that this can happen, the rapid cooling effect can pose a rather insidious hazard.

Such a sudden cooling could freeze water in contact with the equipment. A major accident of this type was caused at Feyzin, France in 1966 (see Appendix H) when an opened sampling valve underneath a propane tank froze up and could not be closed again. Generally, however, accidents caused by such "low temperature" effects have been rather minor and not many have been reported.

When stored above their boiling point, and particularly above ambient temperature, the LP gases exert an increasing vapor pressure, thus increasing the pressure required for liquefaction. Accidental releases of LPG from the failure of pressurized containment present a significant safety concern. When LPG is kept at ambient temperatures under pressure, vessels, pipes and valves can suddenly rupture if the material they are made of has lost its strength (e.g., due to corrosion or mishandling) or if the internal gas pressure exceeds the safe design pressure of the equipment. For protection against overpressure failures, safety relief valves are usually installed at properly selected locations.

3.1.2 Specific Gravity and Flammability Limits

The specific gravities of LP gases, both in their liquid and gaseous states, also have a bearing on their safe handling. These values are included in Table 3.1 for commercial grade propane and butane. At ambient temperatures, the relative liquid gravities of LPG vary from about 0.50 for propane to 0.58 for butane. Relative gas gravities vary from about 1.5 to about 2 for propane and butane. Thus, both are considerably heavier than air. Because of its relative specific gravity, once LPG is released from pressurized containment to the atmosphere, it will spread almost as readily as water. The vapor will displace the warmer, lighter air above it and will resist dispersion from a light wind. Liquid LPG will run downhill and along trenches.

Released LPG will diffuse into the air somewhat so that it is covered by a vapor/air mixture which is somewhere within explosive limits. The limits of flammability represent the percentage of gas required in the gas-air mixture to become flammable. Figures for the lower and upper limits are given in Table 3.1. LP gases become combustible at lower concentrations than most other gases.

The term "limits of flammability" assumes that the gas and air are thoroughly mixed. In practice, LP gas escaping into the atmosphere without ignition tends to settle and is too rich to burn (i.e., the mixture is above the upper limit concentration) except on the fringe of such a gas cloud where mixing with air occurs. When the resultant combustible mixture reaches an ignition source, it will ignite. Heat generated by the burning gas causes turbulence and further mixing so that eventually most of the gas will become mixed within the flammable range and a serious fire hazard may result. As an example, in the Port Hudson incident (see Appendix H) 66 tons of gas from a ruptured liquid propane line spread for 13 minutes in the form of a "pancake" 10-20 ft thick before igniting about 2000 ft away.

Fortunately, such large events do not occur frequently; however, a host of smaller events start the same way inside and outside of the home. It is this stealthy property of LPG which is most difficult for the layman to comprehend and for the expert to anticipate. LPG must be considered unique among hydrocarbon fuels in the combination of properties which lead to this type of hazard. In contrast, lighter hydrocarbons, such as methane, diffuse into the atmosphere, and unless the space is confined, they usually dissipate harmlessly. Heavier hydrocarbons such as gasoline are emitted as liquids and vaporize slowly.

LPG is also unique in that loss of containment (a broken line, or a damaged cylinder) will emit liquid under pressure (not liquid under gravity as with gasoline or gas under pressure as with methane). The mass of material emitted in a given time from a given sized breach will be much more than for either the lighter or the heavier hydrocarbons. LPG in the liquid phase expands considerably when its temperature rises. The coefficients of cubical expansion at 60°F are approximately 0.00089 per °F for propane and 0.0006 per °F for butane. These values are about four times the equivalent for fuel oil and ten times that for water. This high rate of liquid expansion must be taken into account when specifying the maximum quantity of LPG in any pressure vessel. Furthermore, the LPG, when released, will be emitted as a boiling liquid (i.e., a two-phase mixture of spray and gas). Its high density and high velocity will ensure entrainment of air and thus maximize subsequent mixing, making explosion more likely if ignition occurs.

The hazards associated with the release to atmosphere of LPG as with any other boiling flammable liquid show themselves disproportionately when large quantities are involved. They can give rise to two phenomena which are large loss producers; the unconfined vapor cloud explosion and the boiling liquid expanding vapor explosion.

3.1.3 Vapor Cloud Explosions

If a large quantity of flammable gas is released to the atmosphere, there are basically three possible consequences:

1. It can disperse without igniting and without any other ill effects.
2. It can ignite and burn as a deflagration. Such deflagrations or fire balls can cause considerable damage by radiation (e.g., starting secondary fires at some distance).
3. It can ignite and explode, causing high overpressures and heavy blast damage.

The last of these is by far potentially the most damaging, combining the effects of blast with those of fire. In fact, blast waves could, under certain conditions, transform into detonation waves, which are characterized by very high overpressure pulses.

As pointed out by Gugan (1981), there is a continuous spectrum of increasing overpressures leading from deflagration to detonation conditions. For most accident cases, the overpressures generated by blast waves are larger than those for pure deflagration and smaller than those for true detonation. Windebank (1980) calls the explosions with moderate to high overpressures "percussive unconfined vapor cloud explosions" (PUVCE). According to his estimate, a discharge of at least 10 tons of saturated hydrocarbons is necessary to generate a PUVCE. It is understood, however, that only a fraction of the energy contained in spilled material contributes to the blast effect. Generally, most of the energy is converted to heat.

The chances of forming a cloud of more than 10 tons of hydrocarbon, with a large part of it mixed with air within the flammability limits, are not high. The conditions for forming such a cloud require a rapid spill and evaporation before the wind causes significant dispersion of the cloud. Air turbulence must be high enough to assure mixing with the atmosphere; otherwise, the LPG vapor tends to settle in a "pancake" layer.

A liquid held under pressure above its boiling point has the best chance of meeting these requirements. Over half of the forty incidents recorded by Davenport (1977) involved material of this sort, and about half of these involved LPG in its various forms.

Data collected by the Office of Pipeline Safety Operations (OPSO) on the incidents involving LPG pipelines show over an eleven year period (1968 to 1978) 296 spillages, 9 percent of which involved more than 3500 ft^3 of liquid. Eighty-one (27 percent) of these caused fires, but only twenty-five (8 percent) were classed as explosions (i.e., less than one in three). No record was made of the overpressures experienced, and it is likely that only a portion of these would be classified as PUVCEs. According to Windebank (1980), these and other records suggest that of those LPG vapor clouds which are ignited and which are large enough to sustain PUVCEs, less than one in ten does so.

Rare though PUVCEs may be, their damage is large and sometimes enormous. Davenport (1977) lists twenty incidents over the years 1970 to 1975 with average losses of 20 million dollars in property and seven fatalities. However, only four of these were associated with LPG; the remainder were mainly chemical feedstocks including propylene and butylenes.

3.1.4 Boiling Liquid Vapor Explosions (BLEVE)

A BLEVE is a different phenomenon, but again is a consequence of holding a pressurized flammable liquid above its boiling point. In this case rupture of the container is the immediate cause.

Apart from the structural defects discussed above, an LPG container can rupture in practice by overfilling. Thus, this allows inadequate vapor space for the temperature expansion of the contents so that the vessel becomes subjected to the heavy hydrostatic load of the thermally expanding liquid. It is also possible that the yield stresses of the vessel's walls are reduced by mechanical damage as in a road accident, or by high temperature often caused by a neighboring vessel leaking and catching fire.

The failure pattern usually starts as a rupture along a longitudinal weld of an over-stressed mid-tank section. This unwraps instantaneously leaving two dished ends full of flammable liquid above its boiling point. The ejection of the boiling liquid produces a reaction which turns each end into a rocket projectile. Records document cases where tank ends have been propelled almost 1500 ft. More often than not, the liquid ignites as the cylinder fractures, providing still more energy for propulsion. Although the 'hot' BLEVE is more spectacular than the 'cold' BLEVE, both will almost always start fresh fires at some distance from the initial incident, which makes it a difficult situation to handle.

BLEVEs do not have quite the same potential for blast wave loss that the PUVCE has, but their potential for spreading fire over a wide area makes them significant in terms of both life and property. They are particularly hazardous to fire-fighters. According to Windebank (1980) the BLEVE is probably a little more frequent on a world scale than the PUVCE. Most reported BLEVE's have occurred in transportation accidents in the United States.

3.1.5 Releases from Refrigerated Storage

As explained above, both PUVCE and BLEVE basically resulted from accidents involving LPG stored as a liquid under pressure. Therefore, today when large quantities are involved, there is growing use of refrigeration to store and handle LPG as a liquid at atmospheric pressure. The hazards associated with releases from refrigerated storage are different from those with pressurized storage. Because refrigerated spills will be of cold liquid LPG, contact with the warmer surroundings will cause vaporization. However, auto refrigeration will reduce the liquid temperature and, in turn, the evaporation rate. A vapor cloud would, of course, form, but the chances are that it will form a dense vapor 'pancake', without appreciable air pre-mixing if the atmosphere is calm. If it is windy, the rather slowly forming cloud could be dispersed quickly without ever having a chance of being ignited.

If the cloud ignites, the usual flashback to the liquid would most likely occur, resulting in a pool fire. Ignition temperatures for LP gases depend on a variety of conditions, including the air/fuel ratio and the pressure of the gas/air mixture. Table 3.1 shows typical ranges for propane and butane which have been observed with air. A PUVCE, though possible, is considered extremely unlikely under the above release conditions (Windebank 1980). A BLEVE is close to impossible because there is no overpressure in the container as long as the LPG is at its atmospheric boiling point. It is most likely the LPG would simply flow out and evaporate (Windebank 1980). It is conceivable that if a pool fire is started around the leaking tank, the remaining liquid contents in the tank would be heated, thereby increasing the pressure in the tank. This could conceivably cause the leak rate to increase and provide further fuel to the pool fire as an escalating sequence. If the circumstances are such that the internal pressure rises faster than the leak can relieve them, this could lead eventually to a BLEVE. This is particularly possible if the leak is located (or becomes so during the course of the incident) above the liquid level in the container.

Similar hazard problems can arise with safety valves. It appears that circumstances always can be imagined severe enough that a given valve would not relieve the internal pressure buildup fast enough to prevent rupture of the vessel.

3.2 PUBLIC RISK IN USING LPG

The very serious LPG incidents (PUVCE and BLEVE) discussed above, can cause large losses in property and some fatalities. However, their rate of occurrence is relatively low even though their frequency and severity appear to be slowly increasing (Gugan 1981). On the average, there are about 30 such serious incidents per year with total losses reaching possibly ten million dollars. Most of these incidents happen in areas that are remote from the public. In addition to these relatively few severe accidents, there are many smaller incidents per year which are caused principally by the mishandling of LPG by or on the premises of individual consumers. On a total loss basis, these smaller incidents are estimated to outweigh the larger incidents by at least an order of magnitude.

In view of the anticipated growth of future LPG use, it is highly desirable to determine the annual risk of fatalities from all accidents involving LPG in the U.S. Unfortunately, statistics for deriving such figures are not readily available, nor is there any central authority that collects, or is responsible for such information. The following subsections provide estimates of the average individual risk (probability of death per person per year) from accidents involving the various LPG production, transportation, and consumer use operations in the U.S. These estimates are tentative and are based upon diverse data sources.

3.2.1 LPG Production Accidents

Data on the fatal accident frequency rate for LPG plant operations could not be found. It may be assumed that this rate is close to 10^{-4}/year, cited for workers in the chemical industries (Kletz 1980). However, such fatalities are due to "voluntary" risks assumed by the chemical workers, because they voluntarily chose to work in these industries. Therefore, they should not be considered here along with the "involuntary" risks incurred by the general public as a result of LPG use in this country. Risks of the latter type would be encountered by people living or working close by and by those who happen to be near a LPG plant when a large accident happens that extends beyond the plant's boundaries. No statistics could be found which would help define this particular risk rate and it is doubtful that they even exist. Public involvement in this type of scenario appears to be small.

3.2.2 LPG Pipeline Accidents

LPG pipeline accidents are rather well documented, since they have to be reported in detail to the authorities. For 1977 the average individual risk due to LPG pipeline distribution was 1.5 x 10^{-8} (NTSB 1978a). Under current conditions, it may be assumed that this is a representative figure.

3.2.3 LPG Truck and Rail Car Accidents

Because of the relatively high hazards in all types of road and rail transportation, it is to be expected that accidents involving LPG shipments are correspondingly high. Geffen (1980) conducted a study on this particular subject and estimated an average individual risk of 7 x 10^{-8} due to LPG truck and rail car accidents. This study assumed an expanded LPG trade projected for 1985; therefore, the estimate of average individual risk stated above is conservatively assumed to apply to the present situation.

3.2.4 LPG Land Transportation and Storage Accidents

As part of their work under the DOE Program, Martinsen and Cavin (1981) studied LPG land transportation (i.e., railroad tank cars, trucks, and pipelines) and storage safety. Their report contains an analytical examination of fatal accidents involving LPG releases during transportation and/or transportation-related storage. The resulting data indicate that, on the average, releases of LPG during transportation and intermediate storage cause approximately six fatalities per year to members of the general public. The individual risk is about 1 death per 37,000,000 persons or 2.7 x 10^{-8}; about the same as the risk of a person on the ground being killed by an airplane crash, and much less than the risk of death by lightning, tornadoes, or dam failures.

3.2.5 LPG Ship and Barge Accidents

So far, there have been no reported fatalities in this category including marine terminals (Poten and Partners 1981). As with public involvement in LPG production accidents, the risk to the public from this type of scenario appears small.

3.2.6 LPG Accidents at the Consumer Level

Even though the number of accidents involving the consumer of LPG may be quite high, most of them are of minor magnitude when compared with accidents in the transportation category. Therefore, the number of fatalities is not expected to be large. Wakamiya and Calvano (1977) report that the Consumer Product Safety Commission had 105 death certificates on file for LPG container accidents at the consumer level. This is estimated to apply to an 8 year period which would correspond to an average individual risk of 6 x 10^{-8}/year for the current U.S. population. This probably represents a lower bound. 6503 fatalities occurred in the U.S. in 1975 due to fires. LPG caused fires represent only a small fraction of these. Data are scarce, but if figures from Britain can be taken as a characteristic indicator, about 1.5 percent of all fires are connected with LPG (Windebank 1980). Assuming this same fraction applies to fire fatalities, an estimate of the average individual risk due to LPG accidents at the consumer level is 4 x 10^{-7}. This estimate probably represents an upper bound.

3.2.7 Overall Risk to the Public

By summing the above average individual risks, it is estimated that the current annual risk to the U.S. public of being killed in an LPG accident appears to be in the range between 2 and 5 x 10^{-7}. The risk of fatalities from lightning and tornados falls in the same range. Many voluntary risks considered generally acceptable (e.g., driving a car, exposure to medical x-rays) are orders of magnitude larger.

In conclusion, therefore, the use of LPG appears to involve much less risk than the average person is exposed to in daily life from other sources.

4.0 LPG PRODUCTION

Liquefied Petroleum (LP) gases are derived generally from two different sources. One source of LPG is from oil and gas fields, where the LPG is removed as condensable products from natural gas or from crude oils during the stabilization process which is normally applied prior to shipment to reduce vapor pressure. LP gas is also generated in most oil refineries as a byproduct of crude oil processing. Since such processing methods can vary considerably between refineries, the composition and quantity of LPG derived by this process varies accordingly.

In the U.S. approximately 70 percent of the LPG produced comes from oil or gas fields and the remaining 30 percent from petroleum refining. Before LPG reaches the consumer, it is handled by successive elements of the LPG production and distribution industry. This section provides a general description of LPG production methods and facilities.

4.1 LPG REFINING

Crude oil commonly contains a small percentage of LPG components which may be recovered during refining operations. The fraction of LPG recoverable depends on the composition of the crude oil and the refining process used. LPG produced from refining operations may contain significant amounts of sulfur which must be removed before the product is marketable as commercial LPG. Alternatives to removing the sulfur are to use the LPG components for producing gasoline and other petrochemicals, or as a refinery fuel. Refining operations that produce LPG are briefly described below.

4.1.1 Catalytic Cracking

Catalytic cracking processes are the major source of LPG in refining operations. In these processes, heavy gas oils are brought into contact with a hot catalyst and thereby converted into cracked gas, LPG, naphthas, fuel oil and coke.

4.1.2 Hydrocracking

Hydrocracking is a catalytic high pressure pyrolysis process in the presence of fresh and recycled hydrogen that produces middle distillates and gasoline from heavy gas oil. Since the hydrogen is recycled, the heavier gases produced in this process are commonly separated from the hydrogen stream.

4.1.3 Coking

Liquefied refinery gases are also produced during processing of the heaviest crude fractions in coking or visbreaking facilities. In coking, a residual fuel oil or heavy gas oil is preheated and contacted with hot carbon. This results in extensive cracking of the long chain hydrocarbons to form products ranging from methane to heating oils.

In a visbreaking facility, fuel oil is passed through externally fired tubes where liquid phase cracking reactions occur. In this process a gas mixture, consisting primarily of hydrogen, methane and ethane, and lighter fuel oil components are formed.

4.1.4 Catalytic Reforming

Catalytic reforming of hydrocarbon fractions in the C_7 or C_{10} range is a common processing step in the production of gasoline. These processes involve conversion of parafinic and napthenic hydrocarbons into aromatic hydrocarbons in the presence of hydrogen and a catalyst. Isomerization of the original straight-chain molecules to more highly branched hydrocarbons also occurs in this process. Since catalytic reforming produces recyclable hydrogen, it is common practice to separate the gases into light and heavy gas fractions including the LPG components. The gases produced in hydrocracking and catalytic reforming are commonly used in refinery processes for catalytic desulfurization in which middle distillates, ranging from light to vacuum gas oils, along with fresh and recycled hydrogen are passed over a hydrofining catalyst at pressures of 500 to

1000 psig. The process converts organic sulfur compounds to hydrogen sulfide. Some light hydrocarbons are also produced by hydrogen cracking. These light hydrocarbon gases include propane and butane.

Recovery of LPG from the mixtures of the various liquefied refinery gases involves processes similar to those in natural gas extraction plants which are described in the following subsection.

4.2 LPG EXTRACTION PROCESSES

Processes presently used for extracting LPG from natural gas all follow the same general principle. The overall concept is to convert the gaseous LPG components in an inlet gas mixture into their liquid states. The processes can generally be broken down into two basic steps. The first involves extraction of the desired LPG components from the inlet gas stream. Components that are undesirable in the finished LPG are also removed. The second step involves rejection of the undesirable components into a residual gas stream for disposal.

Various types of processing methods are used for liquefied petroleum gases. These methods include oil absorption, refrigerated oil absorption, refrigeration or compression, adsorption, and combinations of these processes.

4.2.1 Oil Absorption

The absorption-stripping cycle for the recovery of LPG and natural gasoline is widely used in the industry. Essentially, the absorption processes add a heavier component, such as a high molecular weight absorption oil (100-180 molecular weight) with properties closely resembling kerosene, to the inlet raw gas stream. This addition increases the condensation which would occur at a given pressure and temperature level. To further improve the extraction selectivity, this interaction of lean oil and gas takes place in an absorber tower. In this trayed tower, natural gas introduced into the bottom flows upward and countercurrently contacts lean absorption oil introduced into the top. As the oil and gas make contact, the lean oil absorbs the desired components from the inlet raw gas stream. Lean oil containing the absorbed gas liquids is termed "rich oil" and, after being stripped of the desired components by distillation, is recycled to the top of the absorber tower. The lean oil processes were initially used to improve extraction efficiency and to increase the range of liquid products obtainable. A flow diagram for the process is shown in Figure 4.1.

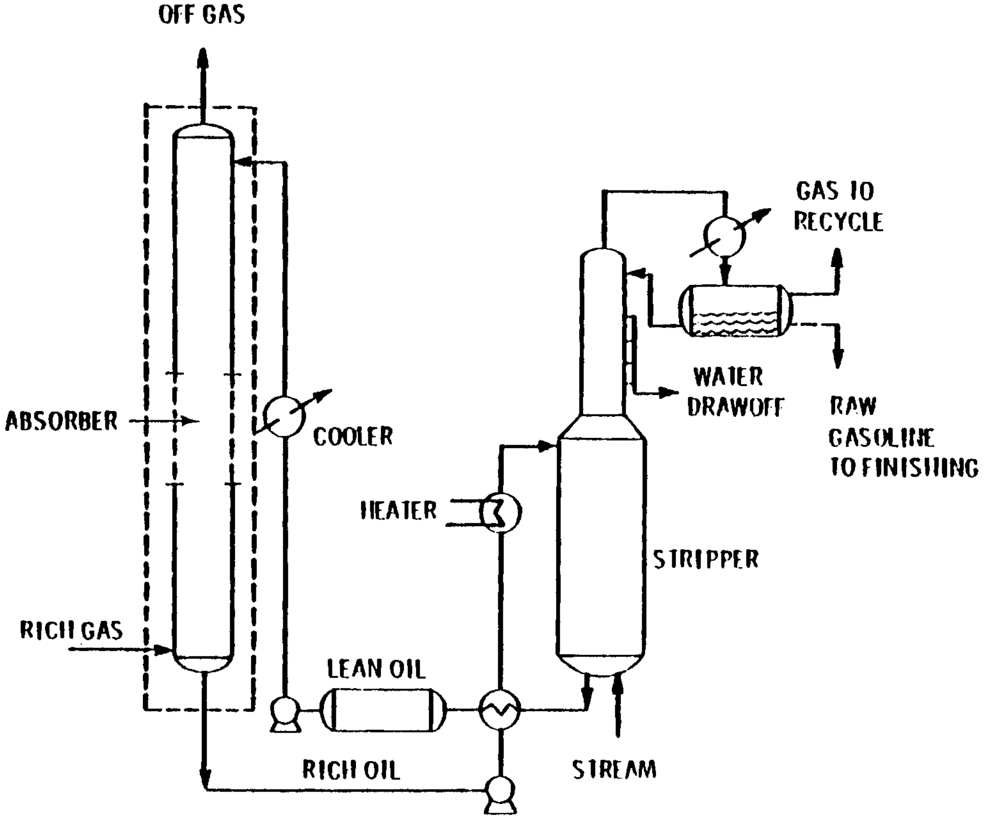

FIGURE 4.1. Oil Absorption Flow Diagram

4.2.2 Refrigerated Oil Absorption

As the demand for LPG components continued to increase, it became economically desirable to refrigerate both the lean oil and the inlet gas. The combination of this inlet gas chilling plus lean oil refrigeration improved recovery efficiency and provided for economical removal of 80% to 95% of the propane component.

In a typical refrigerated oil absorption process, the gas is first stripped of hydrogen sulfide and dried, often by extraction with a glycol at about -22°F. The glycol is recovered from the aqueous condensate by distillation and recycled. After drying and chilling, the gas flows to a demethanizing absorber. In the absorber, operated at essentially feed pressure, the feed gas is contacted with refrigerated absorption oil which is composed of natural gasoline components recovered from the gas itself. The off gas contains methane and ethane. The rich oil from the absorber, along with small quantities of dissolved ethane and methane, flows to the deethanizing absorber operating at a lower pressure. The remaining methane and ethane are used as plant fuel. Any excess is recompressed to pipeline pressure. The bottoms product flows to the debutanizer in which the recovered propane and butane are separated from the gasoline absorption oil and net gasoline product. The gasoline lean oil is cooled and returned to the demethanizer and deethanizer. The overhead propane/butane mixture is then processed in the depropanizer for final product separation. Refrigeration for the process can be supplied by a propane cycle.

4.2.3 Expansion Refrigeration

The expansion-refrigeration system uses the Joule-Thompson effect to increase the recovery of condensate and at the same time lower the water content of the gas. When high wellhead pressures permit, hydrocarbon hydrates are intentionally formed to dehydrate the gas. The temperatures achieved through refrigeration by expansion also permit a considerable increase in the recovery of natural gasoline.

In a typical process, shown in Figure 4.2, the high-pressure well stream enters the heat exchanger in the bottom of the low-temperature separator; the gas exchanges heat with the solid hydrates and cold liquids. The cooled gas stream passes through the regenerative heat exchanger where it is further cooled (when permissible from hydrate formation standpoint) by countercurrent heat exchange with cold-separator off gas. Any water condensed in the cooling process is removed in the free-water separator. The stream then passes through a pressure-reducing separator where the Joule-Thompson expansion occurs. The gas separated from the liquid and solids by density difference leaves the top of the separator. A temperature controller is set up to permit part of the cold dehydrated gas to bypass the regenerative heat exchanger when all the available refrigeration is not needed. The liquid condensate and water are removed through separate outlets from the bottom of the hydrate separator.

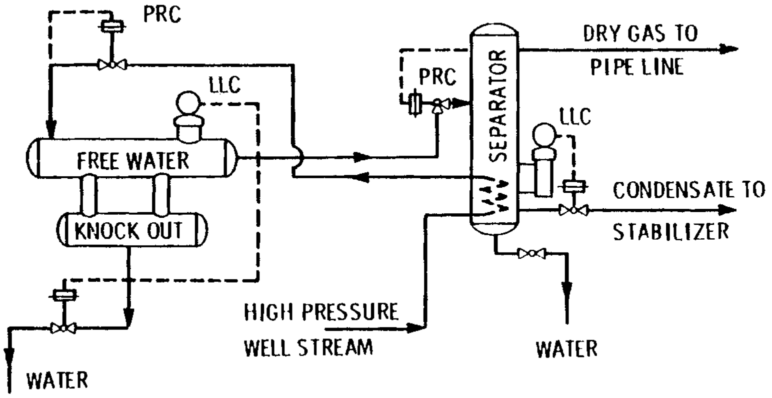

FIGURE 4.2. Expansion Refrigeration Flowchart

4.2.4 Cryogenic Refrigeration

As demand for ethane increased, it became apparent that a tremendous increase in lean oil circulation rates would be required for high ethane recoveries using oil absorption processes. For example, to increase ethane recovery level from 30% to 70% requires about twice as much lean oil at a given level of pressure and temperature. The resulting increase in investment and operating costs made it necessary to investigate other processes to improve the recovery of this LPG.

For any specific composition of the gas stream, selective levels of extraction may be accomplished by adjusting the degree of refrigeration at a fixed pressure level. A decrease in refrigeration temperature at a given pressure will increase the amount of liquid produced and will increase the recovery of each individual component of the gas stream. In practice, the simplest and most economical process in all cases is to provide only enough compression and refrigeration and/or both to effect the desired recovery of the lightest component to be recovered, manipulating pressure and temperature to minimize the compression and refrigeration costs.

In the late 1960's, processing economics favored refrigerating the gas stream to cryogenic levels. To reach these low temperatures, a combination of external refrigeration (using propane or ethane as the refrigerant) and gas expansion utilizing the Joule-Thompson effect was initially used. An improvement of this expansion process is to expand the gas through an expander turbine which is directly connected to a gas compressor or electric generator. Additional energy is extracted from the gas during the expansion, and in so doing, the temperature can be dropped to a lower level than when utilizing only the Joule-Thompson effect. At the same time, energy is recovered for recompressing the residue gas resulting from the extraction of the liquids. This process evolved into what is known today as the cryogenic process for high ethane recovery of natural gas liquids.

The turbo-expander process is the state-of-the-art cryogenic process for the recovery of liquid hydrocarbons such as butane, propane and ethane from wellhead gas streams. With this process, up to 85% ethane extraction is possible. Feed gas pressures range from 60 to 2,500 psig. Plant capacities vary from 10 to 1,200 MMscfd. Plants have been designed for residue gas pressure restoration to the full pressure of the inlet gas and higher. The inlet gas stream must be conditioned, where necessary, to eliminate lube oil, sulfur compounds and foaming agents. The CO_2 level must also be controlled.

In the process, feed gas is dehydrated with molecular sieves to less than 1 ppm of water to protect against hydrates and ice, and then it is split into two streams and chilled. One stream is chilled by heat exchange with residue gas; the other stream is used to reboil the column in one or more reboilers and/or side reboilers. Refrigeration recovery with a side reboiler makes greater extraction or horsepower savings possible. If necessary, supplemental heat can be added with a trim reboiler. Condensed liquids are then separated and fed to the middle of the demethanizer, and vapor is delivered to the expander.

Expansion is nearly isentropic and cools the gas much more efficiently than an adiabatic expansion. Direct-connecting the expander to a booster compressor recovers work from the expander. Although the booster compressor usually compresses residue gas, it can also be used to compress inlet gas. Liquids are condensed at the expander outlet and piped to a stripper or demethanizer. The ethane rich bottoms can be stripped to the desired methane content. A process flowsheet of the turbo-expander process is shown in Figure 4.3.

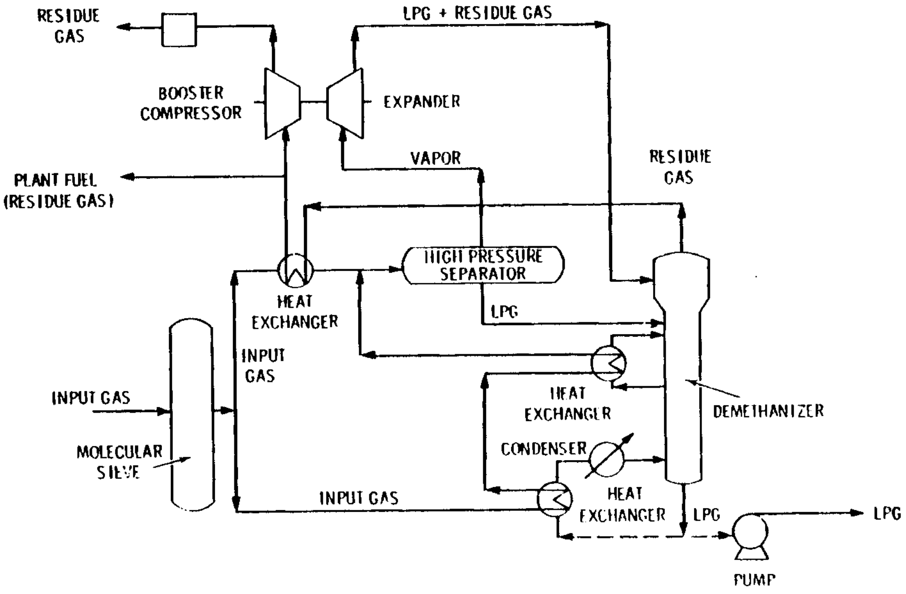

FIGURE 4.3. Turbine Expander Extraction Process Flowsheet

4.2.5 Adsorption

The adsorption process is generally used in field treatment of relatively rich gas to remove high molecular weight fractions which could cause trouble by condensing in collecting lines. Although up to 85 percent of the natural gas liquid component can be recovered using this process, adsorption type plants account for only a minor fraction of the total U.S. LPG production.

At first adsorbents used in the process were designed only to dehydrate natural gas. Design evolution has resulted in development of adsorbents for removal of select hydrocarbons (e.g., C_5^+); however, no currently available adsorbents can be used universally with optimum results.

Basically all quick-cycle adsorption processes use adsorption, heating and cooling cycles. In the process, inlet gas enters a scrubber which removes all free liquids. The gas then passes into vertical towers filled with adsorbent where hydrocarbons and water are adsorbed. Water and other polar compounds are more strongly adsorbed than hydrocarbons. Within the same chemical family, higher molecular weight compounds are more strongly adsorbed than low weight ones. Therefore, as gas passes through the bed, heavier hydrocarbons displace light ones and push them down through the tower. The cycle is switched before the water vapor adsorption zone starts displacing all condensable hydrocarbons. The adsorbent is regenerated, and liquids are recovered. Maximum liquid recovery occurs at regeneration pressures between 400 psi and 700 psi. Dry gas is returned to the pipelines.

Towers are internally insulated to reduce heat requirements, save fuel gas and permit shorter cycles. Also to conserve heat, regeneration gas is preheated by passing it through the tower which has just completed the heating cycle. This also cools the bed so it can go back into adsorption service.

The preheated stream passes through the heater and then into the tower being heated to remove all liquids from the bed. The rich stream then goes through cooling equipment to condense and recover stock tank liquids in the regeneration gas separator. Cooled gas reenters the main inlet gas stream going to the tower on adsorption.

Two- three- or four-tower arrangements with open or closed regeneration cycles are in use. A process flowchart for a three-tower short-cycle unit is shown in Figure 4.4.

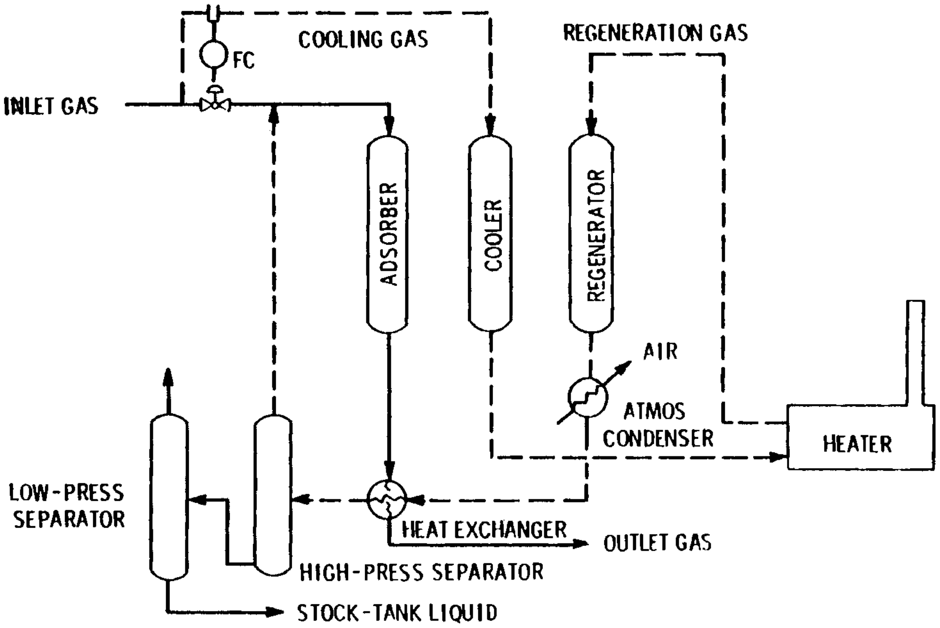

FIGURE 4.4. Three Tower Adsorption Unit Using Open-Cycle Regeneration

Other open-cycle regeneration units use heating-to-cooling regeneration flow- or tail-gas cooling regeneration flow. Closed-cycle regeneration systems are in use as well. The process chosen depends on economic and feed gas parameters.

Molecular sieves are the basis of most adsorption processes, especially those intended for dehydration only. Molecular sieves are synthetic metal aluminosilicates having a three-dimensional power structure and a one-to-three micron particle diameter. They are manufactured as pellets size 1/16 in. or 1/8 in. for easy handling and use. The size and position of the metal ions in the crystal control the effective diameter of the interconnecting channels. Commercial molecular sieves which are used in many LPG sweetening plants will absorb molecules with critical diameters of up to 10 Angstrom units (10 Å). H_2S, mercaptans, and LPG hydrocarbons fall into this category. The separation of the sulphur compounds depends therefore on selective adsorption due to the higher polarity of these compounds rather than on conventional molecular sieve action. The adsorptive capacity of H_2S and lower mercaptans are about 1% and 0.5% by weight, respectively. Higher sulfides and disulfides are not removed.

The preferred feedstock is an LPG stream containing saturated hydrocarbons, H_2S and a relatively small amount of mercaptans. Typically, a regenerative unit might be sized for 40 gal/min throughput with a product specification of 1-2 ppm H_2S or mercaptans and 10 ppm maximum outlet water concentration. More than one dozen units are used for sweetening over one billion scf/day natural gas. Molecular sieves are commonly used for gas drying. All cryogenic gas processing plants in the U.S.A. except two use molecular sieve dehydration. In the process, the liquid LPG is percolated through solid adsorbents for water removal. The cheapest adsorbent is often a once-through system using calcium chloride. Other adsorbants in common use on a regenerative basis in refineries at the present time are alumina, silica gel or 4 Å molecular sieves. In all such adsorption processes, the packed towers are protected from slop water carryover by upstream separators.

Adsorbents must be periodically renewed. Several techniques help assure full adsorbent service, one of which is the use of a buffer bed. This buffer, equal to about 5% of the total bed volume, protects the adsorbent in the remainder of the tower from temperature extremes and entrained liquid water in the inlet gas stream. Another technique to assure full service life is to properly break in adsorbents. During the first 12 cycles, only 50% of the designed gas flow should be used. To assure full adsorbent service, however, rapid pressure changes in the adsorbent must be avoided. These changes can cause adsorbent breakage, especially when depressurizing a bed fully saturated with liquid hydrocarbons.

4.3 PRODUCTION FACILITY LISTINGS

Two appendices are attached to provide a summary of LPG production plant statistics. Appendix A is a compilation of information on U.S. refineries involved in production of liquefied refinery gases. Included is information on total processing capacity and charge capacities of liquefied refinery gases (LRG) producing processes such as thermal operations, catalytic cracking, catalytic reforming, catalytic hydrocracking and catalytic hydrotreating. The stream day figures represent the amount that a unit can process when running full capacity for short periods.

This information shows that, of 289 refineries involved in producing LRG, 112 are owned by the 20 largest refinery companies. Their crude oil processing capacity of 13,427,600 barrels per calendar day is approximately 80% of the total capacity for all the refineries listed (Cantrell 1978).

Appendix B is a listing of facilities engaged in extraction of natural gas liquids (NGL) from natural gas. This listing includes the gas capacities and gas throughputs for each plant, the processing methods, and average daily amounts of NGL produced based on the previous 12 months production. Out of a total of 762 U.S. gas processing plants, 333 engage in the production of propane, isobutane, normal or unsplit butane, and LP gas mix (Cantrell 1979).

5.0 LPG TRANSPORTATION

LPG is transported from production facilities either directly to large consumers and consumer distribution facilities or to primary storage and peakshaving facilities. Whenever possible, pipelines are used for transporting LPG over long distances. However, rail tank cars, tank trucks and barges are also employed to reach markets that are not supplied by pipelines. LPG from foreign sources is shipped to the U.S. in ocean-going tankships.

The U.S. demand for LPG peaked in 1976 and is estimated to have fallen off in the last four years. However, a current world surplus and rising prices of other fuels are expected to increase future U.S. demand above 1976 levels (NLPGA 1978). Information published by the National LP-Gas Association indicates the typical modal contributions to LPG transportation (Table 5.1).

TABLE 5.1. LPG Transportation by Mode in 1976(a) (NLPGA 1977)

Transport Mode	Volume (10^6 gallons)	Percent Of Total
Truck	678	3.4
Rail	180	.9
Pipeline-Truck	17,879	90.6
Pipeline-Rail	899	4.6
Tanker or Barge(b)	107	.5
TOTAL	19,743	100.0

(a) The figures reflect movements of LP-gas from points of production or import to distributor or storage facilities.

(b) Approximately 90% of the volume in this category is transported by pipelines prior to tanker or barge movement.

Most of the propane (90.6%) shipped in the United States in 1976 was by pipeline to storage or distribution terminals and then to trucks for final delivery; 4.6% was shipped by pipeline to intermediate terminals and transferred to rail tank cars. Only 3.4% of the total was transported by truck from the point of origin directly to the consumer; direct rail delivery involved 0.9% of the total. Less than 1% traveled by tanker or barge. Most imported propane is moved by rail tank cars, while most local distributions are by truck.

Figure 5.1 shows the principal transportation links used by the LPG industry which are described in this section.

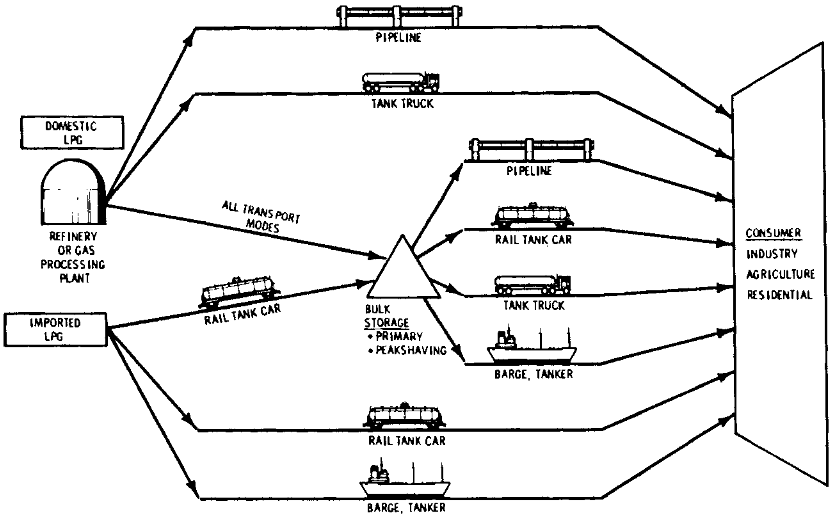

FIGURE 5.1. General Patterns of LPG Transportation and Distribution (Geffen et al. 1980).

5.1 LPG PIPELINES

By the late 1970's, 225,000 miles of pipelines had been constructed and were being used to transport about 19 billion gallons of LPG (NLPGA 1978). As indicated in Table 5.1, this accounts for 95 percent of the LPG moved by all modes of transportation. Figure 5.2 shows the LPG pipeline systems existing in 1978 (Congressional Research Service 1978).

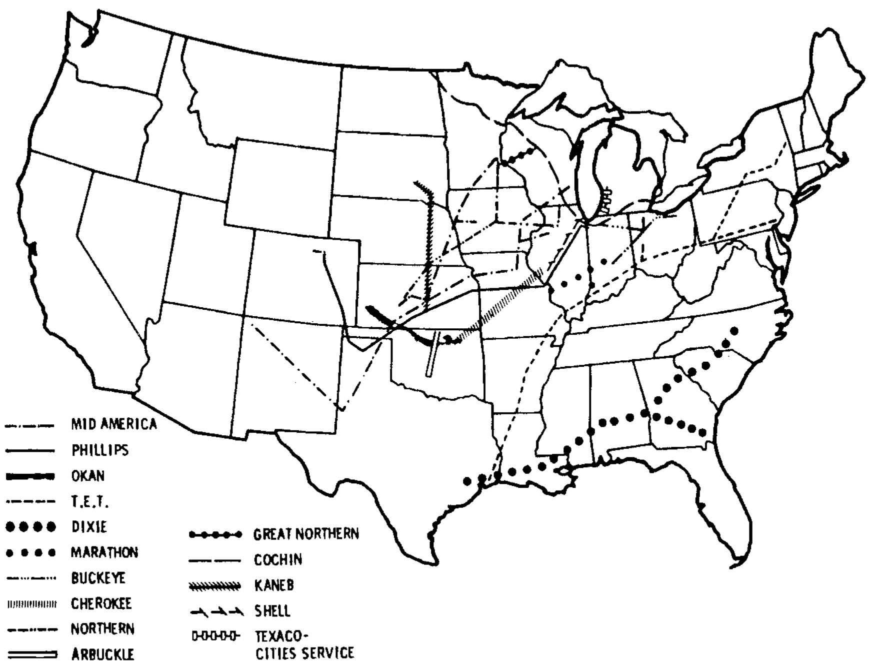

FIGURE 5.2. Map Showing LPG Pipelines (Congressional Research Service 1978)

5.1.1 LPG Pipeline Design

LPG pipelines are designed and constructed so that they will have a high degree of structural integrity. The incentives for this include the safety of the general public, safety of employees, avoiding loss of LPG and damage to company and non-company property, and preventing costly shutdowns. Toward this objective, the pipeline operators, trade and professional organizations, and the federal government have established specifications, standards, and regulations which are intended to produce the desired integrity and serviceability of pipelines. The industry standards and specifications and the requirements established by federal regulations are described in detail in Section 9 of this report.

Overall pipeline design considers LPG flow maintenance and section isolation for operation and maintenance and safety purposes. Pumping stations must be spaced along the pipeline to move the LPG at the desired flow rate and maintain a sufficiently high pressure head to prevent vaporization of the LPG. The spacings between the pumps typically range from less than 10 miles to as much as 150 miles depending on design and economic considerations. For example, a pipeline may be planned for a certain ultimate capacity, but will be built initially with enough pumping capacity to handle only the present need.

5.1.2 Pipeline Construction

Construction of a products pipeline is a large and expensive undertaking. The route is determined on the basis of geographical, environmental and safety requirements, and, of course, the economics. A right-of-way is obtained and cleared to accommodate the construction operations.

The joints of pipe, usually 40 feet long, are laid end to end along the right-of-way. Pipe procured for construction of a pipeline may be used in various ways. The most common practice is to order pipe of nominal wall thickness and grade for a section of pipeline between pump stations by choosing the thickness and grade appropriate to the highest pressure location within that section. An alternative construction technique is to telescope the pipe. In this technique, pipe of decreasing nominal wall thickness or lower grade steel is used in sections farther from a pump station in accordance with the pressure gradient along the pipeline.

A ditch, the depth of which is dependent upon the pipe diameter, is dug. The pipe is butt-welded by highly-trained welders, is bent if necessary to fit the ditch and topography, and the welds are examined by means of X-rays for flaws. The completed pipeline is then wrapped and coated to protect it from corrosion and is lowered into the ditch. Most pipelines also use cathodic protection against corrosion by imposing a small electric potential on the pipeline. Additionally, when crossing rivers, marshes, or other bodies of water, the pipe is encased in concrete to assure that it will not float. After the pipeline has been laid, it is tested hydrostatically for leaks. After successful completion of the hydrostatic test, the trench is filled, compacted, and leveled. In most cases, the prior use of surface lands can be resumed.

5.1.3 Pipeline Operation

LPG moves through pipelines in a continuous flow moved by pumps. Storage facilities are provided at both ends of the line and at any input and off-take points in between to even out variations of supply and demand. LPG is transported either in dedicated pipelines through which no other product is transported or as batches between other oil products in "all products" pipelines.

Batching within the pipeline is often accomplished by placing a batching sphere in the liquid stream when a switch is made from one product to another. A batching sphere is normally an inflatable, water-filled rubber sphere which fits the inside diameter of the pipeline and helps to reduce the interface area between two adjacent products. The intent of using a batching sphere is to reduce the amount of mixing which occurs between products. Despite the use of the spheres, some mixing does occur, and this portion of the pipeline charge must be channeled into an interface tank to be separated later into different products or to be blended with sufficient quantities of pure product so desired specifications can be met.

The liquid pipeline is controlled from a central location (Koch and Lutz, 1969). The type of control of operation of these pipeline systems ranges from manual call-in of parameters which are then usually displayed on a control panel, to complete computer control. In some cases, a main computer is used in conjunction with various types of peripheral equipment such as hard-wired controllers. A large number of pipelines also use a manually operated control panel with various types of peripheral equipment.

and blowdown from seals, piping, and valves. The disposal system may be a lighted flare, or the LPG may be vented to the atmosphere. Final disposal may be delayed by moving the vapor or liquid into a pressurized holding tank. The vapor space in the holding tank may be connected to a flare or vent, or the tank may supply a recovery process that recycles the product back into the pipeline. LPG pipelines use all these types of systems, with the choice being dependent on specific situations and local conditions.

5.2 TRUCK TRANSPORT OF LPG

Transport of LPG by truck alone in 1977 amounted to almost 800 million gallons which was about 4% of the total shipped by all modes. In that same year, however, transport of LPG by the pipeline/truck combined mode amounted to slightly over 18 billion gallons or about 91% of the total. According to NLPGA, there were about 25,000 transport and delivery trucks in operation in the U.S. in 1977 (NLPGA 1978). Highway transportation of LPG is accomplished using large truck and trailer combinations and small delivery transports.

Large truck and trailer transports with carrying capacities ranging from 3,500 gallons up to 10,000 gallons are used primarily for transporting LPG from production facilities to bulk storage facilities. A small amount of LPG is carried directly from production facilities to consumers by large truck and trailer transports. These trucks, which are owned and operated by common carriers and private carriers, are constructed and operated in accordance with U.S. Department of Transportation regulations. It is estimated that some 7,000 of these trucks are in operation in LPG service.

The cargo tanks for these trucks are generally constructed of high-strength steels of the quenched and tempered type. They are also equipped with safety features required by federal regulations and trade codes.

Small delivery transports, commonly called bobtails, are used for local distribution of LPG from consumer storage facilities to consumers. The tanks on these trucks are usually constructed of conventional mild steels and generally range in capacity from 800 to 3500 gallons. However, tank design and construction must conform to U.S. Department of Transportation and trade codes. There are some 18,000 trucks of this type in operation.

5.2.1 Tank Truck Design and Construction

Federal regulations and industry standards specify requirements for tank truck design, construction and operation. These specifications are discussed in detail in Chapter 9. For instance, federal regulations establish the requirement that LPG be transported in trucks whose tanks are designed and constructed to conform to MC-331 specifications. The regulations require that all tank trucks constructed since September 1, 1965, conform to the MC-331 specification and that all MC-330 tanks be altered to include the MC-331 type release prevention valve system. Because of the length of time these requirements have been in existence, it can be assumed that the vast majority of tank trucks transporting LPG conform to the MC-331 specifications. A relatively small number of trucks used exclusively in intrastate LPG service do not conform to this standard.

MC-331 tanks are constructed in accordance with applicable ASME codes (ASME 1974). Tanks must be seamless or welded steel construction and are uninsulated. Typical tank dimensions and material properties for MC-331 type tanks (both larger cargo tanks and bobtail tanks) are listed in Table 5.2 (Geffen et al. 1980). Figure 5.3 is a photograph of a typical tanker unit used for transporting LPG on the highways. A photograph of a bobtail truck transport is shown in Figure 5.4.

TABLE 5.2. Typical Dimensions of MC-331 Tanks

	Large Cargo Tanks	Bobtail Tanks
Container Diameter	7.15 ft	6.0 ft
Container Length	33.80 ft	15.42 ft
Container Wall Thickness	0.40 in.	0.40 in.
Water Weight Capacity	11,600 gal	2,800 gal
Maximum Permitted Filling Density	45% Water Weight Capacity	45% Water Weight Capacity
Tank Working Pressure	250 psig	250 psig
Tank Test Pressure	500 psig	500 psig

The capacity of MC-331 tank trucks can vary. Any tank with a water weight capacity of 3500 gallons or larger is required by 49 CFR 178.337 to have a manway which is usually located in the rear head of the tank. Typically, the manway opening is about 15 in. in diameter and is reinforced with a 20-in. diameter, 1.5-in. thick steel ring welded to the tank material. Tank trucks have two other outlets at the top of the tank. These are safety relief valves which are spring-loaded and are required to be set to discharge within a particular pressure range. For LPG tank trucks, the start-to-discharge setting of these valves typically is about 270 psig. The valve itself fits inside the tank and is in direct communication with the vapor space. Each valve is required to have a protective cap that prevents the entrance of dirt or water into the valve but does not impede the flow of LPG during release events. Other outlets communicating directly with the tank contents are the openings that lead to loading and unloading lines. The liquid lines are generally 3 in. in diameter and the vapor lines are usually 2 in. In the MC-331 tank, it is required that each liquid or vapor discharge opening larger than 1 1/4 in. in diameter be equipped with a remotely-controlled internal shut-off valve.

The critical shut-off parts of the valve are located within the tank so that the shutoff assembly will remain intact even if the internal valve is sheared off in an accident below the flange or the container pipe coupling. The valve assembly system has a section exterior to the tank that is designed to break under undue strain. This helps insure that the flow rate through the broken connection will be great enough to cause the excess flow valve to operate.

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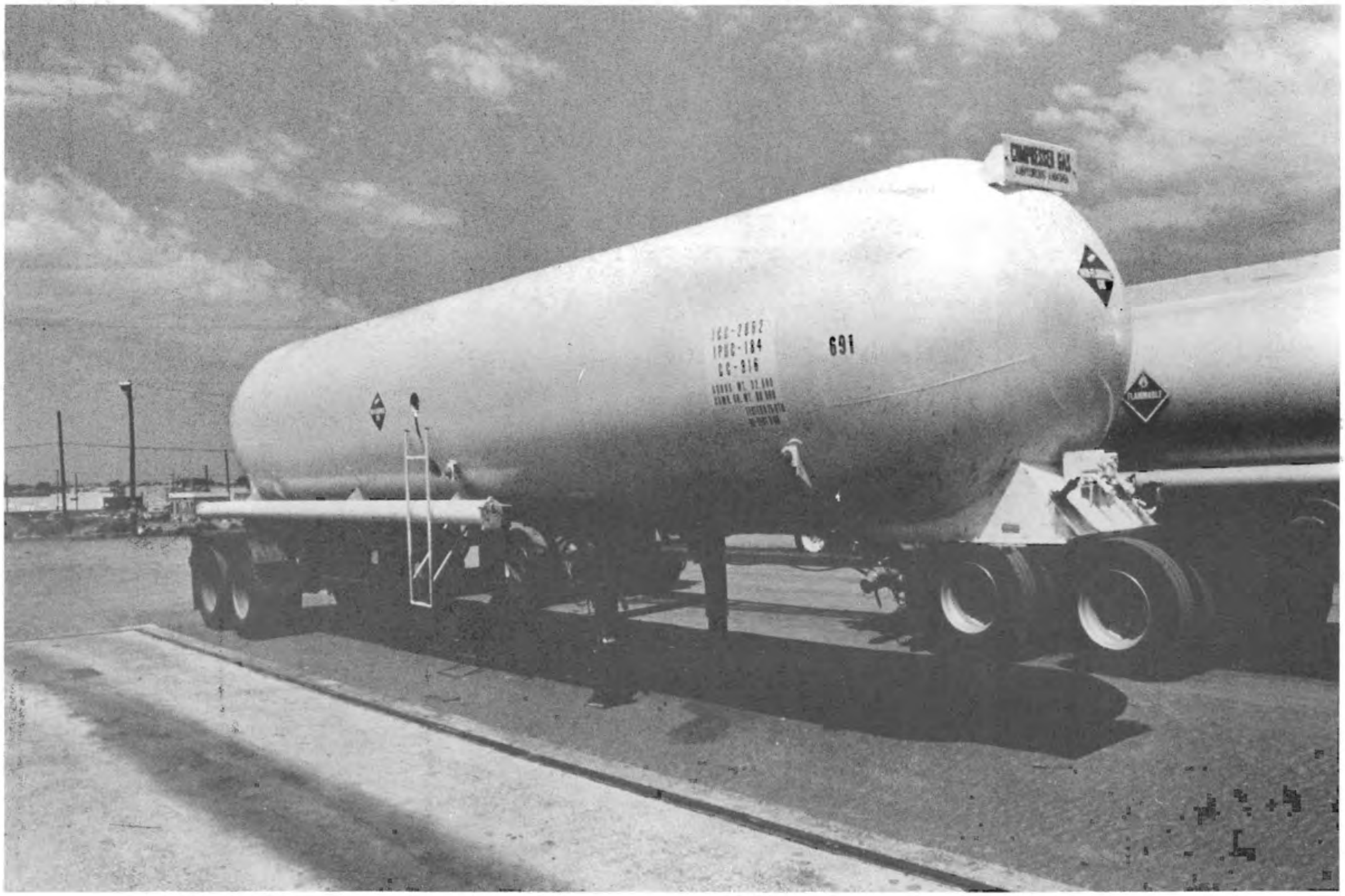

FIGURE 5.3. Typical MC-331 Tank Trailer

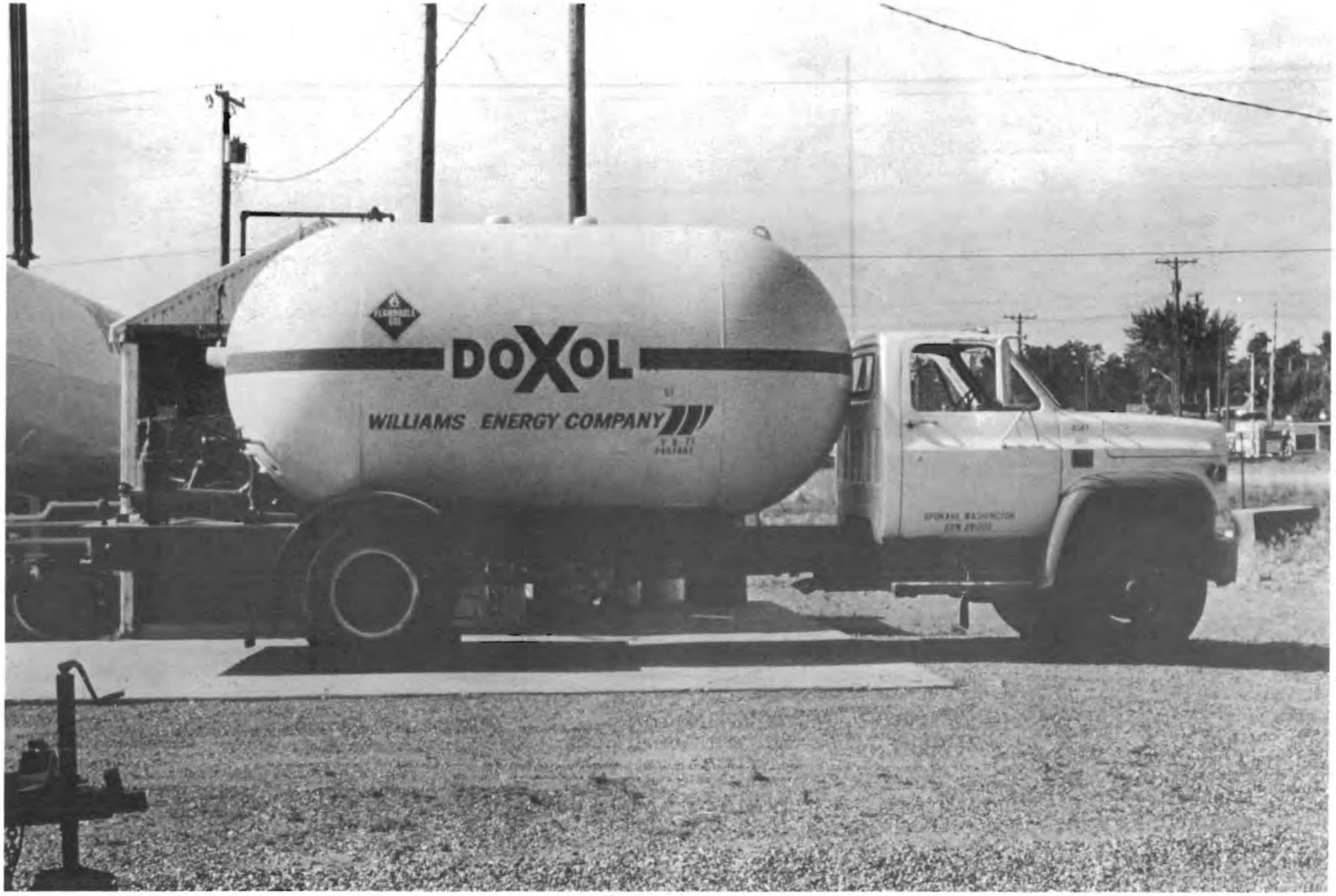

FIGURE 5.4. Bobtail Transport Truck

MC-331 cargo tanks are tested for integrity at least once every five years. The tank is given a hydrostatic pressure test to one and a half times the tank design pressure. The tank is also inspected for corrosion, bad dents, and other weaknesses.

There are three major differences between the bobtail tank truck and the larger tank truck. First of all, the bobtail has no manhole. Secondly, it is much smaller than the large tank truck and is used primarily for local propane deliveries. Finally, most of the piping and valves on the bobtail transport are located at the rear end of the truck, rather than underneath, as on the large tank truck, although the pumping system for transfer operations is located underneath the bobtail tank. The basic valves included are the same. These items are all specified in the federal regulations and are discussed in detail in Section 9.

5.3 RAIL TRANSPORT OF LPG

Shipments of almost 1.1 billion gallons or 5.5% of the total LPG transported in 1976 were either by rail alone or a combination of pipeline and rail (NLPGA 1978). As of 1977, the NLPGA reported 22,000 railroad tank cars in use for transporting LPG.

LPG tank cars are different from railroad cars used for other petroleum products in that they operate under pressure to maintain the gas in a liquid state. LPG rail tank cars built prior to 1961 typically had a water weight capacity of 11,600 gallons. After 1961 much larger tank cars became available with water capacities up to 34,000 gallons.

Liquefied petroleum gases are almost always transported in DOT-105, DOT-112 or DOT-114 specification tank cars. The DOT-114 specification car is designed for multipurpose service and is used to transport materials other than LPG including anhydrous ammonia. The majority of DOT-105 tank cars are small and comparable in capacity to the tank truck. In recent years the majority of LPG transported by rail has been carried in DOT-112 type tank cars (RPI-AAR 1972). By December 31, 1980, all existing and subsequently constructed 112 and 114 specification tank cars used to transport flammable LPG were required to have thermal and tank head protection, and must also be equipped with special (shelf) couplers that are designed to resist vertical disengagement (e.g., during a derailment).

The tank car basically consists of a carbon steel shell with a manway structure. Unlike tank trucks, all the tank outlets on a rail car, including the safety relief valves, are gathered in one location at the top of the car. An illustration of a typical 112 specification tank car is shown in Figure 5.5. Typical tank dimensions and material properties are listed in Table 5.3.

TABLE 5.3. Typical DOT-112 Rail Tank Car Specifications

Container Diameter	10.0 ft
Container Length	60.0 ft
Container Wall Thickness	0.75 in.
Water Weight Capacity	33,500 gal
Maximum Permitted Filling Density (insulated cars measured as % of water weight capacity)	
April-October	46.75%
November-March	48.51%
Tank Test Pressure	335 psig
Tank Rupture or Burst Pressure	855 psig

Source: RPI-AAR 1972, 49 CFR 173.314 and Geffen et al. 1980.

5.4 SHIPS AND BARGES

Transportation of LPG by water is accomplished using both ships and barges. Ships are used in the LPG industry as the principal means of importing or exporting LPG. Barges use navigable inland and intracoastal waterways for transporting LPG to storage facilities or directly to consumers.

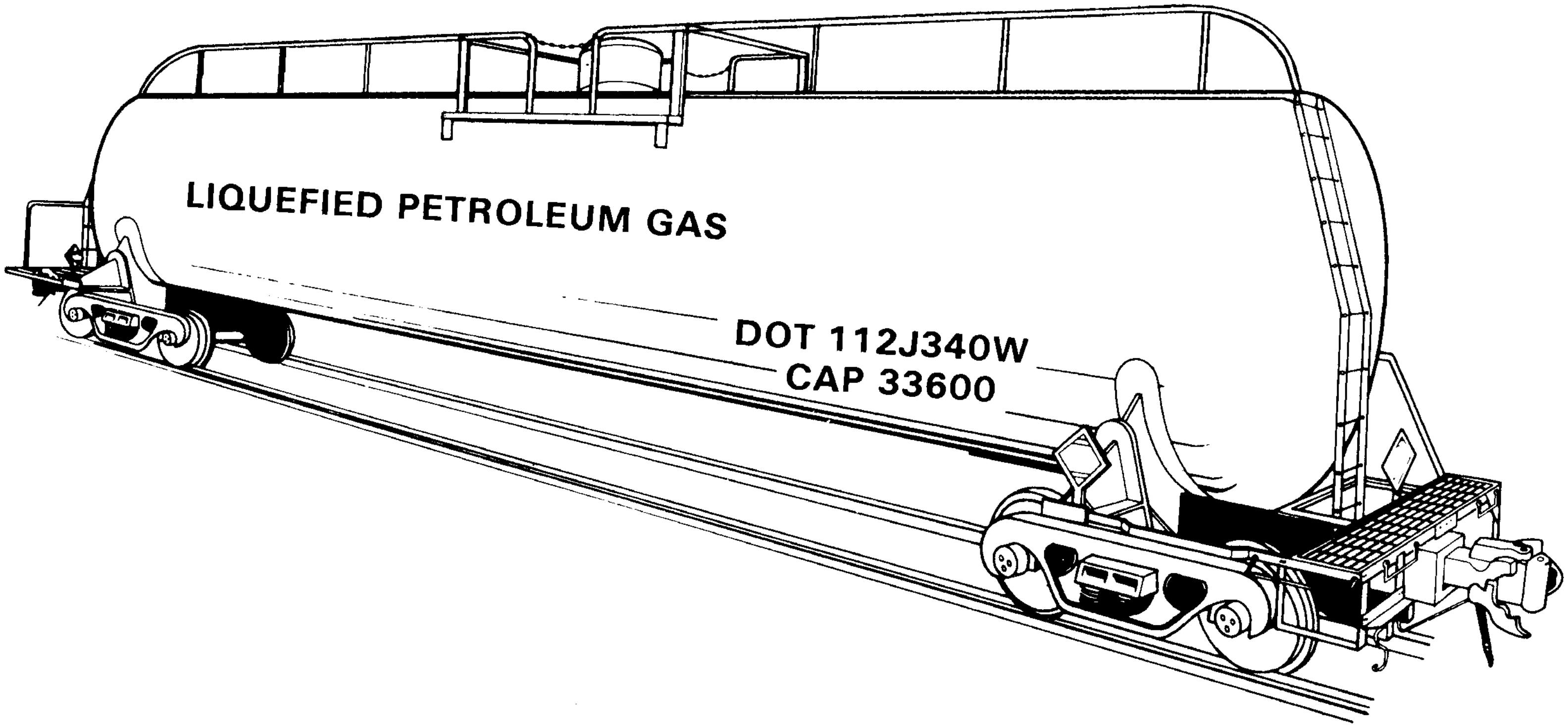

FIGURE 5.5. Propane Rail Tank Car

The 1979 Clarkson's Liquid Gas Carrier Register listed 554 gas carriers in operation with a total cargo capacity of slightly more than 9,000,000 m^3 (H. Clarkson & Co. 1979). An additional 55 carriers with a total cargo capacity of 4,632,300 m^3 were on order as of that date with deliveries scheduled through 1984. A majority of these liquid gas carriers are involved in trades which would not bring them to U.S. terminals.

In 1976, the combined volume of LPG transported within the United States by the ship and barge modes amounted to about 107 million gallons, which was less than 1% of all LPG transported in that year (NLPGA 1978).

5.4.1 Ships

Designers of liquefied gas tankships are aware of the hazards posed by accidental releases of liquefied flammable gases and have, therefore, concentrated on developing systems designed to prevent any cargo releases during normal operating conditions and to minimize the amount released should part of the system fail. This has led to development of better tank designs that have built-in redundancy; tanks that will leak before they fail; or tanks that, through conservative design, can be shown not to fail in ordinary service.

Liquefied gas ships built since 1976 must basically conform to the Inter-Governmental Maritime Consultive Organization Code for the Construction and Equipment of Ships Carrying Liquefied Gases in Bulk (IMCO Code). Prior to that time, all LPG tankers used in U.S. waters were to conform to applicable portions of 46 CFR and/or be approved by the U.S. Coast Guard through their Letter of Compliance (LOC) program. These codes are discussed in detail in Section 9, particularly as they relate to release prevention and control of the LPG cargo. Certain design features are specified in detail in these codes, while others are given only in general terms, thereby leaving the designer some leeway in determining how the specification should be met. In general, an LPG ship is designed to minimize the probability of a cargo tank being damaged by low-speed collisions or groundings; to prevent the spread of damage due either to cryogenic effects or fire; to detect any leaks or spills at an early stage; and to protect the crew from the hazards of the cargo.

The evolution of LPG ship design has been directed by technological advances and economic considerations. Initially the cargo containers on LPG ships were thick-walled pressure vessels, and the practical limit for size was 5,000 m^3. The development of combination ships (i.e., ships in which the cargo is contained under combined conditions such as semipressurized/semirefrigerated or semipressurized/fully refrigerated) allowed for enlargement of cargo capacities up to 15,000 m^3; whereas, the latest development, the fully refrigerated LPG ship, can be built to carry cargos up to 125,000 m^3 in volume.

As mentioned above, ships with pressurized cargo containment systems represent the first generation of LPG transport ships. Their tanks are cylindrical and normally designed for a working pressure of about 250 psi, which corresponds to the vapor pressure of propane at 113°F (which is the highest ambient temperature in which the ship is likely to operate). The primary advantages of this type tank are that they are built of ordinary grades of steel, no liquefaction equipment is required, and operations are simple. However, there are disadvantages including the tank shape which disallows optimization of underdeck space utilization and the high cost and weight associated with the tank wall thicknesses required to accommodate the high design pressure requirements. A lower ratio of cargo carried to weight of tank when compared to refrigerated carriers makes pressurized ships less attractive from an economic perspective.

The development of semipressurized ships with refrigeration plants started in 1959. Through the 1960s, the design of these ships developed very rapidly as the market required more flexibility. This new generation of ships was designed with the capability to heat or cool the cargo during loading operations, or while at sea and also to be able to raise the temperature of the cargo when discharging.

The cargo capacity of semipressurized/semirefrigerated ships ranges normally up to 15,000 m^3. The tanks are usually cylindrical in shape, and the propane is carried at a pressure of about 120 psi and at a temperature of 14°F.

Semipressurized/fully refrigerated ships generally range up to a maximum size of about 12,000 m^3. They are designed to carry the full range of LPG cargos in cylindrical or spherical tanks and designed for a maximum service temperature of 118°F and working pressure in the order of 70 to 115 psi.

The advantages of semipressurized/refrigerated ships over fully pressurized ships are:

- More cargo can be carried in a tank of the same volumetric capacity.
- A tank of the same volumetric capacity is lighter and cheaper to build.
- Much larger and, therefore, more economical ships can be constructed.

The pioneer design of several companies showed the economy of transporting even larger quantities of LPG in fully refrigerated, nonpressurized condition. The self-supporting, prismatically-shaped cargo tanks allow better utilization of the available hold space than the semipressurized/refrigerated ships. The tanks are usually designed for a maximum working pressure of about 4 psi and a minimum working temperature of -58°F. Cargo capacity of this type LPG ship ranges up to about 125,000 m^3.

Cargo containment systems using integral tanks, membrane tanks, semimembrane tanks, and independent tanks are used on LPG ships. Secondary barrier requirements for each cargo containment system vary with the type of containment system used.

5.4.2 LPG Fleet Description

Figure 5.6 shows the development of the world liquefied gas carrier fleet by type and number of ships since 1950 (Rasch 1978). By 1962 the world fleet included 52 ships of the pressure vessel type. In the subsequent three 5-year periods, the fleet of pressure vessel type ships increased by 70, 62 and 45 units. No ships of this design are planned for construction at this time. The fleet of combination ships, which included only 21 ships in 1962, increased by 39, 49 and 35 units in the

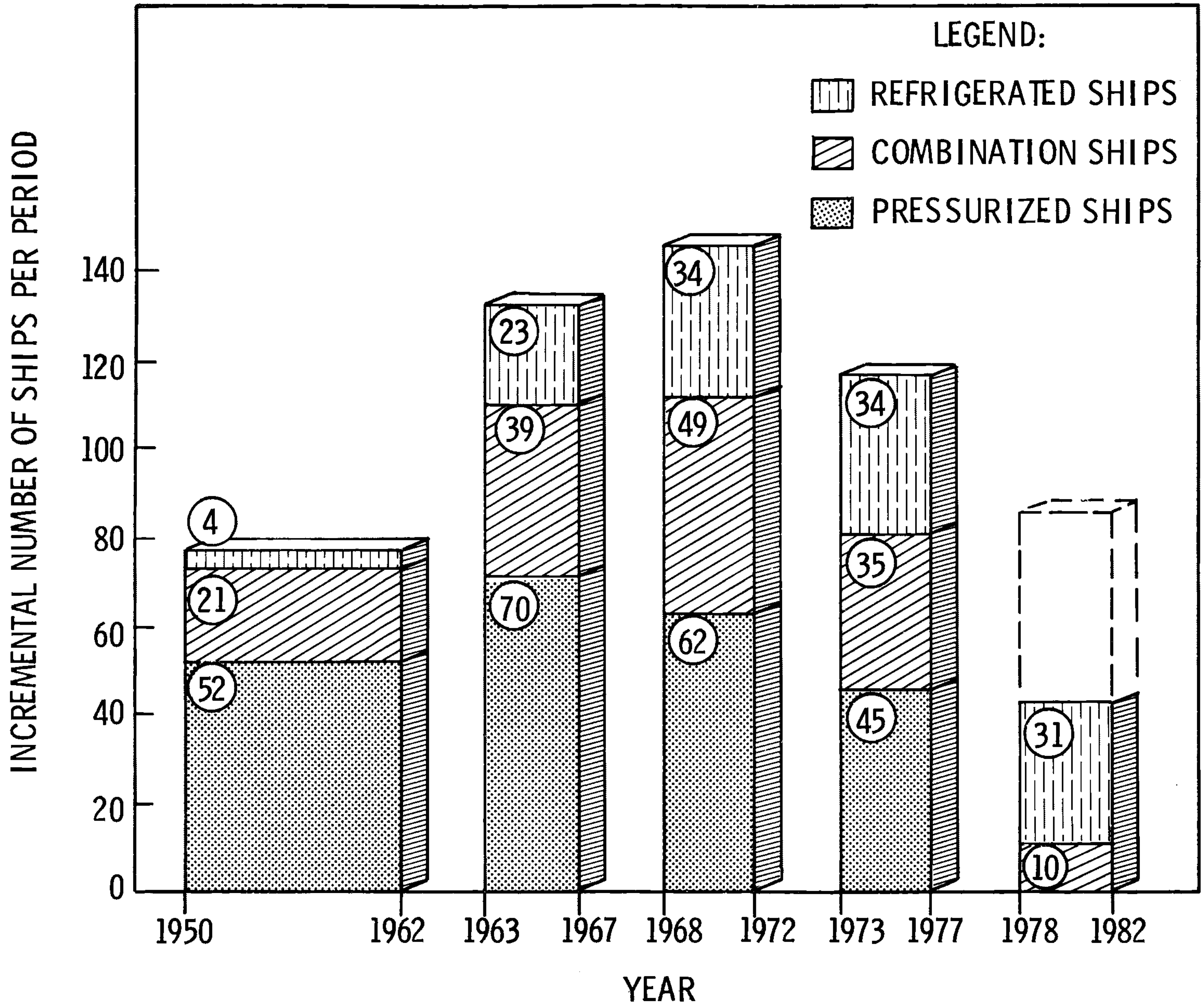

FIGURE 5.6. Development of World Liquefied Gas Carrier Fleet by Type and Number of Ships. (LNG Ships Not Included)

subsequent three 5-year periods. Ten combination ships were on order for delivery during the four-year period from 1978 through 1982. The fully refrigerated fleet included only four ships in 1962, and two of those were delivered to their owners during that year. Deliveries within this category of ships took place at a rate of 23, 34 and 34 ships in the subsequent three 5-year periods. In 1978, 31 ships of the fully refrigerated design type were on order for delivery in 1982.

Figure 5.7 shows the growth trend of the fully refrigerated LPG fleet in terms of consolidated cargo capacity. Only about 60,000 m^3 of capacity were available in this type ship in 1962. During the decade from 1967 through 1977, this fleet grew from about half a million cubic meters of capacity to over 3.7 million cubic meters. Subsequent orders for additional fully refrigerated ships leads to a projection for growth of total fleet capacity to almost 6 million cubic meters by the end of 1982.

Table 5.4 is a listing of vessels over 500 m^3 capacity that delivered LPG cargos to U.S. ports in 1977.

TABLE 5.4. LPG Ships Making Deliveries to U.S. Ports in 1977

Ship Name	Capacity (Cubic Meters)
Antilla Bay	53,424
Antilla Cape	29,540
Capo Ovest	14,800
Cavendish	40,213
Devonshire	52,650
Faraday	31,215
Fernwave	12,000
Gambada	29,791
Garinda	53,000
Garmula	52,647
Havis	15,285
Hoegh Multina	52,000
Inge Maersk	12,060
Lincolnshire	31,290
Luigi Casale	14,268
Malmros Multina	53,400
Marian P. Billips	2,033
Monge	70,700
Mundogas Atlantic	8,565

TABLE 5.4. Continued

Norfolk Multina	25,102
Pascal	15,022
Providence Multina	53,400
Pythagore	14,258
Reliance Gas	26,504
Sine Maersk	12.060
Trina Multina	18,422
Tropigas Far East	5,000
Wiltshire	15,495

These twenty-eight vessels represented approximately 25 percent of all vessels with cargo capacities greater than 500 m^3 that could carry LPG.

The U.S. Coast Guard maintains and periodically updates a listing of status of foreign flag vessels with respect to the Letter of Compliance (LOC) program. This list indicates the dates of first contact with a vessel owner for the vessel, when the vessel plans were satisfactorily reviewed, the vessel was satisfactorily examined for an LOC, and the LOC was issued. Also listed for each vessel is the date the LOC expires, and, if applicable, the date the LOC was invalidated.

The 1979 Coast Guard list is summarized in Appendix C. The list includes those vessels with propane and butane listed as cargos. Under the column headed "Remarks," the expiration date of the LOC is given for 57 vessels for which the date is 1979 through 1981. The remaining 117 foreign flag vessels may have expired and unrenewed LOCs or may have been at any of the various stages in obtaining an LOC. It should be noted that an LOC is only valid for a period of up to two years from its date of issue. Of the 57 foreign flag vessels, listed in Appendix C, 31 have refrigerated cargo tanks, 24 have semirefrigerated cargo tanks, and 2 have cargo tanks of the pressurized type. The two pressure-type vessels are of small capacity; 227 m^3 and 2033 m^3. The 57 vessels have an average cargo capacity of 20,904 m^3. The cargo is contained in four tanks for the average vessel corresponding to approximately 5,075 m^3 per cargo tank.

5.4.3 Barges

Barge mounted pressure cylinders for transport of LPG have been used since the mid 1940s . The first propane barge of large capacity for use on inland waters was a hopper-type barge in which six pressure-type

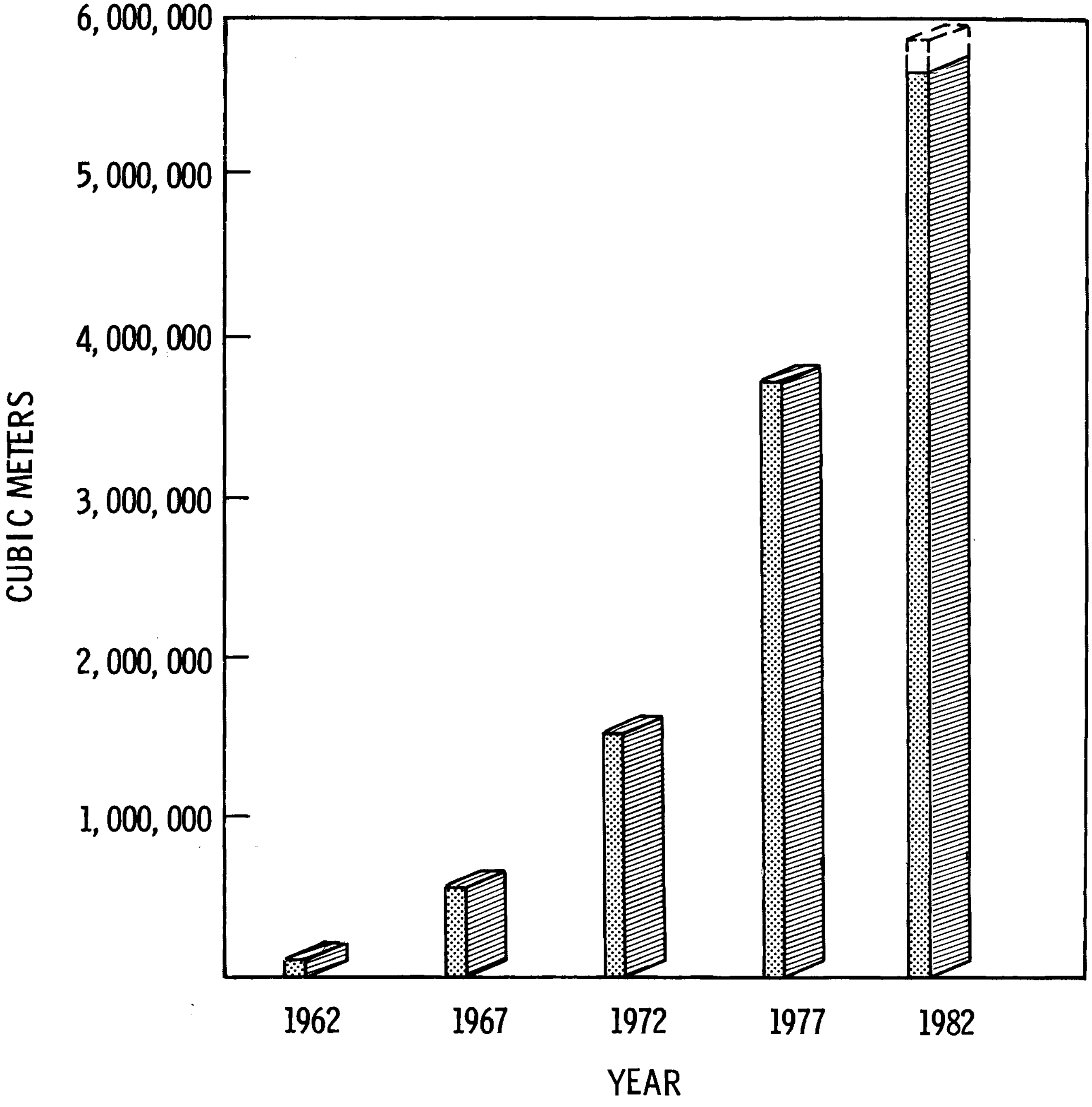

FIGURE 5.7. Consolidated Capacity of the Fully Refrigerated LPG Carrier Fleet

cylinders of 60,000 gallons capacity were installed. The total water equivalent of 360,000 gallons was approximately a net cargo capacity of 320,000 gallons. The barge was designed to be operated on both the intracoastal canal and inland waterways systems.

Capacities of individual barges in operation today vary from approximately 1400 gallons to 4 million gallons. All of these barges are believed to utilize pressure type cylindrical cargo tanks of the independent type.

All barges carrying liquefied flammable gas cargos must be inspected and certified by the United States Coast Guard. A list of barges which have been inspected is maintained by the Coast Guard and updated semiannually. The January 1979 listing of barges certified to carry LPG is presented in Appendix D (U.S. Coast Guard 1979).

The study "Fire Safety of LPG in Marine Transportation" performed by the Applied Technology Corporation as part of the DOE Program includes a detailed description of LPG ships and barges and related operational procedures (Martinsen et al. 1980).

6.0 STORAGE OF LPG

The bulk storage of LPG can be broken down into four major categories: 1) primary storage, 2) peakshaving storage, 3) distributor storage, and 4) consumer storage. Primary storage is used to store LPG as it is produced or imported prior to transportation. Peakshaving storage accommodates fluctuations in consumer demand while maintaining a constant nominal supply of LPG. Distributor storage principally involves LPG stored in tanks owned by retail distributors that supply individual customers. Consumer storage covers a range of quantities from a few pounds to more than 2000 gallons. Much of this category is "bottled gas" storage and includes the LPG cylinders and containers used domestically and in other applications where portability is required. The basic features of LPG storage systems are described in this section together with brief comments on LPG transfer operations.

6.1 PRIMARY STORAGE

Primary storage facilities provide storage for the transportation, distribution, and utilization segments of the LPG industry. Most of these facilities provide storage as a commercial service in which the owners act as brokers. Primary storage facilities are generally not located at refineries or gas processing plants although they may be adjacent to such facilities.

6.1.1 Underground Storage

While some primary storage is accommodated in aboveground tanks, the major primary storage facilities utilize underground storage caverns mined in rock formations and salt domes. The quality of the stored LP-Gas is periodically checked to guard against contamination. These underground facilities are usually owned by private companies which rent storage capacity to utilities.

The standards for designing, constructing and operating these facilities have evolved with the LPG industry and are based on refinery practice and trade and professional standards (e.g., ASME, NFPA, API). The capacities of underground storage sites used primarily as merchant storage are summarized in Appendix E from data published by the Gas Processors Association (GPA 1979a).

6.1.2 Import/Export Terminals

Up to the present time, LPG has been brought into U.S. ports in relatively limited amounts. This situation is expected to change because U.S. domestic production is not expected to keep pace with industry demands. Therefore, it is anticipated that changes will occur in the storage mode and capacity characteristics of LPG terminals.

Most of the smaller capacity terminals in the U.S. utilize refrigerated storage. A notable exception is the Tropigas Inc. of Florida terminal at Port Everglades, Florida, which has pressurized storage in steel tanks. Other terminals, such as the Sun Gas Company terminal at Marcus Hook, Pennsylvania, utilize underground storage caverns mined in granite or other rock.

Terminals at Gulf coast ports may be connected directly to underground storage in salt formations. These formations provide enough storage to take the full cargo from the largest LPG carriers.

If the operational mode of LPG import terminals is to change with increased imports, it is probable that larger size gas carriers will be employed in LPG trades. These larger ships can transport cargo at a lower cost per unit over any distance. The longer the voyage, the more important this advantage. The larger ships, of course, require larger storage capacities which, in turn, favor the lowest unit cost terminal storage. The lowest cost terminal storage is underground storage in salt formations.

The large liquid gas carriers in the 75,000 to 125,000 cubic meter capacities are designed to discharge their cargos in a period of 18 to 24 hours. This translates to pumping rates of approximately 4,000 to 7,000 cubic meters per hour (18,000 to 31,000 gallons per minute). Such rates require large pipelines to carry the product to the storage containers. Existing pipelines and storage wells at the Gulf coast are not designed for these high flow rates. Large gas ships have discharged at Corpus Christi and Houston on the Gulf coast but have required proportionately increased discharge at times.

A compilation of existing terminals has been prepared by the Gas Processors Association (GPA 1979b). The data were obtained from survey responses and other sources believed by GPA to be reliable. Table 6.1 summarizes the characteristics of existing terminal facilities, and Table 6.2 lists proposed LPG import terminals. Detailed information on existing import/export terminals is provided in Appendix F.

6.2 PEAKSHAVING FACILITIES

Sendout requirements of gas distribution companies are characterized by hourly, daily, and seasonal variations. In order to achieve minimum costs for the transmission of natural gas supplies from distant producing areas, the pipelines must be operated at or near capacity every day of the year. To reconcile these opposing requirements, a distribution utility must supplement pipeline purchases during periods of peak sendout. This practice is termed peakshaving.

Supplementing natural gas supplies with LPG/air mixtures is carried out in areas where local conditions make it economically favorable. A secure source of supply must be available to meet all or part of the peak load requirements. This type of peakshaving is generally produced by vaporizing LPG and mixing it with air. The type of LPG used for supplemental natural gas is almost always propane. Propane-air plants are used by utilities in all parts of the country. Their simplicity makes them well adapted to operation on short notice and with the minimum of operating personnel.

Peakshaving with propane, rather than with natural gas, gives rise to two operating problems for the distribution system: 1) the gases differ in calorific value (i.e., the quantity of heat produced by burning a given volume of the gas); and 2) the gases have different burning characteristics which may cause unsatisfactory operation of natural gas burners and appliances when using propane. The first problem is overcome by diluting the propane with air in order to reduce the available heat energy per unit

TABLE 6.1. Summary of Existing Import/Export Terminals

TERMINAL FACILITIES	DOCK RESTRICTIONS, FEET			PRODUCTS HANDLED						UNLOADING CAPACITIES, MBH		STORAGE CAPACITY, MB		
Company--Location	Length	Draught	Breadth	C_3	iC_4	nC_4	C_4Mix	BP	Other	C_3	C_4	C_3	C_4	LPG Mix
Atlantic Energy, Inc. --Norfolk, VA	720	35(B)	110	X	X	X	X			8-12[1]	8-12[1]	--	--	--
California Liq. Gas Corp. --Ferndale, WA	1,000	37(S)	--	X	X	X	X	X		--	10	350	350	350
Cities Service Co.[5] --Lake Charles, LA	750	38(F)	125	X	X	X	X	X		10	10	2,000	2,000	2,000
Coastal St. Crude Gathering --Curpus Christi, TX	--	40(B)	--	X	X	X	X	X		4.5	4.5	6,600	1,060	1,600
Exxon Co., U.S.A. --Everett, MA	800-1,000	38(S)	120	X						10	--	400	--	--
Gulf Oil Corp. --New Orleans, LA	650	38(F)	--		X	X	X			--	4	72	170	--
Gulf Oil Corp. --Philadelphia, PA	735	32(B)	110			X	X			--	1.5	--	200	--
Petrolane Inc. --Los Angeles, CA	--	34(B)	--	X			X			10	10	300-600	300	--
Petrolane Inc. --Providence, RI	700	34(B)	--	X						7.5	--	--	400	--
Petro-Tex Chemical Corp. --Houston, TX	674	36(F)	74	X	X	X	X	X	X	2	2	no limit[2]	no limit[2]	no limit[2]
Phillips Petroleum Co. --Houston, TX	750	36(F)	110	X	X	X	X	X		5	1.7	--	no limit[3]	--
Promix[4] --Plaquemine, LA		Incomplete		X	X	X	X	X		--	--	--	--	--
Dorchester Sea-3 --Portsmouth, NH	640	32(B)	100	X						12.5	--	400	--	--
Sun Gas Co. --Marcus, PA	1,000	38(F)	--	X	X	X	X			18	9	1,170	400	--
Tropigas Inc. of Florida --Port Everglades, FL	378	25(S)	65	X						83	--	10.5	--	--
Warren Petroleum Co. --Houston, TX	750	39(F)	116	X	X	X	X	X	X[7]	15[6]	15	no limit[3]		--
Warren Petroleum Co. --Port Everglades, FL	700	39(S)	--	X						1.58	--	38	--	--

Footnotes:

[1] Fully Refrigerated

[2] Commercial Storage at Pierce Junction

[3] Commercial Storage at Mont Belvieu

[4] Privately Owned--Not a Merchant Terminal

[5] In Operation Last Quarter 1979

[6] Fully Refrigerated; 4 MBH Ambient

[7] Natural Gasoline, Naphtha

(B) Brackish Water

(F) Fresh Water

(S) Salt Water

TABLE 6.2. Proposed LPG Import Terminals

PROPOSED LPG IMPORT TERMINALS	PORT	DAILY THROUGHPUT CAPACITY (Million Barrels Per Day)	STARTUP DATE
Company			
Gulf Coast			
Coastal	Freeport, Tex.	35	?
Tet, Nng Mobil, Texaco	Sabine Pass, Tex.	300	1981
Conoco	Houston, Tex.	150	1982
Oil Tanking	Houston, Tex.	?	?
East Coast			
Gulf Interstate	Moorehead City, N.C.	?	?
Sea-3	Arthur Kill, N.J.	75	?
TOTAL CAPACITY		560	

volume of gas mixture to coincide with that for natural gas; about 1300-1400 Btu/ft^3. The solution to the second problem is to use the propane-air mix to replace only a portion of the natural gas. It has been demonstrated that limited proportions (up to about 50%) of propane-air mixtures with natural gas give suitable performance. This depends to some extent on the type of burners found in the system.

The basic components of all propane-air peakshaving plants are liquid storage facilities, vaporizers, and a proportioning system, as shown schematically in Figure 6.1.

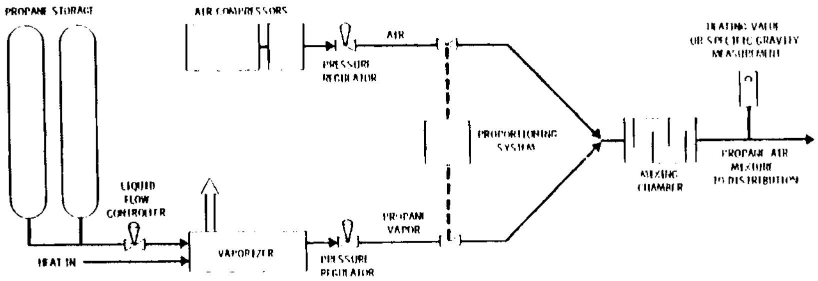

FIGURE 6.1. Propane-Air Plant

These plants vaporize liquid propane and mix it with the desired proportion of air. If the plant is to feed gas into a distribution system at high pressure, air compressors are needed to supply air to the proportioning system at the required pressure. If desired, automatic control of the proportioning operation or remote control of the entire cycle of operation from start-up to shutdown can also be incorporated into the plant.

Peakshaving plants store liquid propane in steel pressure tanks, in refrigerated tanks, and in mined underground cavities. For many years, pressure tanks were used almost exclusively for this purpose, and some storage of this type is installed at nearly all propane-air plants.

LPG vaporizers are simply heat exchangers designed to evaporate liquid propane and add more heat to the vapor. The heat sources used are steam, hot water, or burner flue gas.

Accurate blending and thorough mixing of propane vapor and air is the important final step in the production of propane-air peakshaving gas. A number of packaged units have been developed to achieve it.

In addition to the basic components just discussed, considerable auxiliary equipment is required to create a working peakshaving plant. The components must be piped together, and propane must flow through the plant. Most plants have a liquid pump in the line to the vaporizer. Unloading facilities must also be provided to transfer liquid propane from the delivery vehicle, normally railway cars or trucks, to storage.

A total of 491 plants operated by 212 gas utilities in 39 states were estimated to be in operation in 1978 (Hale 1978). Since then, it is estimated that approximately 120 additional plants have been constructed, representing an increase in sendout capacity of approximately 25%. In the meantime, it is also probable that older facilities dating back to 1946 have undergone significant modifications or may be decommissioned. To obtain a more complete listing of peakshaving facilities, inquiries were sent to 13 suppliers of peakshaving component hardware such as storage tanks, vaporizers, blending equipment, etc. and engineering/contracting firms. Responses were received from 7 of the 13 inquiries. The list thus generated provides a starting point for compiling an up-to-date facility list.

The resulting list of LPG-air peakshaving facilities, contained in Appendix G, represents approximately 210 million gallons of storage capacity. Although this list is incomplete, it summarizes the characteristics of most of the LPG peakshaving plants currently operating in the U.S.

6.3 DISTRIBUTOR AND CONSUMER STORAGE

Before the LP-gas reaches the consumer, it passes through various distributing plants or points located at convenient places throughout the country. Distributing plants are facilities which receive LPG in tank cars or by truck transport and distribute it to the end user by portable container delivery, by tank truck, or through gas piping. Such plants have bulk storage capacity in excess of 2,000-gallons water capacity and usually have container-filling and truck-loading facilities on their premises.

Distributing points are smaller facilities, such as LP-gas service stations, which normally receive gas by tank truck and then fill small containers or the engine fuel tanks of motor vehicles on their premises. Any such facility having an LPG storage capacity of more than 100-gallon water capacity is considered a distributing point (NFPA 1980).

In recent years the LPG service stations have greatly extended their service to dispensing LP-gas as motor fuel, particularly in rural areas. The outward appearance of their equipment appears very similar to that of regular gasoline stations.

Individual users obtain the LPG usually in portable containers, which include "cylinders," "portable tanks," and "cargo tanks." The maximum size for cylinders is a 120-gallon water capacity. Various shapes and sizes for LPG cylinders are available. They are mainly constructed of steel, but aluminum cylinders are also available.

In most cases smaller cylinders are delivered already filled to replace empty ones. Larger tanks, which are mounted permanently, are usually filled from a tank truck. Portable tanks (also called skid tanks) are larger than 120 gallons and are mounted on skids or sunners. Skid tanks can be moved from place to place.

There appear to be no readily accessible statistics on the number of LPG tanks and cylinders in current operation. Table 6.3 provides an indication of the number of tanks in different sizes produced in the period 1968-1977. Production increased during this period from approximately 1 to 1.6 million containers per year.

TABLE 6.3. LP-Gas Cylinder Production, by Size

SIZE IN POUNDS[a]	5	10	20	33	40	60	100	200	300	All Other
1968	38,736	50,789	630,812	148,635	14,761	28,312	257,013	11,033	21,922	1,290
1969	67,516	32,431	481,211	115,439	10,779	8,290	192,456	5,714	16,124	--
1970	124,231	40,823	492,911	132,277	8,777	11,139	194,371	12,399	29,670	35,310
1971	53,202	80,947	819,120	150,346	7,411	18,111	196,743	4,388	27,416	15,123
1972	29,619	31,264	858,836	374,450	52,870	12,888	138,994	9,683	1,699	52,133
1973	36,375	33,997	902,107	340,085	60,188	21,559	144,553	10,373	1,737	33,100
1974	23,862	44,911	624,856	248,271	40,828	24,443	107,184	9,483	2,065	32,247
1975	13,958	36,373	721,714	266,214	52,583	39,733	81,469	8,608	881	73,341
1976	13,264	32,532	953,719	400,971	47,382	33,344	69,543	12,782	53	40,548
1977	27,682	50,880	902,726	381,742	56,826	19,189	61,362	9,742	--	50,249

[a] For clarity and convenience, size classifications have been listed by the common cylinder sizes.

SOURCE: NLPGA (1978)

All LPG containers are equipped, depending on the prevailing code requirements (see Section 9), with up to 8 fittings, valves, or other accessories. These are: safety relief valve, service line valve, filler valve, vapor return valve, liquid level indicator, excess flow checkvalve, pressure gage, and pressure regulator.

For protection against mechanical damage, the fittings are normally contained in a dome on top of, or at the end of, the tank. Also, containers are always identified by nameplates or markings, indicating their individual characteristics. LPG tanks on consumer premises may be installed above or below ground.

Any LP-gas cylinder is only partially filled with liquefied gas which remains in equilibrium with the vapor occupying the space above the liquid. When the valve on the cylinder is opened, some of the pressurized vapor leaves the tank. Some of the gas in the liquid state immediately evaporates to maintain equilibrium, thus always keeping the pressure inside the cylinder constant with constant temperature regardless of the fluid level. The pressure in the LP-gas tank is equal to the vapor pressure of that gas or mixture of gases stored in the tank at that temperature. Thus, since propane and butane have vapor pressures at 70°F of 132 psig and 17 psig, respectively, the pressure in the tank containing these gases will be between these two extremes, depending on the mixture of these two gases. At a temperature of 130°F the same tank would have a pressure somewhere between 69 psig and 300 psig. However, since the safety relief valve on propane tanks is set somewhere between 200 and 250 psig, the temperature of the propane within the tank can never rise above about 115°F.

In most small applications the LP-gas is vaporized by extracting the required heat of vaporization from the environment. For larger installations this method is usually not adequate and special vaporizing equipment must be installed, particularly in cold climates. In most cases, some of the LPG is burned in the vaporizer to provide the required energy for vaporization. Further discussion of transfer methods is included in Subsection 6.4 below.

6.4 TRANSFER OPERATIONS AT STORAGE TERMINALS

At each storage terminal, there exists one or more means of transferring LPG to and from storage. In many cases, transfer is achieved by means of permanent connections to pipelines. However, when truck and rail cars are used, transfer operations are generally performed with temporary connections between the storage tank and the vehicle. These transfer operations are somewhat unique in that the design specifications of the carrier tanks vary, and the connections between storage tank and carrier must be frequently changed.

6.4.1 Transfer Operations with Railroad Tank Cars

Before loading or unloading connections are made to a railroad tank car, the car is spotted at the correct location, the brakes set, the wheels blocked and a derail placed between the spotted car and the carrier's tracks. During the time that the tank car is connected, and before the outlet valve is opened, caution signs must be placed on the track and cars to warn persons approaching the car. These signs must be left in place until the liquid transfer is completed and the car is disconnected. At least one attendant is present during the entire transfer operation. Before starting liquid transfer, the carrier tanks are inspected for exterior damage and possible leaks.

Connections for loading or unloading, safety valves, gaging and temperature devices are housed under a protective steel shell at the top center of the tank car. The cars are loaded in accordance with the Department of Transportation Table of Filling Densities. Before filling a storage container, the gaging device is checked and set or marked to indicate the maximum quantity of liquid to be placed in the container.

Various methods are used to generate the pressure differential necessary for transfering the LPG. The principal methods are:

- With a compressor - compressor withdraws and compresses vapor from storage tank and transfers it to the tank car's vapor space. A pressure differential is thus developed, and liquid transfers from the tank car to the storage tank.

- By gas repressuring - non-condensable (in some cases, inert) gas is used to pressure the tank car and drive the liquid into the storage tank.
- By direct-acting liquid pump - with this method it is not possible to remove vapors from the tank. Therefore, a vapor compressor is almost always used in addition to the liquid pump.
- By vaporizer - liquid from storage is vaporized to exert pressure on the top of the liquid in the tank car.

Before unloading any tank car, it is necessary to make sure that the vapor pressure of the product to be unloaded does not exceed the maximum vapor pressure at 100°F, marked on the receiving container. The invoice or delivery ticket should show the specific gravity and vapor pressure at a temperature of 100°F. When possible, the consignee measures the temperature and pressure of the product and checks them against the shipper's figure.

6.4.2 Transfer Operations with Tank Trucks

A liquid pump is the usual method for transferring LPG from a tank truck. Due to the low vapor pressure of butane at low temperatures, the transfer of butane in cold climates may require both a liquid pump and a vapor compressor. If a compressor is used, it draws vapor from the discharge vessel, compresses and discharges the vapor back into the delivery tank, thus creating a pressure differential and forcing the liquid to flow from the delivery to the receiving tank.

When undertaking transfer operations, propane must not be loaded into a vessel designed for butane, although the converse is allowable. Overfilling must be guarded against by constantly monitoring the liquid level. Road trucks must be electrically grounded to prevent the build-up of static electricity. In addition, transfer lines must have electrical continuity across any connectors.

7.0 CONSUMER USE OF LPG

LPG is used as a source of energy, mostly where natural gas is not readily available and as a feedstock in the petrochemical industry. The LPG industry serves about 18 million customers, including business and government as well as homes, farms, and other users. It is estimated that roughly 60 million people are dependent on LPG for one use or another.

7.1 HISTORIC CONSUMPTION AND GROWTH TRENDS

During the 15-year time span from 1936 to 1951, LPG use in the United States increased about 100 times to nearly 2×10^9 gal/yr. In the following 15 years (1951-1965), it increased about 8 times to nearly 1.5×10^{10} gal/yr, but in the most recent 15 years (1965-1980), it increased at a much slower rate(a) to about 2.3×10^{10} gal/yr. Cited figures vary, depending on the source of information. Government figures are usually lower than industry figures. Presently (1981) the growth amounts to about 2% per year.

The U.S. has been essentially self-sufficient in LPG in the past and has even been in a position to export small amounts. However, of all the LPG consumed in this country, about 10% is now imported. This amount is expected to increase in the coming years as the domestic supplies of crude and gas are further diminished. By 1985 LPG imports will probably double. Compared to other major energy sources (oil, coal and nuclear), the contribution made by LPG to the total energy consumption in the U.S. is still relatively small (at about 2%). This is expected to rise slowly to at least 3% by 1985.

(a) There was actually a temporary decrease in LPG use during 1978. On the average, the LPG consumption in the U.S. has been about constant from 1977-1980. With the 1978 drop in LPG consumption, prices fell from 31 cents/gal in 1977 to a low of 21 cents/gal in 1978, but by 1980, they had risen to 42 cents/gal.

The chemical feedstock market for LPG is essentially dominated by economic factors (i.e., price) since it can be substituted with relative ease by other petrochemicals such as naphtha. Therefore, this use is highly susceptible to the cyclic ups and downs of the petrochemical industry and to prevailing government regulations. In contrast, the LPG fuel market is much more stable. In this area, LPG is used in residential/commercial applications (for heating, cooking, etc.), in utility applications (for peakshaving), in transportation (e.g., as a substitute for gasoline), in industrial applications (e.g., for fuel-switching to cheaper and environmentally more acceptable fuels), and in agricultural applications (for use in farm vehicles, for grain drying, etc.). The main incentive to use LPG has been its relatively low cost (on a Btu/dollar basis) compared to other fuels. LPG is used either as propane or butane alone or as a mixture of the two. Propane use ranks first with about 80% of the market, compared to butane with 16% and propane/butane mixture with about 4%.

For the future, it is expected that worldwide LPG production will grow and that a surplus of it will persist at least through the 1980s (Oil and Gas Journal 1980). This surplus is expected because the processing capacity for LPG will grow rapidly in the coming years, as less and less gas will be flared(a) at the well head but, rather, will be processed for sale as LNG and LPG. Projects for this purpose were started by OPEC nations several years ago and are coming on stream now, and it would not be economical to cancel them or to let them sit idle. Therefore, it is expected that in the future all exporting countries will insist that their customers buy surplus LPG along with the crude.

Another reason for a larger production of LPG is that refineries are expected to increase cracking severity to meet the octane requirements for the more fuel-efficient cars. This will, inevitably, produce more LPG as a by-product.

(a) Around 1980 about 4 times as much gas was still flared or vented as was liquefied and traded.

The largest commercial user of LPG is the chemical industry, which uses LPG as a feedstock for producing a variety of synthetic materials. Another sizable commercial application of LPG is its use for peakshaving in natural gas supply systems. The number of these large industrial and utility users of LPG, however, is quite small compared to the number of small individual users. The balance of this section describes LPG uses in the retail market.

7.2 RETAIL CUSTOMER USE OF LPG

In 1977, the NLPGA reported the number of LPG retail customers as follows (NLPGA 1978):

Homes	10,772,600
Farms	1,394,000
Commerce and Industry	944,400
Recreation Vehicles	5,502,700

LPG sales were also reported with the following breakdown:

	10^9 Gallons
Residential and Commercial	7.17
Internal Combustion	1.14
Industrial	1.04
Utility Gas	0.57
Chemical and Synthetic Rubber	3.58
Other Uses (including agriculture, SNG feedstock, and secondary recovery of petroleum)	2.01

The major consumer applications for LPG are listed in Table 7.1. Descriptions of these applications follow.

7.2.1 Residential Applications

LPG use in the home is primarily concentrated in rural areas where piped natural gas is not available. It is also used frequently in isolated cabins, mobile homes and campers. In these applications, its use extends mainly to cooking, water heating, refrigeration, and space heating.

TABLE 7.1. Consumer Applications of LP Gas

Residential & Commercial	• Kitchen Ranges
	• Barbecue Grills
	• Water Heaters
	• Clothes Dryers
	• Central Heating Systems
	• Central Air Conditioning Systems
	• Space Heaters
	• Patio and Yard Lamps
	• Patio and Pool Deck Heaters
	• Recreation: Camp Cookers, Heaters, Lights, Refrigeration
	• Infra-Red Heaters
	• Laundry Equipment
	• Greenhouses
	• Incinerators
	• Roofing Kettles
	• Street Pavers
Industrial	• Standby Fuel
	• Flame Cutters
	• Metallurgical Furnaces
	• Industrial Dryers
	• Electric Generation
	• Construction Heaters
	• Refinery Fuel
	• Solvents
	• Raw Materials
Agricultural	• Crop Dryers
	• Flame Cultivation
	• Weed and Insect Control
	• Tobacco Curers
	• Poultry Brooders
	• Stock Tank Heaters
	• Pig Farrowers
Internal Combustion	• Farm Tractors
	• Industrial Lift Trucks
	• Industrial Sweepers
	• Stationary Engines
	• Bus and Truck
	• Automobile
	• Portable Engines
	• Refrigeration

Source: NLPGA (1978)

Cooking

Modern LPG ranges are very similar to conventional gas ranges, and are equipped with top burners, ovens and broilers, which can be thermostatically controlled and automatically ignited by pilot burners. With only minor modifications, standard gas ranges can be converted to LPG use and vice versa.

Water Heating

LPG water heaters can be of the manual or automatic type. Their construction and outward appearance is again very similar to conventional gas water heaters in which the water in a tank is automatically held at a preset temperature. Instantaneous water heaters, which are less frequently encountered, are also available for LPG use. Such heaters essentially contain a long coil of tubing with a large capacity gas burner below it. They are preferred when large amounts of hot water are required at one time.

Refrigeration

LPG refrigerators employ absorption-type refrigeration in practically the same way as standard gas-operated refrigerators.

Heating

LPG space heating has, for the most part, been limited to the installation of room heaters and floor furnaces. Such heating units require the smallest investment and are easy to install. In many cases they supply auxiliary heat, e.g., where the existing system may be inadequate. A wide variety of room heaters and floor furnaces are commercially available. So-called "unit heaters," normally suspended from the ceiling, are primarily used in commercial and industrial applications, such as stores and shops, eating places, garages and warehouses.

Domestic Appliances

LPG home appliances which are used to a lesser degree are air conditioners and dishwashers. The outdoor grill is probably the most widely used LPG appliance today, particularly because it is so popular in urban areas as well. Other recreational uses of LPG, (e.g., for camping, operation of recreational vehicles, pleasure boats, etc.) has also increased considerably in recent years. Miniature cylinders containing 2 lbs or less of propane are widely used for hand torches, lights, and camp stoves. Twenty-pound cylinders are most commonly used by small domestic users or for trailers, temporary installations, and other applications where portability is most important.

7.2.2 Commercial Applications

The commercial applications of LPG range from those in neighborhood restaurants and bakeries to food processing. Again, most of such applications are found in rural areas. Military camps are also often large users of LPG.

Typical restaurant utilization of LPG may include the use of a heavy-duty-type range for cooking, baking, frying, broiling, roasting, and even for popping corn. Lunch wagons almost always depend on LPG. In bakeries, the bake ovens operated on LPG are preferred over crude fuel ovens because they are more efficient and require less space. Since the various sections of such bake ovens can be individually heated with gas burners at different temperatures, their versatility is superior to crude fuel ovens. Other utensils common in the restaurant and bakery business, such as coffee makers, dry food warmers, toasters, deep fryers, refrigerators, dishwashers, and space and water heaters are also available in LPG versions in a wide variety of sizes.

7.2.3 Agricultural

Ranchers and farmers are particularly heavy individual users of LPG since only rarely do they have piped gas on their premises. They usually have LPG storage tanks between 1,000 and 10,000 gallons capacity. However, some use bottled gas. The largest volumes of LP gas are probably used for tractors and other types of mechanical farm machinery, such as grain harvesters, irrigation pumps, etc. In another wide variety of applications, the heat obtainable from burning LPG is utilized for crop drying, flame weeding, tobacco curing, defoliation, poultry and pig brooding, stock tank heating, greenhouse heating, frost protection in orchards, dehydration of alfalfa hay and processing honey. Dairies also use LPG for much of their power and heating needs.

7.2.4 Industrial Applications

It has been estimated that there are more than 10,000 industrial applications of LPG (Clifford 1952). Again, the heaviest industrial use of LPG appears to be in remote areas. This would apply particularly to the food processing industry that is often located quite close to the producers. LPG is used here both for heat processing of food and for use in refrigeration of perishable food products. Food processing may include food ripening and coloring, drying, cooking, roasting and smoking. In other industrial applications, LPG is used for small power generation in which an LPG fuel combustion engine is coupled to an electric generator.

Light industry utilizes LPG in a wide variety of applications which include such different activities as welding and flame cutting, flame soldering, pipe bending, melting tar, road clearing, heating starting plugs of diesel engines, barrel drying, drying of new plaster, retreading tires and blacksmith forging. It is even used for incinerating human waste in portable toilets on construction sites.

Heavy industry has been using LPG mainly for reasons of economy since in certain cases it proves to be more cost efficient than other fuels, particularly for enrichment of more conventional fuels and for peak load and standby uses. In other heavy industrial applications, LPG is used in a wide variety of heating processes, such as gas-fired furnaces. For example, butane is used in heat treating and hardening furnaces for automobile manufacturing. For continuous gas carburizing furnaces, a mixture of propane, butane, and air is used. This permits close control of carbon content, depth, and character of the carburized casing. For baking automotive finishes, a mixture of propane and butane is used.

As mentioned previously, the chemical industry is, by far, the largest consumer of LPG in bulk, using both propane and butane as a feedstock in the production of a wide variety of synthetic products.

7.2.5 Internal Combustion

LP gas has been used for many years as a fuel for internal combustion engines, particularly in farm tractors. Other types of vehicles, such as forklifts and mine locomotives, also have a long history of using LPG as a fuel. This is due to the clean burning characteristics of LPG, which allow it to be used as an internal combustion engine fuel for vehicles operating in enclosed spaces.

The use of LPG as a fuel for trucks, buses, and regular automobiles is increasing at present because LPG is available at a lower price than gasoline. Its clean burning characteristics and high octane rating are also adding to its popularity. The main hindrances to wider public acceptance as an automotive fuel are the high cost of conversion (from gasoline to LPG), the lack of refueling facilities, and the high number of miles that must be driven to recoup the original investment. It is also probable that if LPG gains enough popularity it will have a tax imposed on it similar to the gasoline tax. This would obviously take away part, or all, of its economic edge.

8.0 CURRENT KNOWLEDGE OF LPG RELEASE PHENOMENOLOGY

If release prevention devices or precautions fail, LPG may be released and dispersed into the environment as liquid and vapor. The effectiveness of release control methods then becomes crucial in preventing injury and damage in the vicinity of the spill. Release control practices depend primarily on knowledge of release progressions and consequences. Research and development activities relating to vapor generation and dispersion, pool fires and explosion phenomena can provide this knowledge. Even though there are quantitative differences in these phenomena among the various types of hydrocarbon fuels, there are nevertheless many similarities. Because LPG has not been researched and tested under accident conditions as thoroughly as some of the other liquefied fuels, such as LNG, much of the evidence presented in this section is actually based on knowledge and experience gained with LNG.

8.1 LPG VAPOR GENERATION AND DISPERSION

In the storage and transportation of LPG, spillage has occasionally occurred and remains a distinct possibility in the future. Whether spilled on land or on water, LPG will spread over the surface and will boil vigorously because of heat input from convective air currents and from the substrate, generating vapor at a high rate. In addition, the initial release of pressure in the LPG during spillage will cause flash vaporization of some LPG without contact with or heat from either land or water. The resulting vapor will often be colder than the surrounding air, and will condense moisture from the air, thus rendering the LPG vapor cloud visible. The vapor cloud is normally more dense than the atmosphere by 50 to 100%; and as a result, the LPG vapor cloud tends to reamin near the surface of the land or water. As the vapor is carried by winds, the LPG cloud will be gradually diluted by mixing the air, mainly due to atmospheric turbulence. Despite mixing an warming, the LPG vapor cloud never becomes buoyant as LNG vapor clouds normally do. However, by the time an LNG cloud becomes buoyant through mixing and warming, the buoyant portion of the cloud has usually been diluted below the lower flammable limit (LFL). Within a certain range of LPG concentration, the vapor cloud is flammable, and the greatest distance downwind to the LFL is of great interest with respect to the safety of people and property in the vicinity of an LNG spill. The greatest downwind distance to the LFL is affected by the size or rate of spillage, the prevailing dispersive character of the atmosphere, locale, specific conditions of the spill (height above the ground, tank or pipeline pressure, etc.), and local topography.

In measuring this greatest downwind distance, and in other measurements of fuel concentration to permit the development of predictive equations from tests, compromises have to be made because of the variability of the wind and the limited number of sensors which can be used. To accommodate the meandering of a plume caused by shifts in wind direction, several sensors can be placed in a plane normal to the expected wind direction. The readings of each sensor in a given time interval will then yield an average concentration for the sensor location. The maximum of these readings among all the sensors in the same plane is close to the "maximum average concentration" at that distance from the source. It is expected that such "maximum average values" would be comparable among various tests. However, the concentration value of prime interest in terms of safety is one which exists just above the lower limit of flammability for a sufficiently long time, so that an ignition source in this region would produce a flame burning in a sustained manner and propagating back toward the spill source. If one could predict the average concentrations related to the peak concentrations existing throughout a volume sufficient to propagate a flame, the measurement of the average would be sufficient. However, the scant amount of published data on this matter indicates that the relationship changes at least with wind velocity, weather stability class, and possibly with distance from the source, as well as with the concentration level itself. An adequate study of the relationship between peak and average concentrations would therefore be desirable, although it is realized that this would be a difficult task.

The following discussion brings together the current knowledge and the published data on vapor generation and dispersion in LPG spills. As noted above, the more extensive information on LNG spills is cited here to provide wind. The presence of a dike might confine the vapor cloud from sufficiently small spills until diffusion and wind disperse the vapor cloud. In a sufficiently large confined spill, the cold vapor could fill the dike and overflow, dropping to the ground and spreading in much the same way as the vapor cloud from an unconfined spill.

The high density of the LPG cloud can result in the vapor flowing almost like water (e.g. through natural drainage channels, ditches, sewers, culverts, creeks). Eventually, as the vapor cloud spreads and becomes more shallow, the gravitational forces become less important (because of a loss of gravity "head"), and the atmospheric forces become more important in the motion of the cloud. It then begins to disperse (due to turbulence) and move downwind. Such turbulent dispersion will be enhanced by the presence of structures near the spill, such as tanks, dikes, buildings and ships.

The LPG concentration in the dispersing cloud will decrease from the spill site in more or less all directions; of primary concern is the possible maximum distance from the spill to the LFL. Since the concentration of LPG vapor in the cloud is not uniform, local pockets may develop which are rich or lean as compared to an average vapor concentration. Further complicating the problem of predicting cloud behavior is the fact that dispersion of the vapor cloud occurs close to the ground and that the atmospheric velocity gradient in the boundary layer can strongly influence the dispersion and the safe (maximum) distance to the LFL. The importance of estimating the distance to the LFL is not so much to estimate the distance accurately as to provide a guide for siting, construction materials and approaches to controlling vapor dispersion so as to prevent ignition at or near the spill site.

So far, few studies of LPG spill dispersions have been reported in the literature. They will be discussed in the next section.

8.1.2 Analytical and Experimental Work with LPG

Hardee and Lee (1975) developed a three-stage model for cloud growth, following the spill of pressurized liquid fuels (including butane and propane). The model is developed on analytical reasoning, but is compared with experimental data for small spills. The first stage of their model is useful additions to the rather sparse data on LPG. Generally, experimental data and their analyses will be discussed separately from theoretical studies. It must be recognized, in comparing data, that different experiments may have had different objectives and thus have been carried out with different limitations. The vagaries of the weather and the management of large data gathering systems have often presented serious problems especially for large-scale spills in the open.

8.1.1 Physical Description of LPG Spill Phenomena

An accidental spill of LPG on land or on water involves a sequence of events which are of importance in determining the nature and effect of a given spill. Most spills will probably involve the simultaneous occurrence of several processes, all of which are coupled, and all of which may be strongly influenced by certain key factors of the spill.

In its simplest form, an LPG spill may occur as a discharge of a continuous stream of LPG. Some will flash or vaporize while the stream is still up in the air, and the remaining LPG will begin to form a boiling pool or several pools on land or water.

High velocities in the liquid LPG stream or jet can entrain liquid droplets in the vapor flow and thus enhance vapor generation. They also can result in significant spreading of the liquid pool on land or water by splashing, thus favoring the formation of several small LPG pools. Subsequent heat transfer from the land or water surface will vaporize the LPG, thus generating a cloud of cold vapor denser than ambient air.

The rupture of an underground pipeline may result in the formation of a crater, forming a natural partial confinement for the LPG pool. Unconfined spills are possible on flat land, e.g. from tank trucks, rail tank cars, aboveground pipelines, or storage tanks. Large spills on water, such as from an LPG ship accident, would also be unconfined. Since such spills often have a much larger heat transfer area between the surface and the pool, their vaporization rate could be much higher than for a confined spill. Once the cold vapor cloud has been formed over the spill, it will tend to disperse due to gravity (negative buoyancy), diffusion, and local an expansion process, wherein the liquid expands from its storage tank pressure to atmospheric pressure. The resulting two-phase flow is considered to be isentropic. During the second and third stages, the cloud entrains ambient air at constant total momentum, hence the expansion velocity decreases as the cloud expands. During Stages 1 and 2, the vapor cloud expands radially and vertically; during Stage 3, only radial growth is considered. Wind and atmospheric turbulence are not considered in dispersion of the vapor cloud. The model has been applied to the simulation of the dispersion of several fuel vapors, including propane and butane. Experimental results were obtained for 1-lb and 64-lb spills of a welding gas known as MAPP. Comparisons of theoretical predictions and measured propane cloud dimensions are also presented for a 930-lb release in a forested area. Data for all releases are presented for times up to 5 seconds after the spill and up to a 100-ft cloud radius.

Reid and Smith (1978a) studied the boiling of LPG on water and, in a few experiments, on ice. In contrast to previous tests using other cryogens, mainly LNG, LPG showed an initially violent reaction upon spillage. Water and ice were thrown into the vapor cloud for a few seconds after the spills began. Subsequent boiling was of a quieter nature and was described by a simple, one-dimensional heat transfer model. Tests were conducted using an insulated calorimeter, initially containing distilled water, the surface area being 30 in.^{2}. Liquid spills involving a few hundred grams of LPG or propane were made on water or ice. The mass rate of boiled LPG was found to vary as the square root of time for all tests. Additional work by Reid and Smith (1978b) has been reported involving the boiling of other liquids on water, ice, and other (gel) surfaces. In the work reported by Reid and Smith (1978a,b), the spills were confined by the calorimeter walls.

Valentine and Moore (1974) describe a transient mixing process for propane and air in a confined space, where stratification results in pure air immediately above a flammable mixture of propane and air. Experiments were conducted using an air filled vertical pipe, 6 inches in diameter and 8.2 ft high. Propane vapor was introduced through the base of the pipe and samples for gas chromatograph analysis were drawn at selected time intervals.

The details of the loss of LPG from a pipeline rupture are given by Bearint et al. (1976) where the complex interactions of pumping stations and line profile are discussed. The phenomenology of a storage tank (or truck or rail car tank) rupture can be derived from Bearint's discussion by discounting the effects of pumping stations and line length. Hardee and Lee (1973) also discuss the mechanism of LPG spills in connection with subsequent fireballs, and Lathrop and Walls (1974) discuss LPG spill events in a paper dealing with fires involving LP gas trucks.

No other reports of experimental studies of LPG dispersion were found in the literature, other than wind tunnel studies which are discussed in a later subsection of this report.

8.1.3 Studies of Other Heavy Gases

A great deal of experimental work has been done on the spillage of other heavier-than-air gases, including chlorine, some Freons, and LNG. Mackay and Matsugu (1973) considered the evaporation of several liquid hydrocarbons from land and water. Experimental correlations were made between heat transfer coefficient and wind speed, pool size and vapor phase Schmidt number. Estimations of evaporation rates are also discussed. Tests were conducted using 4-x 4-ft and 4 x 8-ft pans, 0.75 inch deep. Jeje and Reid (1974) studied the boiling of liquid hydrocarbons on water, using LNG, liquid methane, liquid ethane, and also liquid nitrogen (LN_2). Boiling rates were measured with a load cell recording the liquid weight. Confined spills were carried out in a 4-in.-inside diameter vessel. Liquid volumes on the order of 0.03 gallons were used in the tests. Different correlations for heat transfer coefficients were developed for each fluid tested. Kalinin et al (1975) studied heat transfer in boiling cryogenic liquids, considering LN_2, CCl_4, n-pentane, and Freon 113; an analytical model for boiling was developed and compared with experimental results.

Experiments on the dispersion of chlorine and Freon were conducted in the Netherlands by van Ulden (1974). Several types of experiments were conducted, including those directed at the spreading of heavy gases. Twenty-two hundred pounds of Freon 12 were spilled in stable atmospheric conditions with a wind velocity of 6.7 miles/hour. Concentration measurements were made at downwind distances of 330 ft, and 3280 feet. Three phases were noted: an instantaneous release phase, a gravity-driven outflow phase, and an atmospheric dispersion phase. Approximately Gaussian distributions of concentration were measured at 1640 and 3280 ft downwind distances. Comparison of experimental results with calculations of downwind concentrations using an unspecified Gaussian model showed significantly greater horizontal spreading and less vertical spreading than would be expected from the analytical models for a neutrally buoyant gas.

The generation and dispersion of vapors of LPG can be expected in many respects to be similar to the vapor generation and dispersion of heavier-than-air gases. The similarities mainly result from similar two-phase liquid/vapor properties. Since the liquids are usually stored under pressure at or near atmospheric temperature, some problems associated with cryogenic storage are not present with these vapors. Further, because the vapors or gases are all denser than air, they are similarly affected by gravity spreading. (Note that methane is denser than air only at temperatures below -160°F.) Even though differences in density must be considered when applying the results obtained for the dispersion of one dense gas to the dispersion of another, a very important factor is that both gases are always more dense than air. Thus, when comparing the generation of LPG vapor with the generation of vapor from other volatile and pressurized liquids, or with the release of pressurized gases, a great deal of similarity in the physical processes is to be expected. High energy streams or jets of liquid and/or gas usually result in a wide dispersion of the vapor or gas from the spill in all these cases.

8.1.4 LNG Experimental Work

The bulk of the experimental work on the generation and dispersion of liquefied gases has been done with LNG. Even so, the work described does illustrate the general character of various phases of liquid hydrocarbon spills, vapor generation, and dispersion. As already noted above, LPG would behave similarly.

Rate of Spread of Spill

The rate at which vapor is produced by an LNG spill is again controlled basically by the area of the spill surface in contact with the heat source, and by the characteristics of that heat source. If a spill spreads slowly, its vaporization rate would be reduced accordingly. Further, depending on the pattern of the spread and the change in temperature of the contacted surface, the rate would change with time. On the other hand, for a steady-state spill, the area grows until the vaporization rate matches the supply rate.

In the case of land spills in a diked area, it is sometimes assumed that the dike is filled instantaneously, but there seem not to be any computations of the spread rate over solid surfaces of various properties. Since supporting information is available (Peterson et al. 1969, Gideon et al. 1974, Smith and Reid 1977), there would seem to be no difficulty in computing the spread rate of a spill on a flat solid surface or one with a simple slope following the example of spreading of spills on water. Little consideration has so far been given to unconfined spills on land; although, Arthur D. Little, Inc. (1974) has performed some work in this area.

There are several experimental and theoretical analyses available for LNG spills on water (R&D Assoc. 1976). Some of the most extensive data were taken by the Bureau of Mines (Burgess et al. 1970a,b), though data are also available in reports from Shell (Boyle and Kneebone 1973, Duffy et al. 1974), ESSO-API (Feldbauer et al. 1972), and Tokyo Gas (1971). Burgess et al. (1970 a,b) suggest that the diameter of the spill area varies linearly with time, even though each set of data seems to fall off with time. Individual points of data were not given by Shell, but the curves did fall off with time when diameter was plotted against time. Gideon et al. (1974), Otterman (1975), and others pointed out the applicability of Fannelop and Waldman's similarity analysis (Fannelop and Waldman 1971), which leads to a linear increase of area with time when inertial-gravity conditions are controlling the spread rate. Fay (1973) reached a similar conclusion. This linear area relationship affords a satisfactory correlation of the data.

The analysis has also been extended to include the effect of a constant rate of vaporization and a finite, short spill time. While the Fannelop-Waldman type of relation is now generally accepted, it does involve similarity assumptions, and it is suggested that some cases be computed in a more exact manner for comparison.

Vaporization Rate

Maximum vaporization rates of LNG spills on land and water at ambient temperature appear to be about the same. This is reasonable, assuming a film boiling limitation on the rate. However, the change in vaporization rate with time is quite different in the two cases. For confined land spills (e.g.,within a diked area), the vaporization rapidly reaches a maximum when the dike floor is covered. Then, as heat is removed from the floor, the rate of vaporization decreases. For simple solids and film boiling, rigorous solutions are available for the vaporization rate. For water spills and unconfined land spills, the total rate of vaporization increases as the LNG spreads out. The formation of a solid sheet of ice in a spill on water would result in a decrease of vapor flux per unit area, as in a land spill. However, such sheets apparently do not form, (Jeje and Reid 1974) and the mixing near the water surface results in a more nearly constant heat flux to the surface. For other hydrocarbon spills on rough water, large increases in evaporation rate have been reported, only part of which can be explained by an accelerated spread rate (Mackay and Matsugu 1973). Thus, while a spill on land shows a rapid rise in vapor production followed by a slow falling off, a spill on water shows a slower rise, with a leveling off, and then a gradual falling off as the thin part of the LNG film breaks up and disappears (Feldbauer et al. 1972).

Vaporization rates in land spill experiments have usually been determined by liquid level or pressure recession rates, and in laboratory experiments by rate of weight loss using load cells under the whole apparatus. Vaporization rates in unlimited water-spill experiments are derived from spill rate and area, for steady-state cases; or from spread rate, quantity spilled, and time, for nonsteady cases. For spills on water in the laboratory, recession rate measurements have been made using load cells, although the measurements may sometimes be complicated by the expulsion of water and the formation of hydrates. Vaporization rates of light hydrocarbon mixtures on water have been studied most recently at MIT (Jeje and Reid 1974, Drake et al. 1975 a, b, Smith and Reid 1977b), earlier by Shell (Boyle 1973, Boyle and Kneebone 1973), Tokyo Gas (1971), and the Bureau of Mines (Burgess et al. 1970a,b). It appears that the vaporization rates have been measured with sufficient accuracy for the pure cryogens and for some mixtures. However, some questions remain related to change of vaporization rate with time, effects of composition, water temperature, and turbulence level on the water surface.

Vaporization rates for soils and for other dike materials have been measured in field experiments directed primarily toward dispersion control. The data are sufficiently accurate for purposes of estimating hazards. Unexpected maintenance of high rates in some studies has been explained on the basis of soil cracking (Duffy et al. 1974, Conch Methane Services n.d., Humbert-Basset and Montet 1972, Reid and Smith 1975, Drake and Reid 1975, Arthur D. Little, Inc. 1974).

Gravitational Spread of Vapors

As the liquid spill is vaporizing, the vapor is building up in a pancake-shaped cloud, which may be carried downwind. With low wind velocity, a large cloud remains nearly symmetrical and may be treated as a "puff." With high wind velocity, a small spill may be treated as a quasi-steady-state evaporation process relative to the wind. Calculations suggest that land spills in diked areas will frequently fall into the latter category. For the maximum spill of 2700 gal in the ESSO-API tests (Burgess et al. 1970a), the turnover point appeared to be at a wind velocity of between 4 and 8 fps.

Vapor clouds of pancake shape are especially apparent for instantaneous unconfined spills (Feldbauer et al. 1972). The extent to which this spreading is important in determining the final dispersion pattern has been pointed out in various publications (Germeles and Drake 1975, Fay and Lewis 1975, Lewis 1974, Raj and Kalelkar 1974, van Ulden 1974, Buschmann 1975).

Unfortunately, there appear to be no experimental data on LNG spills to verify the puff theory. The ESSO-API (Feldbauer et al. 1972) data appear to fall into the quasi-steady-state analysis regime, except for two runs which were apparently unsuitable as data sources. Kneebone and Prew (1974) present data on the steady-state spill from a moving ship in which "instantaneous" evaporation is assumed, followed by lateral spreading of the vapor. In the San Clemente program (Duffy et al. 1974), photographic records were made of LNG spills. These were analyzed to obtain data on plume height and width near the origin of the spill as a function of time. It would seem that van Ulden (1974) has made the only valid check of gravity-spread theory against experiment.

Dispersion of Fuel Vapor By Wind

As the vapor cloud is being carried downwind, it disperses by mixing continuously with air, so that the concentration of fuel diminishes with distance, x, from the source, as well as with cross-wind distance, y, from the plume centerline and vertical distance, z, from the ground plane. It can thus be expected that the peak concentrations will generally occur at or near ground level on the plume centerline, and hence, can be considered as functions of distance, x, only. Turbulent diffusion models ordinarily predict maximum average values of concentration, C, by relations of the form

$$\frac{C}{\dot{M}} \sim \frac{T}{U} \text{ function } (x),$$

where, $\dot{M}$ is the steady-state source strength, U is the average wind velocity, and the coefficient T is an empirical turbulence factor. This turbulence factor has been tabulated in terms of cross-wind and vertical dispersion coefficients, σ_y and σ_z, as a function of distance x, for various classes of weather (Turner 1969). The customary use of these atmospheric dispersion equations in analyzing vapor dispersion data from LNG spills and in predicting downwind concentrations has been treated in several references. For instance, Welker (Duffy et al. 1974) used the Brookhaven National Laboratory curves (Turner 1969, Smith 1968) in analyzing the San Clemente spill data and Drake, (Duffy et al. 1974), used the Gifford-Pasquill curves (Turner 1969, Gifford 1961).

It was recognized very early by various investigators that the negative buoyancy effects tended to flatten the plume, that is to decrease the value of σ_z/σ_y, so that the horizontal dispersion coefficient was increased and the vertical dispersion coefficient was decreased. However, it was usually assumed that the product of the two, for quasi-steady-state conditions, would give an accurate representation along the line of maximum concentration. Comparisons with data indicated that this might be a reasonably good assumption (Duffy et al. 1974).

May and co-workers (May et al. 1973a,b, Feldbauer et al. 1972) indicated that their water spills at Matagordo Bay might have been accompanied by more stable weather conditions than indicated by the normal interpretation of the air temperature gradient and wind velocity. For instance, recent research (Raynor et al. 1974a, b, Michael et al. 1973, Panofsky 1974, Slade 1962, Parker 1970) has indicated that when water temperature is lower than the air temperature, strong inversions can exist in the lower levels of the atmosphere. If dispersion data under these conditions were interpreted as being characteristic of a neutral atmosphere condition, this might result in an excessive estimate of the most dangerous spill condition.

The peak-to-mean composition is obviously a significant parameter in determining the limit of concentration beyond which there could be no flame propagation upwind. However, there are problems both in defining peak-to-mean composition and in measuring its value. For steady-state discharge of a gas into air, one can observe that the downwind plume usually meanders back and forth. As a result, a composition measurement will show peaks and valleys and thus give a series of peak-to-mean measurements. For measurements farther from the mean axis of flow, these peak-to-mean values (due to meandering) tend to increase (Burgess et al. 1970b, Singer 1961, Ramsdell and Hinds 1971). However, they apparently are not pertinent to the present problem and it is interesting to note that they will not occur in wind tunnel modeling.

A more pertinent peak-to-mean value is the ratio of the peak concentration in a given downwind cross section from a steady-state source to the value on the plume centerline averaged over a short period of time.(a) Lewis (1974) and Fay and Lewis (1975) carried out this type of analysis on the ESSO-API spill data using an average concentration, $\bar{C}$, defined as the sum of concentrations over the total signal time divided by the total signal time at each station for which data were analyzed. While this would be a satisfactory base for an idealized "puff"-type spill (i.e., large water spill with low wind), the data do not appear to correspond to puff conditions (as discussed above). Their analysis produces a relation for the probability that the concentration ratio, $C/\bar{C}$, exceeds a given value. The authors conclude that the relation is independent of size of spill, wind speed, and distance from the spill. Erroneous results can be obtained, however, from using this analysis with a quasi-steady-state spill, without proper adjustment of $\bar{C}$. Further, the wind class for the ESSO-API runs appeared to have been quite constant. Since it is felt that the peak-to-mean values would show a lesser spread of data with the more calm wind conditions, use of the Lewis-Fay curve for all wind conditions would be ill-advised at the present time. The data were also limited by the slow response time of the instrumentation (order of 10 seconds). The response time is not short enough to give data on cells of the probable size that will just permit flame propagation upwind.

In drawing conclusions from the tests, the small range of base data relative to wind velocity, spill size, and distance from spill, as well as the scatter of the data, must be noted. Therefore, further data of this type would be needed before probability distributions of peak-to-mean concentrations under various wind conditions over land and water can be predicted with some degree of confidence.

(a) If the source is constant, this time is related to the time that the plume centerline can be maintained under observation. If the source is varying, the time will be related also to the accuracy with which the data may be normalized.

8.1.5 Mathematical Modeling

Mathematical modeling of LPG spills has received only a little attention in the literature. Burgess and Zabetakis (1973) applied a simple Gaussian dispersion model to the analysis of an LPG accident at Port Hudson, Missouri, in 1970. Other than this study, little has apparently been reported on LPG vapor generation and dispersion. Considerably more work has been done on other heavy gases, including chlorine, Freon, and LNG. Since much of this work can be applied to the analysis of LPG spills, a review of the LNG-oriented modeling work is useful at this point. However, before proceeding with the description of a few specific mathematical models, some general features will be discussed which must be considered in the development of such models.

8.1.6 Modeling Considerations

A mathematical model for the generation and dispersion of vapor from a large LNG or LPG spill can be very complex if it is to take into account all the factors affecting the dispersion of and ultimate safety from the vapor cloud. Of considerable importance are 1) spill rate, 2) pool spread (and confinement, if any), 3) vapor generation (heat transfer from ground or water), 4) gravity-driven spread of vapor cloud, and 5) later, turbulence-driven, atmospheric dispersion of the cloud. These and other aspects are discussed briefly below.

Spill Rate

In most modeling efforts it is assumed that there is either an instantaneous spill or a rapid, continuous spill. In an accidental LPG spill, it is unlikely that either assumption will be valid: real spills may well begin at a high flow rate which decreases with time as the internal tank pressure decreases.

Spread Rate

The rate of spreading of the LPG pool has a strong effect on the vaporization rate and distance to the LFL, which is more important in unconfined spills than in spills contained in a dike. Adequate models of the spreading of the liquid pool must take into account the spill rate, the evaporation rate, and the details of the surface (and confinement) where the spill occurs.

Heat Transfer

The rate of heat transfer between the subsurface land or water and the LPG pool is also of key importance in determining the vaporization rate. In spills onto water, there is a large heat source of a high effective thermal conductivity available to attain and maintain high vaporization rates.

Vaporization Rate

The presence of a free jet of LPG would seem to be a probable occurrence in some accidental LPG spills. Initial flashing and breakup of such a jet would seem to increase the anticipated vaporization rate in a difficult-to-predict manner.

Gravity Spreading

The negative buoyancy of the LPG vapor cloud has a strong effect on vapor dispersion in both confined and unconfined spills. In a confined spill, the negative buoyancy of the vapor may initially trap it within the dike. When the vapor cloud overflows the dike, the negative buoyancy may accelerate the spreading of the vapor cloud. With no wind, the LPG vapor could spill over all sides of the dike, leading to a more complex initial dispersion problem than the usual case of spillage over the downwind side of the dike. For unconfined spills, the hydrostatic head in the vapor cloud would cause the cloud to flow uniformly in all horizontal directions on level surfaces but downhill in hilly terrain. Although during the gravity spreading phase the wind forces on the vapor cloud have been generally neglected in favor of the gravitational forces, they can have a significant influence.

Turbulent Dispersion

The turbulence mechanism for vapor dispersion dominates the final portion of the vapor dispersion process as it is usually modeled. Some turbulence models are available but are mathematically very complex and are difficult and expensive to use.

Wind Condition

The vapor cloud will generally drift with the wind, but winds tend to change both speed and direction almost continuously. The velocity gradient in the earth's boundary layer may also be important to dispersion.

Near- and Far-Field Effects

It is in the far-field region that ultimate dispersion of the vapor occurs and where the population may most likely be affected. The far-field can be characterized by low concentrations of vapor and rather slow transients of the average concentration. Far-field regions are usually characterized in models by essentially flat topography whereas the near-field region vapor concentrations are higher, transients are more rapid, and the effects of nearby structures must be considered. The near-field region would appear to be the more difficult to model well. Certainly, the near-field provides the boundary and initial conditions for far-field modeling and this alone makes it necessary to accurately model near-field phenomena.

Transient versus Steady-State Effects

Although an LPG spill is a transient phenomenon, the high complexity of transient models may lead to the preferred use of steady-state approximations. Some use of transient models remains necessary, if only to indicate where the simpler steady-state models are valid.

The nine considerations discussed above are important in any given LPG spill and provide a set of guidelines for evaluating various mathematical models. Strict consideration of all aspects would result in complex, expensive computer models. Less complete models may often be adequate and useful if they can be verified by more exact models and more experimental data.

The early modeling work of Burgess and Zabetakis (1973) was presented as a part of an accident report for a specific LPG accident which resulted in a significant drift of a vapor plume (approximately 980 ft), from the LPG source to an ignition point. Two different models for atmospheric dispersion (Turner 1969, Havens 1977) were used in that study to predict horizontal and vertical dispersion, particularly the UFL and LFL for up to 1500 ft.

Published mathematical models of spills onto water were recently reviewed comprehensively by Havens (1977) and Harsha (1976). Havens' review considered only the mathematical modeling of LNG vapor dispersion from large (25,000 m^3) spills onto water. Harsha's review considered both land and water spills, but his primary concern was on land spills. Because of their timeliness and detail, these two recent reviews will form the basis for the following description of the state of the art of mathematical modeling of vapor dispersion. These reviews appear to include most if not all models described in the literature for modeling LPG vapor dispersion.

8.1.7 Havens' Review

Havens performed a detailed analysis of seven models which have been developed for describing LNG dispersion phenomena. He compared the results predicted by each model for the downwind distance to the LFL (5% time-averaged concentration) for a standard spill scenario: an instantaneous spill of 6.6 million gallons of LNG. Although Haven's standard test is for LNG, not LPG, and is for a much larger spill than is probable for LPG transport, his review and Harsha's review do show relative variations among the models considered.

The seven models considered by Havens are:

1. U.S. Bureau of Mines - (Burgess, et al. 1970a, 1972)
2. American Petroleum Institute - (Feldbauer et al. 1972)
3. Cabot Corporation - (Germeles and Drake 1975)
4. U.S. Coast Guard CHRIS (Chemical Hazard Response Information System) - Arthur D. Little, Inc. (Raj and Kalelkar 1974)
5. Professor James Fay, Massachusetts Institute of Technology (Fay and Lewis 1975)
6. Federal Power Commission (FPC) 1976
7. Science Applications, Inc. (SAI) 1975a

These models were categorized to be

1. Classical dispersion models used to describe the motion of neutrally buoyant materials in the atmosphere. Essentially, these are air pollution models and assume low concentrations of pollutants, and assume a Gaussian distribution of pollutant downwind of a point or area source.

Two subcategories were identified:

- A "puff" model, used by Fay, by Germeles and Drake, and by CHRIS, wherein the pollutant is instantaneously released into the atmosphere in a cloud which then disperses as it is moved by the wind.
- A plume model, used by Burgess, by Feldbauer, and by the FPC, wherein the pollutant is continuously released at a steady rate.

2. A type of model used only by SAI, based on solutions to the combined mass, momentum, and energy equations. This model is considered proprietary by SAI, and its details were not fully discussed by Havens.

Different atmospheric conditions were assumed for each of these models. Usually a stable atmosphere was assumed, but SAI and FPC assumed a neutral atmosphere--the FPC used Pasquill D (Pasquill 1961) dispersion values, and the API assumed that the vertical dispersion was characterized by Singer and Smith (1966) D and the horizontal dispersion by Pasquill C conditions. Another factor affecting these results is the vaporization rate and the gravity spread of vapor before turbulent dispersion begins. The vaporization rates range from a low of 1.4 x 10^5 ft^3/sec (FPC) to a high of 2 x 10^6 ft^3/sec (Burgess). Two of the models do not consider gravity spreading of the vapor cloud (and thus are inappropriate for LPG studies) and four models make specific, but different, assumptions about this effect. Another factor affecting the final results is the assumed nature of the source, either a point source (two models) or an area source. The assumed nature of the area source has a strong effect on the results given by the API model.

Havens draws several conclusions as to the sensitivity of the various models to certain key factors, as to which models provide unrealistic results due to basic assumptions, and which models are preferable and/or more likely to provide usable, accurate predictions. In evaluating the six atmospheric diffusion models, Havens uses four key factors:

1. Atmospheric turbulence (stability)
2. Area source allowance
3. Vapor release rate
4. Allowance for gravity spread and air entrainment.

One of these models was selected for further development and evaluation; the others were dismissed as providing unrealistic predictions or less plausible estimates of the dispersion process. The FPC model is judged to be unrealistic because it uses an "unrealistically low vapor release rate and . . . neutral atmospheric stability characteristics" (FPC 1976). The Burgess models do not account for any heat transfer or momentum transfer effects and thus are judged to provide unrealistically conservative results. The API model allows air entrainment during the gravity spread, based upon observations of smaller (10 m^3) spills. For the spill Havens considered, this assumption leads to a 2-mile-wide cloud before turbulent dispersion begins, using a series of point sources along a line perpendicular to the wind as the source for the dispersion model. The scale-up from 10 to 25,000 m^3 was considered "uncertain" and the source simulation was considered "not realistic" by Havens. These considerations may be inappropriate for smaller LPG spills. The Fay, CHRIS, and Germeles and Drake models provide similar results for distance to the LFL as they were used by Havens, and he judges that the Germeles and Drake model is "more plausible" than the other two models because it incorporates a "rational, if simplified, description" of the gravity spread phase.

Criticism of Havens' review is that he did not consider the chronology of the development of any of the models reviewed. Considering only the six atmospheric dispersion models, the Germeles and Drake model is the most recent except for the Fay and the FPC model. The Fay model was reported in 1975, but was in use and reported by Fay in 1973. The CHRIS model was based on earlier atmospheric diffusion work and some new models for pool spreading and evaporation, developed by 1974. The FPC model is also based on a combination of new and old models and appears to have been assembled recently, the primary criticism being of some of its assumptions. The Burgess, API, and Fay models predate the Germeles and Drake model, by 2 to 5 years, so it is hardly surprising that the later model is preferred.

The SAI model is unique among those reviewed by Havens in that it alone is based on a solution to the mass, momentum, and energy conservation equations. For a larger spill, this model predicts a shorter downwind distance to the LFL than any but the FPC model. Havens attributes this to the "predicted highly turbulent motion and associated air entrainment induced during the gravity spread of the cloud."

Havens presents five recommendations based upon his review and indicates the order in which all five recommendations should be enacted:

1. The SAI or similar models should be evaluated further. Specifically, a) turbulent mass, momentum, and energy transfer models should be reexamined; b) the confidence level of the values used for turbulent transfer coefficients should be estimated in an error analysis; c) the sensitivity of the results to uncertainties in those coefficients should be determined; d) the effects of assumed models for pool spread, vapor generation, and heat transfer on the predicted results should be determined; and e) the numerical techniques should be critically evaluated.
2. The SAI or a similar model should be used to characterize the relationship between spill size and distance to the LFL.
3. The Germeles and Drake model should also be used for a similar characterization and the results compared.
4. Experimental data requirements should be reviewed for experimental verification of results predicted for small spills.
5. Experimental spills should only be done to verify model predictions; large demonstration spills should be avoided without heavy reliance on models.

Even though these recommendations were formulated for LNG spill dispersion analysis, they can also stand for LPG spill analysis, since they are independent of the specific gas properties.

8.1.8 Comments on Models Reviewed by Havens

The seven models reviewed by Havens can be compared on the basis of the nine modeling considerations mentioned above. The models have already been compared to experimental data during their development; but since no large-scale spills have been conducted, the ultimate comparisons cannot yet be made.

The Burgess model is the oldest model of those considered and it is the least detailed and least adequate when compared to the model requirements. This model has been used to study instantaneous spills, but it and the other "plume" models may also be applicable to steady, continuous spills. Burgess' model does not consider:

1. pool spreading,
2. heat transfer,
3. gravity spreading of the cloud,
4. detailed turbulence, or
5. area source correction.

In common with other "puff" and "plume" models, Burgess' model considers atmospheric turbulent dispersion only in terms of empirical coefficients and, hence, can to a degree simulate the effects of turbulent dispersion but not model the process. Since these dispersion coefficients were developed for atmospheric pollutants in very low concentrations which do not disturb the flow field, these models would not be applicable to near-field conditions due to the high concentration of LPG vapor in the air. The Burgess model seems to be well suited only for continuous or steady dispersion problems. Vaporization modeling is based on experimental data and is not coupled to any heat transfer phenomena.

The API model is of the same type as Burgess' model and may be viewed as an improved version in two respects. The API model does consider the finite area of the source after a vapor cloud has mixed with air over the pool. The vapor spread and air entrainment of the API model leads, in Havens' opinion, to the widest vapor cloud of any model, and this accounts, in part for this model's relatively short distance to the LFL.

The FPC model represents again an improvement over the Burgess and API models in regards to a more adequate representation of pool spread and gravity spread. The FPC model considers vapor release from the gravity spread cloud to be controlled by heat transfer only from the air to the top of the vapor cloud, resulting in very low vapor release rates as compared to experiments or other models. Only the FPC model controls the release of vapor in this restrictive manner.

The Fay model is the oldest of the three "puff" models considered by Havens. It is apparently well suited to an instantaneous spill, but it would be less appropriate than plume models for continuous spill modeling. The Fay model does consider both pool and cloud spreading and includes heat transfer from the substrate to determine the vaporization rate.

The CHRIS model appears not to consider gravity spread of the vapor cloud or heat transfer for control of vaporization rate. The arbitrary assumption is made that the pool vaporizes instantly after spreading to its maximum extent. In light of the other models, the CHRIS model appears to be too simplified for quantitative use.

The Germeles and Drake model is the most detailed model of those based on the pollutant dispersion equations. It considers pool spread, heat transfer control of vaporization rate, gravity spread of the vapor cloud and air entrainment in the vapor cloud. Extended area sources are simply considered as upwind virtual point sources as in all "puff" models above and the FPC model.

All puff and plume models discussed above are based on similar, empirically-based dispersion coefficients which are strictly valid only for dispersion of low concentration pollutants whose presence does not disturb the flow field. Since the presence of obstructions to air flow and high concentration of vapor are common to near-field regions, these dispersion models are, therefore, not well suited to modeling near-field dispersion. Several of the more recent models of this type have included near-field considerations to improve predictions in both near and far fields. This undoubtedly improves their ability to more accurately predict far-field dispersion.

The simple puff and plume models are also limited to steady wind conditions and have been used primarily for instantaneous spills and only occasionally for continuous spills. The limitation on wind velocity appears to be difficult to overcome except perhaps by using a wind "rose" distribution for short spill times similar to the way a wind rose is used for long-term dispersion of pollutants in the atmosphere.

Havens concluded that the Germeles and Drake model was the best of this type of models presently available. When each model is judged relative to the above nine requirements for modeling LPG dispersions, the Germeles and Drake model is again the closest to an ideal model of all the puff and plume models considered.

The SAI model is distinctly different from the other six models reviewed by Havens. It does consider the area nature of the vapor source, pool spread, heat transfer both from the air and from the water, and gravity spreading of the cold cloud with turbulent air entrainment. It assumes vaporization at a constant rate per unit pool area and considers a three-dimensional flow field with a turbulence model. The solution is derived from the coupled equations for the conservation of mass, momentum and energy. This model should provide significantly more accurate predictions of dispersion in both the near and far fields. It does allow for interactions between the gas (LNG or LPG) and the atmosphere and is not limited by assumptions of low concentration. However, the SAI model is not well suited to solve the dispersion problem in a near-field region having air flow obstructions such as dikes, tanks, or buildings. Also, because it is new, complex, and proprietary, it is not yet well understood outside SAI, and as with large, complex computer models, it requires significant amounts of computation time. As Havens notes in his recommendations, further critical evaluation of certain aspects of the SAI model, particularly its turbulence model, is warranted. Other, still more complete and more complex models and approaches are being considered and developed as noted in the following section. However, at the present time, there do not appear to be any models available which are similar to or more detailed than the SAI model.

One of the basic limitations of all the Gaussian models is their lack of including atmospheric turbulence. Only empirically derived dispersion coefficients are used in the Gaussian models to account for concentration decreases in various directions from the spill. Both Havens' and Harsha's reviews point to the strong effect that the choice of wind conditions has over the distance to the LFL. Thus, the critical effect of these coefficients leaves open an important area for further study. A better understanding of the effect of the real wind conditions on the various categories of dispersion coefficients is desirable. Studies towards this end should include both theoretical and experimental efforts.

Another aspect of the same problem is determining what coefficients to use for correlating a given experiment. One approach to this problem might be to parallel each spill test with a smoke plume which could be observed, and thus provide better insight into the exact effects of the local wind on each experimental spill.

8.1.9 Harsha's Review

In 1976 Harsha performed for the American Gas Association (AGA) a review of certain mathematical models for vapor dispersion from LNG spills on both land and water, which included: 1) point, line, and area source Gaussian dispersion models; 2) multi-dimensional "hydrostatic" models; and 3) current techniques for further model development. The Gaussian models considered are similar to those reviewed by Havens above. The "hydrostatic" model is again the SAI model (Science Applications Inc. 1975b). The remainder of Harsha's review is devoted to currently available mathematical and numerical techniques which could provide the basis for more advanced models.

One of the major objectives of the AGA-sponsored program was the evaluation of passive control methods which would restrict flammable LNG vapor clouds to within one or two dike diameters of a land spill. Thus there was a need for mathematical models which would provide accurate predictions in the very near field, and which would specifically take into account the high turbulence near dikes and other flow obstacles. Also needed were models which could be used for predicting dispersion from LNG spills onto water, where the dispersion distances to the LFL are usually predicted to be greater than in land spills. In view of these objectives and areas of interest, a set of five requirements for adequate models was developed. These five requirements, which were partly based on wind tunnel test observations, were:

1. consideration of a full, three-dimensional velocity field,
2. a nonlinear mathematical formulation of the problem,
3. inclusion of time-dependent terms in the equations of motion
4. consideration of coupled energy, mass, and momentum equations with buoyancy terms, and
5. use of an adequate turbulence model.

Harsha considered two Gaussian plume models: the ADL model developed by Arthur D. Little Inc. and the UE model developed by University Engineers (Duffy et al. 1974). As mentioned above, also explicitly reviewed was the SAI model.

Harsha found that the predictions made using Gaussian plume models were too sensitive to the choices made for four key parameters: source strength, source type, dispersion coefficients, and weather conditions. In addition, he found that not only did these models not meet any of the five model requirements to some extent, but they were not satisfactory in some other requirements. In particular, the assumption of negligible vertical velocities was considered to be a major limitation to the use of the SAI model in near-field situations where vertical velocities are important. Such situations are common in flow around tanks and dikes but are generally not a problem in spills on flat ground or in water spills unless a ship or other structure is upwind or downwind of the spill.

Harsha's review of LNG dispersion problems indicated that one approach, described by Markham and Lewis (1975), is under development for an improved near-field model. This approach is also not an ideal approach for a general model, since it, too, does not completely address all five model requirements. An extensive discussion of the requirements for turbulence models and the details of some available turbulence models are presented in Harsha's review.

Harsha concluded that 1) an improved dispersion model is needed for near-field modeling where flow obstacles are present, 2) the "hydrostatic" model approach is valid for many less restrictive near-field and all far-field modeling studies, and 3) all models need improved turbulence models.

8.1.10 Recent Modeling Efforts

Several recent studies have been reported which were not available to Havens and Harsha. Wassel and Catton (1977) considered the diffusion from a line source in a stratified atmospheric surface layer including the effect of thermal stratification on turbulent diffusion in the atmosphere. Bloom and Mason (1978) have reported a model for both near- and far-field dispersion of dense gases. In the far-field, a Gaussian model is used, but in the near-field, the conservation equations of momentum, energy, and chemical species are applied simultaneously. Water droplet condensation, effects of chemical reactions and mixing with air are also considered. Of significance is the option of including downwind terrain details in the near-field model (Bloom and Mason 1978).

Van Ulden (1974) developed a model for the spreading of a heavy gas cloud (as contrasted to a neutrally buoyant gas) which was used to simulate experiments conducted with Freon-12. He found that the horizontal spread of the vapor was greater than expected and the vertical spread was much less than expected from other models. Dispersion coefficients were calculated from the model and experimental data for distances up to 3280 ft from the spill site. Ooms et al. (1974) describe a plume path model and compare it with experimental data. In particular, predictions of the plume path for a butane vent are presented. His model is based on the use of similarity profiles and conservation of energy. Reasonably good agreement is obtained between model predictions and experimental results. Sykes (1978) has reported an analysis of stratification effects in boundary layer flow around hills. Although his analysis does not specifically consider LPG vapor flow, his three-deck levels of atmospheric stratified flows appears applicable for LPG dispersion modeling.

One of the most recent theoretical vaporization studies (Opschoor 1981) concludes that in the case of a solid substrate (e.g. a spill on land) different dispersion results are obtained depending on whether the substrate is porous or nonporous or whether it is dry or moist. Experimental verification of the results is not yet available.

In a previous publication, the same author (Opschoor 1980) had presented a rather simple analysis of the evaporation and spreading of burning and nonburning LNG spilled on water. Heat radiated from the burning pool fire was calculated using empirical data from LNG fires on land.

8.1.11 Models Developed in the DOE Program

Research on LNG vapor generation and dispersion is a continuing emphasis of the DOE Liquefied Gaseous Fuels Safety and Environmental Control Assessment Program. Progress in the following studies conducted by the Lawrence Livermore Laboratory is reported in the two LGF Program Status Reports published to date.

A model for the unconfined spreading and evaporation of LNG when spilled on a water surface has been developed by Stein (1979). The model addresses the differential boil-off of LNG constituents and can be used to calculate effects for instantaneous or continuous spills. Such calculations have been shown to agree fairly well with data obtained in two LNG spill tests at China Lake.

Another study (Bowman 1979) conducted as part of the DOE Program, compares this and other LNG dispersion models in regard to their ability to predict the maximum distance to the lower flammability limit of a specified LNG spill. So far, the value of the study is constrained by the lack of sufficient test data for a definitive comparison.

In more recent studies, Chan et al. (1980) developed two computer models to simulate the vapor spread and dispersion from large LNG spills. The first model, based on a plain hydrostatic assumption for the vapor cloud, was found to be clearly less accurate than the second model, based on a nonhydrostatic formulation. The latter is now undergoing further development.

Most models developed so far neglect any moisture in the air/vapor mixture, even though such moisture can have a significant effect on the buoyancy of a vapor cloud. Therefore, another recent study (Haselman 1980) tried to evaluate such effects by computer analysis for mixtures of methane, air and water vapor. So far, there exist insufficient experimental data to verify the accuracy of the calculated results.

The effect of buoyancy on the dispersion of a vapor cloud was also the subject of an MIT study (Fay et al. 1980), which specifically tried to model negatively buoyant cloud motion and dispersion. The model is of intermediate complexity in that it assumes merely a hydrostatic pressure distribution, but includes air entrainment to account for vertical mixing. Since LPG clouds are mainly non-buoyant, this model may be of particular interest for estimating the spread of vapors from LPG spills.

8.1.12 Experimental Effort in the DOE Program

The simultaneous boiling and spreading of LPG have been investigated by Reid, et al. (1979, 1980) and Chang, et al. (1981) as part of the DOE Program. In this study, the boiling and spreading rates of LPG were found to be the same as those of pure propane. An LPG spill was characterized first by very rapid and violent boiling and then highly irregular ice formation on the water surface. The measured local boiloff rates of LPG agreed reasonably well with theoretical predictions from a moving boundary heat transfer model. The spreading velocity of an LPG spill was found to be constant and determined by the size of the distributor opening. The maximum spreading distance was found to be unaffected by the spilling rate. These observations can be explained by assuming that the ice formation on the water surface controls the spreading of LPG spills. While the mathematical model did not predict the spreading front adequately, it predicted the maximum spreading distance reasonably well.

LPG boiloff was investigated by Welker, et al. (1980) following rapid spills of less than 40 lb. Tests were run by spilling LPG onto concrete, polyurethane foam, and a sand-soil mix. Both the transient and steady state vaporization rates were measured. Good agreement was achieved between measured and predicted boiloff rates. The predicted rates were based on a film coefficient-limited model of heat transfer by conduction through the substrate with additional terms representing atmospheric convection, solar radiation, and sensible heat from evaporative cooling.

In a continuation of this work, Welker spilled propane on soil, concrete, insulating concrete, asphalt, wood, sod, and foamed polymers and measured boiloff during both the transient and steady-state boiloff periods. Most of the transient tests were run in 5-ft^2 circular pits.

Some of the tests were covered to exclude atmospheric effects. Simple one-dimensional heat transfer theory was used to determine the thermal conductivity of the substrate and the heat transfer coefficient between the substrate and the propane. Thermal conductivities were usually higher than the literature values for similar materials at ambient temperature. Heat transfer coefficients varied with substrate material, but were in the range expected to occur for propane. Mass transfer coefficients were determined for a range of variables and correlated as a function of the Reynolds number.

Vapor concentrations were measured downwind of propane pools 25 to 1600 ft^2 in area. The vapor concentrations along the plume centerline at ground level could be correlated satisfactorily with a Gaussian model. The vaporization rates for the dispersion tests were near steady-state values because the pits had been cooled for at least a half-hour before the vapor concentrations were determined. Pool temperatures were usually in the range of -70°F or less.

In a few tests, pressurized propane was discharged into the open air in an attempt to determine what fraction would vaporize or atomize and what fraction would collect as a pool. Even at rates up to 180 lb/min, no pooled propane could be collected.

Both 5-m^3 and 40-m^3 LNG spill tests have been conducted at the Naval Weapons Center, China Lake, California. Results of the 5-m^3 tests are reported in the Second Status Report of the LGF Program (Koopman et al. 1980). These experiments were successful in their primary purpose of evaluating the gas sensors and showed clearly the need for a large array of instruments to measure both gas concentration and the wind field. Data from the 5-m^3 spill tests also provided information about the dispersion process. Estimates of the dispersion coefficients were obtained by comparison with data from stations at the edge of the plume. The horizontal dispersion coefficient was found to correspond to stable atmospheric conditions as defined by the Pasquill-Gifford dispersion categories. The vertical dispersion coefficient was determined from gas concentration data and was also found to correspond to stable conditions. The vertical dispersion coefficient was found to be nearly the same in all four experiments despite considerable change in wind speed.

The LNG pool size was found to be sensitive to wind velocity. A significant result of these experiments was the first field verification of differential boiloff. The data indicate that, under certain wind conditions, part of the gas enriched in ethane and propane can propagate some distance downwind. Overall, these experiments indicated that a good description of the vapor source as a function of time is necessary to reproduce the downwind behavior of the vapor plume.

8.1.13 Wind Tunnel Modeling

Wind tunnel modeling is a valuable research and development tool for many complex flow situations since scaled experiments can often be easily and inexpensively conducted to gain both qualitative and quantitative knowledge of the flow field. A primary requirement is that the phenomena are sufficiently understood so that the proper dimensionless scaling parameters may be used. In a complex flow situation, such as near-field LPG dispersion near flow obstacles, wind tunnel models could provide valuable insight into the behavior of the vapor. Cermak (1975) and R&D Associates (1977) have provided reviews of meteorological wind tunnel modeling with applications to LNG vapor dispersion.

The use of substitute gases as tracers is common in wind tunnel modeling to obtain fluids of different density (from air) without heating or cooling requirements. Alternate fluids have been considered for LPG in some tests, for example, to avoid having flammable concentrations of vapors in the test area. Special gases and gas mixtures can be used to provide easy tracing capabilities as long as care is taken to properly scale all relevant parameters in a planned wind tunnel modeling test. Turbulence characteristics of the tunnel flow may prove to be a special problem for LPG dispersion modeling.

In the following various examples of wind tunnel investigations of plume behavior and vapor dispersion are briefly reviewed. It shows that basically the same techniques could be applied to the investigation of LPG spills.

Wind tunnel studies of dense stack gases were conducted by Bodurtha (1961) using a 3.5-ft by 7-ft wind tunnel at New York University. The stack gas was modeled with Freon-114/air mixtures covering specific gravities from 1 to 5.17. Graphs of the centerline path of the plume were prepared from photographic negatives. Correlations were developed for the initial plume rise and for rates of plume descent. Bodurtha, et al. (1973) later reported on wind tunnel simulations of the discharge of heavy gases from relief valves. This later work summarizes the results of the earlier work and discusses specific venting applications. Hall, et al. (1976) describe an experimental model for the escape of heavy gases, specifically propane or butane. They present a review of the literature and discuss the required experimental scaling of actual spills. They noted that dense plumes were very wide and shallow as compared with neutrally buoyant plumes, and that significant volumes of vapor can be gravity-driven upwind from the release point. Hilly slopes were observed to affect the plume motion significantly. Meroney, Cermak, and Neff (Meroney et al. 1976, Meroney and Neff 1977, Meroney et al. 1978a,b) have continued their original studies of dense gas (primarily LNG) dispersion in their wind tunnels. Their earlier work had concentrated on determining plume structure for simulated LNG releases. Later Meroney and Neff (1977) reported on the downwind dispersion of LNG vapor from diked tanks in scales of 1:666 and 1:130 and have provided more details of the methods and results of simulating LNG spills in high and low dikes (Meroney et al. 1978a,b).

8.2 LPG POOL FIRES

As a pool, LPG vaporizes rapidly, so that a combustible mixture is readily formed in the presence of air. If a pool of LPG is ignited, it burns similarly to a pool of gasoline or fuel oil. The result is a smoky, quasi-steady-state fire, which differs from transient vapor cloud flash fires. Research on fires from pools of LPG and other hydrocarbon fuels burning freely on either land or water is reviewed in this subsection.

Pool fires can damage nearby objects by heat which is transmitted by radiation and convection. Radiated heat is generally a major fraction of the total heat flux incident on targets, except possibly for targets very close to and downwind of a fire; heat transmitted by convection is rapidly attenuated with distance from a fire by the tendency of hot gases to rise and to mix with cooler air. The intensity of radiation on a target can be calculated either by using formulas based on experimental data obtained with pool fires, or by constructing a mathematical model of a pool fire from first principles. The accuracy of the approach based on existing data depends, of course, on the accuracy of the original data and on the extent of extrapolation which may be required. The accuracy of a mathematical model of pool fires designed to yield radiant intensities on targets will be limited by the compromises which must be made in order to obtain a working model of a very complex phenomenon. This review will begin with a general physical description of pool fires. This provides a basis for rather extensive descriptions and evaluations of mathematical modeling work and of experimental work, some of which can be made specifically applicable to the prediction of thermal flux on targets from LPG pool fires. The status of techniques for controlling the hazards of LPG pool fires by reduction of the thermal flux on targets or by complete extinction of the fires will also be reviewed.

8.2.1 Physical Description of Pool Fires

Freely burning pool fires of LPG, or of other petroleum fuels such as gasoline or fuel oil, have these major aspects:

- Vaporization of fuel from the liquid surface
- Mixing of the fuel vapor with air to produce flammable mixtures
- Combustion of the flammable mixture, releasing the heat of combustion of the fuel and yielding products of combustion having elevated temperatures, i.e., a "flame"
- Radiation and convection of energy from the flame to the surroundings
- Radiation and convection of energy from the flame back to the surface of the pool.

Each of these aspects involves numerous factors which must be recognized and accounted for in the design and execution of useful experiments or in the development of mathematical models for the quantitative prediction of potential hazards. The several aspects will be described below.

Vaporization of the Fuel

A pool of liquid fuel on land or on water will vaporize and feed a fire at a rate which depends on the total heat input rate to the liquid. In general, sources of energy for vaporizing the liquid will include heat input from the substrate (land or water) and heat input from the fire by radiation and by convection. The magnitudes of these heat inputs to the fuel will depend on temperature differences between the fuel and the substrate and between the hot gases of the flame and the fuel, and on the physical properties of the substrate, fuel, hot gases and air.

LPG is generally a mixture of propane and butane and vaporizes at its boiling temperatures which can range from -40°F to 36°F. The initial temperature of the vapor is usually the same as that of the liquid. However, the initial composition of the vapor will be richer in the lighter hydrocarbons than the liquid; this affects the fire because the heats of combustion of the components are different, and the products of combustion are different. In particular, while methane, for instance, burns with a "clean" flame, the heavier hydrocarbons produce significantly larger quantities of soot particles as their carbon to hydrogen ratio increases. Since the characteristics of the radiation from a flame depend partly on the concentration of soot in the hot gases, the thermal feedback mechanism, and thus the pool burning rate, will be affected by the time into the burning process.

Mixing of the Fuel Vapor with Air

Even before a fire exists above a pool, the fuel vapor will mix with ambient air to some degree due to local convection and diffusion, and flammable mixtures will exist in places above or near the pool. A fire produces hot gases which in turn induce strong convection because the hot, buoyant combustion products tend to rise and replacement air must be entrained. When the fire is established in a steady or quasi-steady-state, there will be a flux of air into the flame above the pool, even without wind. With wind, turbulence will tend to be greater, increasing the rate of mixing, and the transfer of fuel and air into the flame, particularly for smaller fires. However, thermal energy transferred into the pool may depend somewhat on the displacement of the flame relative to the pool by the wind.

Combustion of the Flammable Mixture

Combustion is a chemical reaction (or set of reactions) between the fuel and oxygen, as a result of which new molecules are produced, and energy is released. The visibility of the flame results from the high temperatures of the gases and particles (soot) throughout the flame. The visible flame above a pool extends essentially over the whole area of the pool under no wind or slight wind conditions, and extends vertically from a level slightly above the liquid to heights greater than the diameter of the pool.

Radiation and Convection of Energy from the Flame

Radiation is an electromagnetic mode of energy transfer and the radiated energy from a source can be characterized in terms of wavelength. The spectrum of wavelengths from a flame consists generally of two parts, a continuous spectrum due to the incandescent soot particles, and a discontinuous spectrum or spectral "lines" from the product gases. Radiation emanates from individual molecules or soot particles, and travels in straight lines until it strikes a target, or is absorbed along its path. Some of the radiation from the interior of a flame is absorbed within the flame, so that as flames increase in size, the radiant intensity from the "surface" of the flame approaches a limiting value. Total radiation traveling through the atmosphere from the flame to a target is partially absorbed, especially by water vapor (humidity) and carbon dioxide (CO_2).

Convection is a mode of transfer of thermal energy by the motion of gases or liquids. In a free-burning fire, the motion is caused by the hot burning or burned gases moving upward through the denser surroundings and by crosswinds. Thus, for these fires, convection is felt most strongly above the flame or fire, and quite weakly at all sides, except downwind. Even on the downwind side, however, the convection of heat to objects diminishes rapidly with distance from the flame, and in many practical situations convection of heat can be neglected with small error compared to the radiant flux.

Transfer of Energy Back to Pool

The same processes that are involved in transferring energy to external targets are involved in transferring energy back to the surface of the pool. Convection energy transfer may be fairly large because of the high degree of turbulence under the flame. On the other hand, radiant energy, while predominant in large pool fires, is reduced by absorption in the unreacted fuel moving upward from the fuel surface. Further, in deep pools, or for a spill on a water surface, some of the radiant energy may be transferred into the bulk of the fluid, rather than being confined near the surface.

8.2.2 Analytical Modeling Studies

Analytical studies of fires include those dealing with the prediction of reaction rates, velocities, composition, and temperature within the flame and those dealing with the prediction of the radiation from the flame. The "with wind" case is more complex than the "no wind" or symmetrical case. As a result, the "with wind" mathematical treatments tend to be less basic in their approach. It should also be clear that while an attempt could be made to confine the studies of experimental work for the most part to those involving LPG, most of the basic modeling problems do not inherently involve any specific fuel. Thus, many of the approaches developed for liquid fuels, or solid fuels, or gaseous fuels may be applicable to LPG flame calculations.

Modeling Under Conditions of "No Wind"

Mathematical analyses of buoyant and forced turbulent plumes with radiative heat exchange have been carried out by Murgai and Emmons (1960), Murgai (1962), Nielsen and Tao (1965), and Morton (1965). Murgai and Emmons treat the natural convection above fires in a dry-calm atmosphere with a constant lapse rate. They present curves from which the natural convection may be computed over a fire of arbitrary size in an atmosphere with an arbitrary lapse rate variation. This work is confined to the treatment of a fully turbulent convection column. Nielsen and Tao extend the plume analysis to include the variation with altitude of the composition, temperature, and velocity of the gases. The combustible substances at the base of the flame are assumed to undergo pyrolysis and to liberate combustible gases. The energy release due to combustion of these gases within the plume (controlled by the entrainment of air from the surrounding atmosphere) is considered along with energy losses due to radiation. The rate of release of fuel vapor from the source is specified as independent of the character of the fire. Thus, the treatment is actually for a turbulent diffusion flame and not a pool fire in which the radiation back to the pool affects the firing rate.

A useful treatment of turbulent diffusion flames and the strongly heated regions of fire plumes has been suggested by Morton. This approach is based on the theory developed for weakly buoyant plumes with appropriate modifications to allow for the high temperatures and large variations in density. A local entrainment function dependent upon the local mean plume to ambient density ratio, ρ/ρ_0, is suggested as,

$$E = (\rho/\rho_0)^{1/2} E_0 , \tag{1}$$

where E is the entrainment constant for the strongly heated plume, and E_0 is the well-established constant for weakly buoyant plumes. The conservation equations describing the strongly heated plume are shown to be reducible approximately to a form directly related to the set of equations used in the study of weakly buoyant forced plumes by the transformation

$$\rho^{1/2} a = \rho_0^{1/2} b , \tag{2}$$

where a and b are local length scales (essentially plume radii) for the strongly buoyant and weakly buoyant plumes, respectively.

Mathematical modeling of the conditions inside symmetric or "no wind" flames has been based on the principles of conservation of mass, species, momentum and energy over the flame cross section in conjunction with the necessary thermodynamic properties of the air/fuel mixture. Steward (1964, 1970) solved the steady axisymmetric flame problem by assuming constant properties over the flame cross section and the mixing behavior according to the Prandtl mixing length theory for turbulent jets. The model predicts the general characteristics of a free-burning fire in both the region where combustion is occurring and in the hot gas plume above the combustion section. The solution yields the mass flow rate as a function of height which can be directly related to the excess air as

$$\left(\frac{H}{Y_o}\right)_{\text{total}} = [C_1 + C_2 (1 - \omega)^{4/5} [(1 + E)^{3/5} - 1)] (N_{co}/C_3)^{1/5} \quad (3)$$

where,

H = Height of flame at given amount of excess air

Y_o = Jet radius at the flame source

ω = Inverse volumetric expansion ratio due to combustion

E = Mass of excess air/mass of stoichiometric air

C_1, C_2, C_3 = Numerical constants

N_{co} = Combustion number, proportional to V^2/gY_o (V = initial jet velocity and g = gravitational acceleration) multiplied by a constant depending upon the characteristics of the particular flame and burning process

The solution indicates that H/Y_o varies as the 1/5th power of N_{co}.

Comparisons with previously reported data in the literature (Putnam and Grinberg 1965, Putnam and Speich 1963, Blinov and Khudyakov 1957 and 1961, Thomas 1963) by Steward indicated that all data, when plotted as H/Y_o versus N_{co}, fall around the straight line,

$$\text{Log}_{10} \left(\frac{H}{Y_o}\right) = (1/5) \ \log_{10} N_{co} + 1.21, \quad (4)$$

and agrees with the 1/5th power variation predicted by the model. Numerical evaluation of the model gave close agreement with the flame height data, with all points falling near a curve representing 400% excess air.

Steward's model appears to be useful in the evaluation of temperature and concentration profiles in a naturally occurring fire. Such profiles can be used as a basis for the more important evaluation of the radiative flux density distribution around such a flame. Such information is needed to determine the radiative heat transfer to the fuel supporting the fire itself.

Wilcox (1975) approaches the problem through the use of the Karman-Pohlhausen integral method. The model is applied to turbulent diffusion flames having low initial momentum and nongray thermal radiation and was formulated and tested specifically for LNG pool fires.

For weakly or strongly buoyant plumes in which the initial momentum is small, the approach suggested by Morton (1965) leads to prediction of very little entrainment at the pool surface. However, experiments by Thomas et al. (1965), indicate vigorous entrainment occurs close to the fuel source. Wilcox included an ambient air entrainment factor which accounts for rapid fluid acceleration from initially low velocity at the liquid pool to much higher velocities established under buoyant rise of the combustion products.

Average flame conditions were assumed to apply over a given cross section. The mixture was assumed to be in thermodynamic equilibrium and to behave as a perfect gas. The radiation heat loss to the surroundings was obtained by assuming that the flame is optically thin. By using pure methane as the fuel and by assigning a constant temperature to the bottom of the flame, Wilcox computed the flame height based on a criterion established by Thomas et al. (1965). By plotting flame height to pool diameter against initial mass flux per unit area, Wilcox compares his calculated flame height-to-diameter ratio with Thomas' flame data and shows good agreement.

Wilcox assumes that the fire is optically thick in the direction of its symmetry axis and neglects axial radiation heat transfer. Radial-radiation heat transfer is computed with the exact transport equation with all gases represented as realistic nongray radiators. The model is used to calculate spectral distribution of the radiation heat flux. Comparison of the radial component with experiment data for a 6-ft diameter LNG pool fire (at about one diameter above the pool) indicates a reasonably good representation by the model.

Modeling Under Wind Conditions

With wind, the analyses have been carried out with the same set of equations governing the motion of jets or plumes as for the "no wind" case. However, the conservation of momentum is considered for both the vertical and horizontal directions and entrainment is considered to be influenced by the wind. Escudier (1972) integrated these equations by assuming that the plume loses heat to the surroundings as a radiating gray body for the case of burning gaseous hydrocarbon fuels. He obtained the plume height as a function of horizontal distance. For a wind velocity, v, up to v = 15 ft/s the near-field results can be represented in dimensionless form as

$$\frac{z}{l_m} = \left(\frac{R}{\alpha R+\beta}\right)^{1/2} \left(\frac{x}{l_m}\right)^{1/2} \quad \text{for } \frac{x}{d_o} \leq 1.0, \qquad (5)$$

and the far-field results in the form

$$\frac{z}{l_b} = \left(\frac{3}{2\beta^2}\right)^{1/2} \left(\frac{x}{l_b}\right)^{2/3} \quad \text{for } \frac{x}{d_o} \geq 10^4, \qquad (6)$$

where x/d_o is the ratio of the horizontal distance and the initial plume diameter, z is the vertical distance, $l_m = (k_o/\rho v^2)^{1/2}$ is the momentum length scale, k_o is the initial momentum flow rate, ρ is the air density, $R = l_m/b_o$, b_o is the initial plume radius, α and β are entrainment constants, $l_b = f_o/v^3$, and f_o is the initial buoyancy flow rate. For wind velocities higher than 15 ft/s, the ranges of x/d_o for the above equations become further restricted.

Brzustowski (1977) modified Escudier's results by considering that the chemical reaction goes through a pyrolysis process before oxidation. Again, no account was taken of radiation back to the source. He computed the plume centerline distances when the mass flow of the combustion products is constant and compared them with laboratory data using propane and hydrogen. Although the experimental trends of the variation of the plume height with crosswind are well represented by the theoretical prediction, good agreement occurs only in a limited range of data.

Modeling Considerations Relative to Thermal Radiation

A great amount of analytical work has been carried out relative to radiation from flames. Three different general approaches may be taken. In the first, the composition and temperature of the burning gases are computed independently, and then the Hottel Zone Method (Hottel and Sarofim 1967) is applied. After correcting the original fluid dynamic and combustion assumptions for the radiation results, the process must be repeated. In the second approach, a Monte-Carlo Method (Steward and Cannon 1971) is used to compute the radiation results, to provide a quicker (but less accurate) method to replace the Zone Method. The third method is known as the Flux Method. In this case, the radiation terms are approximated by diffusion-like terms, so they can be included directly into the flow and reaction equations. Fluxes may be considered in two or more directions (two flux method as used by Wilcox (1975) considers only radial radiation). Simple flux methods are well summarized and compared by Siddal (1972). A four-flux model with two additional second-order differential equations for axisymmetric coordinates has been developed at Imperial College and incorporated into the Spalding flow and mixing code (Gosman et al. 1969, Gosman and Lockwood 1973). An alternative four-flux model has been developed by Lowes, et al. (1973). It utilizes four first-order differential equations in which the axial and radial fluxes are coupled. This coupling is not present in the Imperial College four-flux model. More elaborate versions are also reported in the literature.

In all these methods, the assumption of a gray gas, that is, one in which the emissivity and absorptivity do not vary with wavelength, permits the simplest solution. However, the approach may be modified to account for nongray gas effects.

The properties of the major radiating gases in the combustion products are known. For instance, tables for H_2O and CO_2 are given by Hottel and Sarofim (1967). The absorption properties of the fuel gases such as will be present between the flame and the pool can also be found in the literature.

If the soot particulate size distribution and amount in the flame are known, the radiative output and absorption of that component can also be handled. Several examples of such calculations are reported in the literature (Beer and Howarth 1969, Taylor and Foster 1974, Felske and Tien 1973).

A major problem is calculating, a priori, the size spectrum and location of the developing and burning soot in a flame. Since the production mechanism is not understood at present, and the burning mechanism is only poorly understood even in a simple laminar premixed flame, let alone a large turbulent wind-blown diffusion flame, "soot formation is excluded" (from modeling turbulent diffusion flames) "because there is no suitable model" (to quote Lockwood and Naguib (1976)). Therefore, recourse must be had to experimental data to provide this component of an analytical approach to pool burning.

Modak (1977) recently proposed a slightly different course for computation of the radiation heat transfer from axisymmetric flame to the pool and to the surrounding horizontal surface. The flame is assumed to be an isothermal and homogeneous mixture of soot particles and gases in thermodynamic equilibrium. The local radiation properties of the flames are characterized by an effective radiation temperature and a gray absorption coefficient obtained from experimental data. A conical flame is matched in flame volume and diameter to the real flame. As a result, a closed form of solution is obtained to the heat flux problem. Orloff et al. (1979) give a related treatment for radiation from smoke layers in enclosures.

8.2.3 Experimental Studies of Pool Fires

The discussion of experimental pool fire studies is divided into three parts. First, the general requirements for physical modeling of pool fires are covered. This serves as a guide to correlating pool fire data, as well as indicating the ranges of fundamental variables of significance with respect to pool fire safety studies. Second, pertinent information from pool fire studies of various fuels, including LPG, are discussed under the following classifications: flame shape with no wind, flame shape with wind, burning rate, and flame radiation. Third, experimental studies of fire control and extinction are covered.

Ranges of Correlation of Pool Fire Data

When studying phenomena with many independent and dependent variables, it has been found useful to place the variables into groups which are dimensionless and of similar character. The phenomenon can then be described completely in terms of these groups. However, most complex phenomena require so many groups for a complete description that a physical model must be the same as the prototype. On the other hand, when the less significant groups are eliminated from consideration, physical modeling becomes more tractable and simple correlation of data may be possible. Following this approach, pool fires have been described with various degrees of success by three dimensionless groups. The fluid mechanics of pool fires are characterized by Reynolds numbers and Froude numbers.(a) A third dimensionless group is required to characterize the effects of crosswind velocity. These three dimensionless groups are briefly discussed below.

Reynolds Number

In free-burning hydrocarbon fires, fuel vapor rises from the surface with a low flow velocity (about 4 in/s or less). For accidental fires having cross-sectional areas of about 10 ft or more, the fuel vapor flow will be turbulent. From the work of Blinov and Khudyakov (1957, 1961) on liquid pool burning, one may deduce that turbulent flames from liquid pools have a minimum Reynolds number of about 5000, based on the cold vapor viscosity. Putnam and Speich (1963) suggested the same value from tests on natural gas diffusion flames. This increase of Reynolds number above the usually expected value of about 2000 probably results from the effect of increasing temperature as in a flame, causing an increase in the kinetic viscosity of gas. This Reynolds number limitation means that the results of many studies, such as those by Rasbash et al. (1966) and Corlett and Fu

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(a) The Reynolds number can be defined as the ratio of inertial forces in the flow to the viscous forces; for the fuel vapor in a pool fire, this becomes VD/ν, where V is the velocity of the vapor, D the diameter of the pool, and ν the kinematic viscosity. The Froude number is the ratio of inertial forces (or thrust) to the gravitational or buoyant forces, and can be expressed as V^2/gD, where g is the acceleration due to gravity.

(1966) have only qualitative value in modeling LPG pool fires. Likewise, some of the data reviewed by Brzustowski (1973) do not meet the Reynolds number criterion.

Froude Number

While some form of Froude number is often used in presenting data on free-burning fires (e.g., Byram and Nelson 1974), little information can be found on a Froude number value at which the flame shifts from thrust-controlled to buoyancy-controlled.(a) Knowledge of the Froude number at this transition point will be important in modeling LPG pool fires physically. Putnam and Speich (1963) on the basis of their experimental studies indicated that the Froude number should be less than 30,000 for modeling pool fire diffusion flames with natural gas spuds.(b) Ricou and Spalding (1961) presented a mathematical treatment of entrainment in burning jets through the range from thrust to buoyancy controlled; from their relation for the transition region, a critical value of Froude number, including a fuel dependent coefficient, may be determined.

Crosswind Effect

When a flame is distorted by a crosswind, the crosswind velocity, U, must be considered. Pipkin and Sliepcevich (1964), Welker et al. (1965), Welker and Sliepcevich (1966) in presenting flame data on the crosswind effect used a correlation group having the form of a Froude number, $\rho_a U^2/gD\ (\rho_a - \rho_f)$, where ρ_a, ρ_f are the densities of air and fuel, respectively, and g is the acceleration due to gravity. The appropriate Reynolds number for flame-bending studies is the one based on drag (i.e., wind velocity). These studies meet this turbulence criteria for the Reynolds number. Putnam (1965) in a study of buoyancy-controlled natural gas flames, used

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(a) In a buoyancy-controlled flame, the aspiration and mixing of air into the fuel plume is controlled by the heat liberation, and thus, by the buoyant forces resulting from the hot gases. In a thrust-controlled flame, the aspiration and mixing of air into the fuel jet is controlled by the momentum of the fuel jet, with no significant effect of buoyancy.

(b) Orifices or nozzles, essentially point sources.

the Froude number[a] in the form U^2/gL^*, where L* is the length of the unblown flame; unfortunately, this requires the use of a prediction equation for the unblown flame length to obtain a dimensionless group of independent variables. In that study Putnam observed a change in flame characteristics for windblown flames at Froude number values of about 0.05 to 0.1; below this value, the flame acted much like an unblown flame. Extending the results with some data from wood crib fires, there appeared to be a further change in windblown flame behavior at a Froude number of about 2, with a cessation of horizontal extension of the flame. Escudier (1972) and later Brzustowski (1977) set up equations for solutions to a bent over flame, but concentrated their attention on thrust-controlled flames rather than buoyancy controlled flames.

Flame Characteristics of Pool Fires

There are four dependent flame variables which may be of particular significance in studies of LPG spill fires: flame shape with no wind, flame shape with wind, burning rate, and flame radiation. Although all four variables have been discussed briefly in subsection 8.2.2, the emphasis in the following is on the extensive background information from studies of free burning fires of various fuels that may be pertinent to the understanding of LPG pool fires.

Flame Shape with No Wind

Blinov and Khudyakov (1957, 1961) presented some of the earliest data on pool fires. Using containers from 0.15 in to 75 ft in diameter, they examined a variety of flammable materials including diesel fuel, fuel oil, gasoline, and tractor kerosene. These data showed a gradual decrease in the ratio of flame height to pan diameter as the pan diameter increased. However, only one datum, from a 75-ft pool diameter for gasoline, was above an 8.2-ft diameter pool. This datum showed no further effect of increased size. This datum was incorrectly interpreted as indicating no effect of size on the ratio of flame height to pan diameter when the flame becomes fully turbulent.

(a) Note that the Froude number here is based on the crosswind velocity, and not an upward velocity as normally used in a Froude number definition.

The first suggestion of a correlating equation between flame height and burning rate appears to be that of Thomas (1963). He proposed a correlation for wood crib fires in essentially the form

$$L/D = C_1 \, [V/(gD)^{1/2}]^{0.61} , \quad (7)$$

where L is the flame height, D is the pool diameter, V is the velocity of the fuel, g is the acceleration due to gravity, and C_1 is an experimentally determined coefficient. The relation has been used widely for correlating pool fire data, assuming a cylindrical shape flame. For a natural gas flame from a spud, Putnam and Speich (1963) suggested the form

$$L/D = C_2 \, [V/(gD)^{1/2}]^{2/5} , \quad (8)$$

where D is now the diameter of the spud opening, and C_2 is an experimentally determined coefficient. Thomas showed that data for each of the relations fell on a continuous curve, that is, L/D = function $[V/(gD)^{1/2}]$. Thus, the dimensionless group in brackets on the right appears acceptable for determining flame length. Putnam and Speich (1963) further showed that natural gas fires consisting of a perimeter of sufficiently close fuel spud fires ("string of pearls") acted the same as a uniformly spaced array of spud fires of the same total firing rate, and also that these fire data agree in form with the data of Thomas (1963). Finally, Putnam (1963) showed that a 2/3 power for an area source (composed of a ring of point sources) was derivable from a 2/5 power for a single point source. This result indicates that even Thomas' results (0.61 power, Equation 7) were obtained for conditions where area is not completely controlling, i.e., 0.61 is less than 0.67.

Putnam and Grinberg (1965) showed for various ambient temperature mixtures of CH_4 (methane), C_2H_2 (ethylene), and C_3H_8 (propane) diluted with various amounts of air, H_2, or O_2, that turbulent diffusion flame data could be correlated with temperature, but that visible flame lengths for the various mixtures could not be correlated. Also, Sunavala (1967) reviewed the work of several authors on the effect of fuel properties (such as density and stoichiometric mass concentration) on flame height. For turbulent, buoyancy-controlled flames, he suggested a fuel dependent form of coefficient for Equation 8 above, namely, $C_2 = 5\ (C_T(1 - R)^{-2/5}$, where C_T is the mass concentration of fuel at the stoichiometric ratio and R is the density ratio of stoichiometric products to fuel; for some data conforming approximately to Equation 7 above, he suggested using an "effective" diameter in order to retain the form of Equation 8.

Steward (1970) managed to fit a wide range of data on flame height, including the gasoline data of Blinov and Khudyakov (1957, 1961) and some gaseous propane data to the form of Equation 8. Raj (1977) indicated the correlating relation (correcting a typographical error) was

$$L/D = 14.7 \left[\frac{\omega(r + \omega/\rho_0')^{\,2}}{(1 - \omega)^5} \right]^{1/5} \left[\dot{m}''/\rho_a \sqrt{gD} \right]^{2/5} \qquad (9)$$

where r is the stoichiometric air-to-fuel ratio, ω is the inverse volumetric expanse ratio of products of stoichiometric combustion, ρ_a is the air density, ρ_0' is the ratio of fuel vapor density to air density, and $\dot{m}''$ is the mass burning rate of fuel per unit area.

Flame Shape with Wind

When a flame is bent over by wind, the prediction of flame shape becomes more complex. Welker and Sliepcevich (1966) derived a quasi-theoretical equation for the angle of deflection of the flame from vertical by the wind, by treating the flame as a rigid cylinder with length and diameter the same as for the unblown flame. Experimental data were used to determine the coefficients in the equation, including a term for the crosswind Froude number, $Fr = U^2/Dg$, and the crosswind Reynolds number, $Re = UD\rho_a/\mu$, where U is the crosswind velocity, D is the pool or flame diameter, ρ_a is density of the air, and μ is absolute viscosity of the air. The result was

$$\sin\theta/\cos^2\theta = 3.2\ Re^{0.07}\ Fr^{0.7}\ (\rho_g/\rho_a)^{-0.6}\ , \qquad (10)$$

where θ is the angle between the axis of the flame and the vertical, and ρ_g is the density of the fuel gas. This result has been used to compute radiation from flames by many authors, including Welker (Duffy et al. 1974) and Atallah and Raj (Duffy et al. 1974). However, Equation 10 was obtained for flames so small that in the unblown condition they would have had to be nonturbulent. Welker and Sliepcevich (1966) also obtained an empirical relation for the amount of flame trailing; their sketch shows that the flame angle measurements were made from the center of the entire flame base. The flame trailing distance correction is expected to be generally less than one flame diameter; in using the flame angle equation, this distance correction appears to be universally neglected.

Atallah and Raj (Duffy et al. 1974) consider two formulas for the angle of the wind-deflected flame. The first formula was developed by Welker and Sliepcevich and was discussed above (Equation 10). The second is a modified form of a development by Thomas for two dimensional wood crib fires. They determined that the measured flame angle for the LNG studies (Duffy et al. 1974) could be correlated by the relations

$$\cos\theta = 1/\sqrt{u^*} \qquad \text{for } u^* > 1.0\ , \tag{11}$$

$$\cos\theta = 1.0 \qquad \text{for } u^* \leq 1.0\ , \tag{12}$$

where $u^* = v/u_c$, v is the wind velocity, $u_c = \frac{\dot{m}''\ gD}{P_v}^{1/3}$ is the characteristic velocity, D is the dike diameter, P_v is the vapor density at the saturation condition at normal boiling point, $\dot{m}''$ is the burning rate in mass units per unit area, and g is the gravitational acceleration. The measured visible flame length L was correlated with the relations

$$\frac{L}{D} = \left(\frac{u_c}{\sqrt{gD}}\right)^{-0.54} \left(\frac{\rho_v}{\rho}\right)^{-0.19} u^{*0.06} \qquad \text{for } u^* > 1.0, \tag{13}$$

$$\frac{L}{D} = \left(\frac{u_c}{\sqrt{gD}}\right)^{-0.54} \left(\frac{\rho_v}{\rho}\right)^{-0.19} \qquad \text{for } u^* \leq 1.0, \tag{14}$$

where ρ is the ambient air density. It will be noted that for no-wind conditions, this indicates a flame length varying with $D^{1.09}$ rather than $D^{0.8}$ of Equation 8. No explanation is given, but the result may be an effect of too narrow a range of experimental data. In any case, Atallah and Raj found the flame angle prediction by either Equation 10, or Equations 11 and 12 were equally good compared to the LNG flame data taken at San Clemente.

Putnam (1965) in an experimental study of natural gas flames for single spuds and for arrays of spuds, correlated the flame data on the bases of the downwind extension of the flame, L_{sh}, and the height of the flame, L_{sv}, both related to the unblown flame height. In deriving correlation equations, the flame height with wind was determined on the basis of the additional aspiration of air by the buoyant flame because of the crosswind. The downstream (horizontal) distance to the peak of the flame was determined from the transit time of gas from the flame base to the flame peak and the crosswind velocity. The Froude number, $Fr = U^2/gL^*$, was also defined in terms of wind velocity and of the unblown flame height, L^*. For Froude numbers up to 0.05, the wind effect on flame height appeared negligible. Insufficient data were available to reach a definite conclusion, but the dimensionless downstream extension appeared to vary with $Fr\ \alpha$ where $0.5<\alpha<0.8$. For higher values of Fr,

$$L_{sv}/L^* = 0.48\ (U^2/gL^*)^{-1/4} \text{ and} \tag{15}$$

$$L_{sh}/L^* = 0.60\ (U^2/gL^*)^{1/6}. \tag{16}$$

For $U^2/gL^* > 2$, the flame ceased to lengthen. Blown flame shape was slightly dependent on the ratio of flame base diameter to unblown flame height; however, insufficient data were available to give a correlation function. This correction may be significant for the flames of LPG pool fires. It should be noted that there is a close similarity between Equation 11 and 12 and the above results.

This model of wind effect on flames assumes a constant firing rate, as is usually the condition in a gaseous fuel system (Putnam 1965). In a pool fire, the rate of fuel consumption will be affected by the angle of the flame and also by any pulling away of the flame on the upwind side. This is indicated strongly by data on burning pools of diesel fuel and tractor kerosene (Blinov and Khudayakov 1957 and 1961, Hall 1973). The burning rate increased by a factor of up to 3-1/2 (diesel fuel), first linearly and then leveling off before the flame blew off at about 25 miles/hr. However, even the larger pool was less than 19 in. in diameter. On the other hand, Burgess and Hertzberg (1974) do not analyze for specific effect of wind velocity but do show the effect is small from benzene in trays over 3 ft in diameter. It would appear that since there is a thermal feedback mechanism affecting vaporization, and the amount of vaporization controls the flame size, and the thermal feedback must be affected by the flame tilt and trailing, there must be some net effect of wind on the vaporization rate. Thus a pool would not be expected to burn at the same rate in wind and no-wind conditions. However, it should be noted that the effect of wind on flammable liquid burning rates might be different from the effect on liquefied gases.

Following the above reasoning, and applying a flame tilt model to a pool flame, a feedback mechanism must be added in order to compute the value of unblown flame height at the pertinent fuel consumption rate. In the Welker-Sliepcevich (1966) model, a rigid bent-over flame is assumed, so that the length of the flame does not enter into the angle computation. To the simple theory, Welker-Sliepcevich added an empirical correction factor which included an effect of Reynolds number and an effect of a density ratio, in order to obtain agreement between theory and experimental data. In order to compare the Putnam model with the Welker and Sliepcevich model, the Putnam relation for the point-flame height should be corrected to the area source form, and a half-diameter added to the computed downstream extension of the flame. Also, Putnam did not analyze the effect of fuel temperature, fuel density, and stoichiometric ratio, but rather used an experimentally determined coefficient for natural gas at ambient temperatures.

The largest single body of data on pool fires was generated on LNG pool fires in the AGA IS-3-1 program (Duffy et al. 1974). The analysis s of these data indicates that the angles predicted by the Welker-Sliepcevich method are reasonable. This seems to indicate that the air turbulence may have overcome the laminar flow condition of the unblown flame in the small-scale experiments by Welker and Sliepcevich.

The aim of the above discussion on flame tilt resulting from cross-wind is to provide a basis for predicting radiation output. The shape of the flame should be considered further, however. Byram and Nelson (1974) and Steward (1970), among others, indicate a commonly observed necking down of the pool fire followed by an expansion as the flame moves upward. A more serious question arises in connection with the often observed periodicity of the unblown flame, from very small pools (Rasbash et al. 1966) to very large pools, and for both gaseous fuels and liquid fuels. Hertzberg (1973) discusses the phenomenon of periodicity to some extent, but proceeds to use steady-state assumptions in his analysis of the comparative effects of radiation and convection on the feedback of energy into the pool. Byram (1966) and Byram and Nelson (1970) consider periodicity from a scaling point of view, but no literature is available on the effect of large amplitude oscillations on the overall burning rate of a pool fire, or on the radiation output. Byram and Nelson (1970) present some data on the value of the dimensionless frequency, $f\sqrt{D/g}$, for ethanol flames from 3 in. to 8 ft in diameter, and indicate a value of $f\sqrt{D/g}$ of about 1/2. The small range of data and the scatter preclude extrapolation of the results. Furthermore, no information is given by Byram and Nelson on the amplitude of the oscillations, which must be involved in any judgment of how important the factor may be. Hertzberg et al. (1977), present an explanation for the observed buoyant flame periodicity, but the observation of similar phenomena in water flow modeling of such flames (Putnam 1961) casts doubts on the validity of the explanation. It is clear that the oscillations, which produce separate, buoyant elements of flame, result in difficulties in determining a characteristic flame height.

Some researchers think that when a sufficient crosswind is present, the periodicity disappears, but this is not true in a broad sense. A photograph in a TRW report (Mullin et al. 1969) shows tongues of flame at angles exceeding 45 degrees from the vertical. That report also comments on periodic rolling vortices in the dispersing LNG vapors in a crosswind (with no flame); this again indicates that the periodicity is basically a fluid dynamic phenomenon, rather than being intrinsically related to combustion. However, periodicity is usually only observed in the presence of a fire.

One can also speculate that if a flame is of sufficiently large size, the height to diameter ratio will be so small that with a wind the flame will break up into an array of smaller flames and, especially with a gradient wind, fire whirls may also appear.

These observations suggest that extremely detailed steady-state mathematical attacks on the problem of predicting flame characteristics may be of limited usefulness.

Burning Rate

The burning rate may be treated as a dependent or an independent variable. When gaseous fuel jets are used to model flames, the burning rate is an independent variable. When the burning rate of a liquid pool is measured, it can be treated as an independent variable to determine the flame size and radiation output. However, for a liquid pool fire, the burning rate is affected by a feedback mechanism involving back radiation from the flame and convection produced by the flame (both of which will vary with flame size, crosswind velocity, and fuel), and conduction from the undersurface of the pool.

For the no-wind case, several sets of data are available for various fuels which contribute to an understanding of the feedback phenomenon. The theory is not well advanced; many of the nonwind-burning models involve an assumption that radiation is directed only horizontally. The computation of the radiant flux to the pool surface cannot be done until after the computation of the flame shape. Computation of the convection flux, if this is important in large fires, is even less adequately treated in literature.

The experimental data on this subject appear to fall into two groups, namely, data related to the recession rate or rate of consumption of the fuel, and data related more to the heat input resulting in the fuel consumption. The two types of data will now be discussed in turn.

Blinov and Khudyakov (1957, 1961) presented some of the earliest data on pool fires, using containers from 0.15 in. to 75 ft in diameter, and examining fuels including diesel fuel, fuel oil, gasoline, and tractor kerosene. These data have been extensively analyzed by several investigators (Hall 1973, Byram 1966, Hottell 1959). The data showed a decrease in burning velocity at very small sizes of pools, followed by an increase at larger sizes and an apparent leveling off as the cold flow Reynolds number increased above about 5000.

Burgess and Hertzberg (1974) and Burgess et al. (1961) reported data on liquid recession rates for pools up to 5-ft diameter, and radiation output for flames up to 2.5-ft diameter. This was followed by a succession of tests at Bruceton and Lake Charles (Burgess and Zabetakis 1962, Conch Methane Services, Inc.) on pools of LNG as large as 20-ft square, plus data on other fuels. The authors observed that burning rate (volume per unit area per unit time) increases with size of pool, but indicated that this tends to level off for pool sizes greater than 5-ft diameter. Furthermore, some crosswind tests on the various pools of benzene indicated the burning rates were all the same, at the rate for the 5-ft pool.

Burgess and Hertzberg (1974) then suggested that the burning rate of a variety of fuels from an infinite area (extrapolated from finite area data) is linearly proportional to the dimensionless group

$$G = \Delta H_C / \Delta H_V \qquad (17)$$

where ΔH_C is the lower heating value of the fuel and ΔH_V, which they call the "total sensible enthalpy", is the enthalpy difference between saturated fuel vapor at the interface temperature and the fuel in the reservoir state (Hertzberg 1973, Spalding 1964). The specific equation suggested was

$$R_{(\text{mm/min})} = .076\ G. \tag{18}$$

A major deviation was observed for hydrogen, which fell below the linear relation at the high end. Several other discussions of the original data from Burgess et al. (1961), Burgess and Zebetakis (1962), Spalding (1964), suggest explanations of the correlation and variations on it, such as the use of Spalding's B factor or transfer number for droplets (Hall 1973, Spalding 1964, Glassman and Hansen 1968, Berl 1964). These discussions all imply the need to include the liquid fuel density in the recession rate equation, and some treatments imply the need to include the mass ratio of fuel to air at the point of stoichiometry (this term is in the Spalding B factor). However, the correlations show less scatter without these terms. An excellent correlation is obtained by Putnam (1978).

$$R_{(\text{mm/min})} = 0.212\ G^{3/4}\ , \tag{19}$$

wherein the 3/4 power is that discussed by Spalding (1964) and Glassman and Hansen (1968), but they did not note the relation.(a) Using this relation, the recession rate for propane becomes 0.28 in./min; by comparison, butane has an observed value of 0.31 in./min.

For mixed fuels, such as aviation gasoline (Zabetakis and Burgess, 1961), kerosene (Hall 1973), and diesel fuel (Hall 1973), as contrasted to pure fuels, the value of recession rate is lowered. However, there is a problem here as to how to interpret the data. As one extreme, batch distillation could occur. In this case, the pool composition and temperature, and the vaporized gas composition vary as the pool burns. At the other extreme, especially for highly viscous fuels, equilibrium flash vaporization may occur from the surface with a uniformity of pool composition and temperature, and vapor composition as the pool burns. A pool of water under a thin layer of fuel, however, could permit the boiling of the water, mixing in the fuel layer, and a trend toward batch distillation. (Note that the water vapor can also suppress the soot formation and thus change the flame radiation properties.)

(a) The computed values of G agree with those of Zabetakis and Burgess (1961) but not of Burgess et al. (1961) and later references.

While the analysis of the Bureau of Mines data was in a consistent form, and the values for different fuels can be compared on the basis they used, there is considerable question on whether the data on burning rate level off at the value they assumed. As an example, Burgess et al. (1961) reports 0.25 in./min for LNG with their largest pool, 5 ft. However, the Lake Charles tests on LNG, hexane, and gasoline, show that the rate continues to increase with diameter with values of 0.39 to 0.47 in./min for the 20-ft square pool (Couch Methane Services, Inc.). The vaporization rate for the San Clemente tests was about 0.51 to 0.63 in./min for 20-ft diameter pools.

Concerning the aspect of heat input to the liquid pool from the flame, there are a number of pertinent data. Thomas et al. (1965), measured the radiative and convective heat flux to a 3-ft pool of ethyl alcohol. The radiant intensity fell off toward the boundary, but the convective flux increased. At the center, the heat flux to the pool was 80% radiant and 20% convective. The total heat flux was about 44% above that required to vaporize the fuel, but the radiative heat flux alone almost exactly equaled that required by the vaporization rate. Yumoto (1971), in a short note covering hexane and gasoline burning in pools of 2-, 5-, and 10-ft diameter, showed that the ratio of radiation to convection levels off at the largest size. For hexane, comparing measured fluxes for the 10-ft diameter pool to that computed as required for vaporization, 70% was from radiation and 23% from convection. For gasoline for the 10-ft diameter pool, the values were 61% from radiation and 34% from convection. Alger et al. (1976), conducted seven tests with 10-ft diameter pools of methanol and JP-5 at wind velocities less than 6.6 ft/s. The radiant heat intensity at the center of the pool for methanol was 1.93 times greater than that required for evaporation; for JP5, it was 1.42 times greater. The same researchers compiled data on the fall off of radiant intensity toward the pool edges, also noted by Yumoto. The ratio for the average intensity was 1.50 for methanol. This result seems to be consistent with the results of Thomas et al. (1965). The lack of reasonable energy balance in much of the above discussed data indicates problems in interpretation of the data or understanding of the phenomenon. For instance, reflection of radiation by the pool surface is never considered in these models.

Wood et al. (1971), used a 5-ft sand pool burning methanol and acetone in a model study of large fires. Radiometers measured the radiative flux to the pool surface. Convective flux was determined by the difference between the total flux required for vaporization and the measured radiant flux. In these studies, radiation was found to contribute less than 60% of the total flux required to account for the vaporization rate, until the fuel consumption fell off and the flame broke up into flamelets. From the relative flux measurement, it would appear that a sand wick pool fire behaves somewhat differently than a liquid pool fire, and thus would provide a questionable model for LPG spill studies.

Corlett (1970) used a porous stainless steel disk fed with a gaseous fuel to simulate a pool fire and to study radiative and convective transfer to the pool. However, the flames were in the nonturbulent regime.

Masliyah and Steward (1969) used their "top-hat"(a) flame model to compute radiant flux to various sinks, including the base of the flame. Included in their treatment is a term for the "degree of darkness" resulting from carbon particulate concentration. Using this model, they found a reasonable fit to the burning rate data of Blinov and Khudyakov (1957, 1961).

Duffy, et al. (1974), analyzed the LNG data from the San Clemente diameter tests for the four wide angle radiometers in the 6-ft and 20-ft pools. All these flames were windblown. As a result of the crosswind, the radiant intensity to the pool on the upwind side of the pool and on the sides of the pool was considerably below that at the center and the downwind side. For the 6-ft diameter pool, the vaporization rate three minutes after ignition averaged 0.38 in./min; for the 20-ft pool, the rate averaged 0.56 in./min. The average radiation input in each case was close to the input required for vaporization. Noting the wind direction and the shape of the flame relative to the radiometer readings and the soil temperatures, it was calculated that the vaporization from the 6-ft pool was consistent with no radiation from the flame on 40% of the area and with maximum radiation over 60% of the area.

(a) "Top-hat" indicates a profile that has a constant finite value near the axis, dropping to zero at some finite radius. In some treatments, composition, temperature, and velocity are all assumed as this shape; in others, only part may be so assumed.

Corresponding values for the 20-ft pool were 20% and 80%. By taking a weighted average of evaporation rates at 3 minutes for the areas without radiant flux and with flux, reasonable agreement was obtained with measured fuel recession rate.

For pool fires on water, the heat flux from the water is expected to remain high, rather than falling off rapidly with time as for land spills. In addition, there is evidence that water, either from boiling under a thin film of fuel or from the sweeping inflow of air, or both, suppresses soot formation and thus increases the transmission of radiant energy. With the radiation input, the total vaporization rate could be considerably greater than on land, and thus the flame size would be greater. The increase in flame size would further increase radiation to the surface and increase the burning rate even more. This might account for the observation at the recent 1500-gallon LNG pool fires on water at China Lake where the height of the flames was 3.5 to 4.5 times the base diameter (Raj 1977).

Duffy et al. (1974), also observed a relation between the total vaporization rate and the radiation flux to a vertical surface 7-1/2 diameters downwind from the pool: with little scatter, the required vaporization energy was 42.7 times the radiant heat intensity at this distance, independent of pool size and wind velocity. There do not appear to be any other data on the feedback radiation to large pools of fuel, though there are data and analyses of the energy feedback to small pools (Andersen et al. 1969) and to simulated pools (Corlett 1970).

Flame Radiation

The radiation from a flame is generally treated as a dependent variable, but with some mathematical models an estimate must first be made of the ratio of radiant energy loss to total chemical energy input in order to determine flame size. Burgess and Zebetakis (1962) found for a 1.3-ft diameter butane burner that 30% of the total thermal energy was released as radiation, and for a 2.5-ft diameter pool fire, 27% of the total thermal energy was released as radiation. The highest values listed were for benzene, with 36% for a 45-ft pool. They suggested, from these data, a value of 25% for large LNG pools. These values were based on the assumption that the flame radiated uniformly in all directions (i.e., over a sphere) from a point source (Burgess and Hertzberg 1974).

Markstein (1977) reported that the total radiation from a turbulent, buoyant, gaseous propane flame was constant at about 25% of the energy released for a range of firing rates of from over 30 to 1. The radiant output was found by summing the horizontal flux over the height of the flame. Several studies have been made of LNG radiation, and comparable values for different treatments should give an indication of the variation of results to be expected with LPG. The value suggested by Burgess and Zabetakis (1962) of 25% radiant energy has already been mentioned. Atallah and Raj (Duffy et al. 1974) report values in the range of 20% and 25% for their analyses of the San Clemente experiments, assuming cylindrical flames for 6-ft and 20-ft pool fires, and assuming the surface flux rate used in their overall data correlation. However, May and McQueen (1973) report a value of about 16% from studies of the radiation of a large, steadily burning pool of LNG (2.3×10^9 to 6.8×10^9 Btu/hr total heat release rate); that value was based on the assumption that the flame radiated uniformly in all directions from a point source. The AGA tests at San Clemente (Duffy et al. 1974), analyzed on the basis of the same assumption, showed 11% of input radiated for 3.4×10^7 Btu/hr total heat input rate and 12% for 5.5×10^8 Btu/hr. A clarification of these results is needed. However, as pointed out by Raj (1977), the assumption of a point radiation source for a flame at distances less than about 5 diameters from the flame axis can lead to large discrepancies.

Markstein (1975, 1976, 1977), in a series of three articles, studied the radiation from several hydrocarbon flames. His first tests (Markstein 1975) involved a row of 1 to 10 laminar flames from flat burners 2 in. x 0.13 in. spaced at 1 in. The smokier fuels, in increasing order of sooting tendency, were propylene, iso-butylene, and 1,3-butadiene. Their radiance was described adequately by a single gray gas assumption. Extrapolating to infinite thickness, the ratios of the black body flame temperature to maximum adiabatic flame temperature were 0.637, 0.607, and 0.567, respectively. The reversal of the trend from the single, thin flame case was ascribed to increasingly incomplete combustion and to radiation energy loss. The data for the remaining fuels were fit adequately by a two gray gas hypothesis. These gases showed the soot was present in a thin layer slightly on the fuel-rich side of the high temperature region. The larger component of the two gray gas equations was associated with the soot layer. For an infinite number of flames, methane had a temperature ratio associated with the soot component of 0.64, and the values for ethane, propane, n-butane, iso-butane, and ethylene ranged from 0.70 to 0.75 with no statistical significance in the difference. The work was extended to turbulent ethane and propane flames, at a spacing of 1 ft to prevent flame interaction. The temperature ratios decreased from 0.755 to 0.702 for ethane and from 0.707 to 0.686 for propane. The distribution between the two gray gas components also changed. This showed that the turbulent flame could not be considered as an array of laminar flames.

In addition, Markstein (1976) investigated the ratio of radiant energy to thermal energy input based on the lower heating value of the fuel for four different turbulent propane flames, over a range of firing rates. Values of 0.206, 0.204, 0.213, and 0.246 were obtained. Markstein (1977) obtained a value of 0.238 for what appeared to be the fuel nozzle which gave above 0.213 value. The range of Reynolds number was about 1000 to 9000 and the maximum heat release rate corresponded to about a normal home furnace rate, 1.0 x 10^5 Btu/hr. However, it was indicated that the linearity between firing rate and radiation had been confirmed up to 3.9 times this value of about 4.1 x 10^5 Btu/hr.

The main point of Markstein's (1977) study was to determine the radiative power per unit height as a function of height. Integrating the bell shaped curve that was obtained gives a coefficient of 0.264 as compared to 0.238 determined by a wide-angle radiometer. The discrepancy was not definitely explained, but the effect of the CO_2 and H_2O absorption in the atmosphere at the greater measurement distance for the overall measurement was suggested as one possibility. In this study, dimensionless relations are given for the flame shape, based on the test results. Two propane flames were brought close together, and it was found there was an increase in radiation output due to flow interaction, reaching a maximum of 10% and then falling off again.

Becker and Liang (1981) studied the emission of soot and thermal radiation by free turbulent diffusion flames of CO, H_2, CH_4, C_2H_6, C_3H_8, C_2H_4, and C_2H_2 over a range of variables going from forced convection to natural convection. For their data and that of several other investigators they found the soot yield factor for CH_4, C_2H_2, C_3H_8 leveled off at high value above a Richardson number(a) of about 2000. For propane, the mass of soot was 0.45 percent of the mass of carbon in the fuel. The fraction of chemical energy radiated was 0.27. While all these data were for flames from spuds, the critical Richardson number can be converted into a size of pool fire, which turns out to be about 3.3 ft. This would imply that for larger LPG pool fires, the soot production would be a constant fraction of the fuel input.

Prediction of Radiant Flux

The background developed above leads to the discussion of radiant flux prediction. Various predictions of pool fire radiation are available. For instance, for a body surrounded by a hydrocarbon fire, it is common to design for 34,500 Btu/ft^2, which corresponds to a black body temperature of about 1650°F hr. The value was based on many years of observation and various types of experiments. Knowing the surface area of a flame, and the flame shape relative to the receiver, one can then calculate the heat flux to the receiver. A second approach is to assume a fractional radiant heat radiated from the flame, and then distribute this output uniformly over the flame surface. It should be noted that in both cases, a further assumption is required to predict the burning rate, or more fundamentally, the energy feedback to the pool. To get around this problem, in the latter case, a burning rate is assumed; in the former case, a relation of flame area to flame diameter, based on burning rate, is assumed.

The next level of approximation is to realize that the smaller flames are not black. Available data are then analyzed on the basis of a beam length proportional to the pool size.

(a) Richardson number is defined as $\pi g \rho_\infty L^3/4 \dot{G}_o$, when L is the flame length and $\dot{G}_o$ is the momentum flux of the fuel. For a specific fuel and a large pool where L/D is about constant, Richardson's number is inversely proportional to the Froude number.

Atallah and Raj (1975) and Welker (Duffy et al. 1974) used this approach for windblown LNG flames on land, using the same data source. Raj and Kalelkar (1973) show how the computational routine for variation prediction can be extended to cover the cases of spills spreading on water.

The final level of complication in data analysis and mathematical modeling that will be mentioned here is to determine a flame emissivity as a function of beam length; this is quite common when using a "top-hat" property profile (Masliyah and Steward 1969). For cases where luminous radiation is unimportant, this can be easily accomplished using, for instance, Hottel and Sarofim's (1967) curves based on the partial pressure and temperatures of the radiating gases. When luminous radiation is important, as is the case for most large flames, some method of estimating the soot concentrations and distribution in the flame must be used. The emissivity can then be computed for any path through a known medium at a known temperature (Felske and Tien 1973, Taylor and Foster 1975, Beer 1974).

To indicate the complications that result in applying theory to the real world data, it has been noted that when the black cloud around higher hydrocarbon flame temporarily clears as hot interior gas breaks through, there is a brief increase in flame luminosity.

Data on LNG fires have received the most complete analysis of any fuel data; for large flames, the radiation output on a percent energy input appears not too different from that of LPG.

Models developed on the AGA IS-3-1 program for the LNG turbulent diffusion flames are essentially empirical correlations based on work for other fuels (Duffy et al 1974). Two groups using the same data derived similar models differing somewhat in the correlations for flame length and tilt, but yielding roughly the same intensities on targets outside the flame. The data are difficult to interpret because of the natural turbulence, swirl, and varying geometry of the flames due to wind effects, and also because of noise in the recorded data. In addition, the radiation data for the 80-ft diameter fires are of limited value because the recording system failed before the radiometer indications had reached peak values. All of the data taken together indicate that the flame surface flux for 80-ft fires is probably very close to the maximum flux postulated for an infinitely large pool fire.

The data correlated on the AGA IS-3-1 program were obtained in the earlier periods of fires before the development of heavy soot due to the increasing richness of the fuel in heavier components (Duffy et al. 1974). The spectral data indicate that the radiation intensities on targets outside the flame, and attributable to soot, increased with pool diameter. However, the data do not permit a definite conclusion as to whether the total external flux is greater or lesser at later times, when the burning of heavy components would become important if the pool were sufficiently deep.

Data published by the Bureau of Mines (Burgess and Zebetakis 1962), Tokyo Gas Co. Ltd. (1971), British Gas Council (Carne et al. 1971), Osaka Gas (Maezawa 1973) and Esso (May and McQueen 1973) were also reviewed and included in the analysis and modeling. The AGA program included seven 6-ft diameter fires, six 20-ft diameter fires, and one 80-ft diameter fire; radiation measurements were made with narrow- and wide-angle radiometers, and spectral scans were made with a McPherson 218 Spectrometer. These experiments also included records of weather data, liquid recession rate, and still and motion pictures. Some radiation data also were obtained during extinguishment tests at Marinette, Wisconsin, on 20-ft x 20-ft and 30-ft x 40-ft pools.

Atallah and Raj (Duffy et al. 1974) evaluated the radiative flux at an observing point by the relation

$$I = \tau \varepsilon F E , \qquad (20)$$

where I is the heat flux intensity in kW/m^2, τ is the transmissivity of the atmosphere, ε is the emissivity of the flame, F is the view factor, and E = 31,500 Btu/hr ft^2 is the black body radiation from an optically thick LNG flame. The computed values were compared with radiometer measurements. It was found that the predicted values generally exceed the experimental means and approach the maximum values. There was better agreement with experimental data of the 6-ft fires than the 20-ft fires, which leads Atallah and Raj to believe that part of the flame was not observed by the radiometers when sighted on the 20-ft fire.

Welker (Duffy et al. 1974) suggested a value of E of 45,000 Btu/hr ft^2 based on the same data, but deduced a different value of "b" in the relation

$$E = E_0 (1 - e^{-bD}) \tag{21}$$

that was used to match the data for the various pool sizes. As a result, the agreement of the empirical computations with the data were equally good, but the predictions for larger pool fires deviated significantly.

Parker (1977) recently carried out a reanalysis of the San Clemente flame radiation data and suggested two equations for the heat flux from the flame considered as a cylinder, namely,

$$q = 142 (7/6 - X/L) (1 - \exp (-0.398D)), \text{ kW/m}^2 \tag{22}$$

and

$$q = 101 (1 - \exp (-0.33D)), \text{ kW/m}^2 \tag{23}$$

where D is the diameter of the spill in meters, L is the flame length, and X is the distance from the flame base along the flame length. Equation (22), which includes a term for the observed decreasing radiant source strength as one moves up the flame, is more physically correct, but Equation (23) fits the available data equally well. Parker showed that the equations are compatible with data from the British Gas Corporation fire tests on 18-ft x 18-ft and 20-ft x 40-ft pools (Carne et al. 1971), the Osaka fire test on a 9-ft square pool (Maezawa 1973), the Tokyo fire tests on a 33-ft square pool (Tokyo Gas Co. 1971), and the Esso-Libya fire tests (May and McQueen 1973).

On a cooperative project among USCG, ERDA, and AGA at the Naval Weapons Center at China Lake, tests were carried out with LNG spills on water of 800 to 1500 gallons to determine the liquid pool spreading rate, evaporation rate, burning characteristics and flame radiation. The data, as analyzed by staff at A. D. Little Inc., were reviewed by Raj et al. (1979). The radiation output was nearly twice that observed with land spills. This could be the result of smoke suppression by the water, though some investigators think that the older data were inaccurate.

A computational program for predicting radiation hazard from an LNG fire on water has been outlined recently (Western LNG project 1978). This could possibly be used readily for LPG pool fires on water as well.

8.2.4 Fire Control and Extinction

Control and extinction of pool fires can be accomplished by physically removing heat, by inhibiting the chemical action, or by reducing the available oxygen. The suppressant materials can be solid, liquid, or gaseous and may act in more than one manner.

Sheinson et al. (1978), analyzed the suppressant action of a large variety of chemicals on n-heptane liquid pool flames to determine whether they showed a physical or chemical inhibition. They found, for instance, that suppression by CF_3Br (HALON 1301) was only 20% physical, while CF_3Cl (HALON 1310) was 45% physical.

Recent work on pool fires of JP-4 has shown that the fire can be quenched by replacing approximately half the surrounding atmosphere with nitrogen (Gann et al. 1978). No effect of pool size was observed up to an equivalent of 1.38-ft diameter.

Chakraborty et al. (1975) determined the inhibiting effect of CCl_4, $CHCl_3$, and CH_2Cl_2 on the flammability limits of vaporized LPG, and found that the inhibition increased with number of chlorine atoms. They indicated the effect was through a destructive action on the flame radicals.

Extensive large-scale fire control and extinguishment tests on LNG were conducted on the AGA IS-3-1 program, and were continued in a separate AGA project (Duffy et al. 1974, University Engineers Inc. 1974). Both projects included work with high expansion foams and several dry chemicals including $NaHCO_3$, $KHCO_3$ (Purple K), Urea-$KHCO_3$ (MONNEX), and a monoammonium phosphate. The results should be generally applicable also to LPG fires. The first program included fires in dikes up to 30 ft x 40 ft; the LNG was supplied to the dikes over long time periods, resulting in frozen ground and low boil-off rates. In the second program, high rates of recession of the pool surface were produced in 5-ft and 10-ft diameter pans by hot water heating through immersed tubes. Some 100 experiments were reported in the second program. High recession rates (1.3 and 1.5 in./min) were achieved twice and rates in the range 0.9 to 1.1 in./min were reported in ten experiments. The rates were 81 in./min or less in the remainder of the experiments. The higher burning rates would correspond to 1.2 in./min of LPG. Comparing this with the burning rates for natural feedback, as given earlier, this is about 4 times the expected burning value under normal conditions.

In the dry chemical tests, minimum allowable application rates and extinguishing times of the several chemicals were determined as a function of LNG recession rate, and data correlations were derived. It was found that the dry chemical requirements for extinguishment are about 2.5 times greater for recession rates typical for early times after spill on warm soil as compared to steady rates over frozen soil. High expansion foams with appropriate characteristics reduce the radiation intensity on outside targets near the fire by large amounts (up to 90-95% reductions). The report presents data correlations of fire control time as a function of physical properties of expanded foam.

University Engineers Inc. (1976) extended their studies with a series of fire extinguishment tests to provide additional design information. Suppression of flames from spills on water by dry chemicals and by water screens was studied. The dry chemicals were found to behave the same as for land spills. Water screens did not compare well with direct use of water to cool irradiated targets, or use of water sprays to reduce gas concentration in the downwind air. Zuber (1976) reviewed several methods for fire protection and control from the point of view of control system design. Included were studies of dry chemical extinction of fires up to

30 ft x 40 ft, and foams on fires of 10 ft diameter, as already discussed above.

Richards and Sheehan (1976) report that in suppressing a propane-air explosion in a closed compartment, about 20 times as much water spray by volume and 10 times as much HALON (CF_3Br, CF_2ClBr, and $C_2F_2Br_2$) as Purple-K ($KHCO_3$) was required.

8.2.5 Related Research in the DOE Program

Several areas discussed in Subsection 8.2 are under investigation in the DOE Program on LGF safety and environmental control. The following studies supported by this program complement the efforts reviewed above.

A rather comprehensive survey of fire and radiation hazards from LNG combustion was conducted by Corlett (1980). Included in the survey are ignition, vapor cloud burnup, pool fires and detonation. The state of current understanding of these phenomena is presented and the needs for additional research are outlined.

The LPG Safety Research Project conducted by Applied Technology Corporation (ATC) included an evaluation of fire-fighting agent effectiveness and the measurement of burning rates in LPG fires (Welker et al. 1980). About 100 tests were run on free-burning LPG fires to determine the quantities of fire-fighting agent that is required to control or extinguish a fire. Tests using high expansion foam were run on fires in 5-, 10-, 20-, and 40-ft square pits. It was found that LPG fires can be controlled but not extinguished by the application of high expansion foam. To provide control within a few minutes, application rates in the range of 0.1 to 0.15 gal/min-ft^2 at an expansion ratio of 500 to 1 were required. The wide variability experienced with LNG fires in the effectiveness of various types of foam was also demonstrated with LPG.

Three types of dry chemical were used: sodium bicarbonate, potassium bicarbonate, and urea-potassium bicarbonate. The chemicals were applied from fixed nozzles located along the sides of the test pits. All the dry chemical agents were found effective in extinguishing LPG fires, the sodium bicarbonate being less effective than the other two agents. However, the sodium bicarbonate is an attractive choice on the basis of cost and range for large fixed systems discharging at high rates.

Burning rates were found to vary with fire size and to a lesser extent with wind speed. Not enough data were obtained to provide quantitative relationships; however, the maximum burning rate indicated was about 0.4 in./min and is reached with fires larger than 20 ft wide. If the fuel is ignited before the ground is well frozen, the burning rate can be substantially higher because of more rapid boil-off.

ATC also conducted burning tests to determine radiant fluxes from LPG spills. In the tests, emitted and incident radiant heat fluxes were measured for fires with base areas from 25 ft^2 to 1600 ft^2. Simplified radiant flux models were found to be adequate to represent the results. The maximum effective radiant flux emitted by the propane fires was about 50,000 $Btu/hr\text{-}ft^2$. Fluxes for smaller fires can be expressed by

$$q_s = 50,000 \left(1 - e^{-1.126\ D}\right)$$

where q_s is the effective surface flux ($Btu/hr\text{-}ft^2$) and D is the flame diameter (ft). Flame heights could be predicted using the Thomas equation. A simple model based on an assumed cylindrical shape for the flame and the surface fluxes given by the equation above predicted the incident radiant fluxes surrounding the fire with good accuracy.

The DOE Program included studies on radiation fluxes from both LNG and LPG. Appendix I contains preliminary analysis of LPG pool fire data provided by Welker, Johnson and Cavin of ATC under subcontract to PNL. This effort represents a first step in the analysis of data obtained at the Naval Weapon Center, China Lake, California, on LPG burn tests conducted in 1977. Appendix I provides a summary of text conditions, a comparison of calculated and measured radiant fluxes and a discussion of some problems in the interpretation of results.

8.3 BEHAVIOR OF COMBUSTIBLE LPG MIXTURES

Within the last decade there has been an increasing incidence of unconfined vapor cloud explosions involving propane, LPG, cyclohexane and other petrochemicals (Strehlow 1973, Strehlow and Baker 1976). These explosions have usually been associated with breaks in pipelines, railroad and tank truck accidents, or failures of equipment in petrochemical plants. They have sometimes caused extensive damage to property and injured large numbers of people. Often damage from the excessive pressures have exceeded the fire damage. The explosions usually follow a massive release, producing a large vapor cloud. For instance, Burgess and Zabetakis (1973) report on the heavy damage following ignition of a propane vapor cloud about 980 ft from a pipeline break in Port Hudson, Missouri.

Bearint et al. (1976) present a scenario of the burst of an LPG pipeline, pointing out the importance of the fuel properties, the temperature and pressure before a break occurs, and the size of the break. For the present discussion, the significant point is that a large amount of the fuel may be immediately flashed to a vapor; Hardee and Lee (1973) estimated about 30% for propane. Further, the vapor may carry more fuel in the form of a fine aerosol if the rupture acts as a jet. They assume such action, with subsequent vaporization as air is mixed into the jet. Experimental work with releases up to 1000 lb of propane and of MAPP (methyl acetylene-propylene-propadiene--a mixture used for welding gas) appears to confirm the analysis. Hess et al. (1974) considering propane dispersed from ruptured vessels of 1.6-and 2.4-in. diameter with a length to diameter ratio of 3.5:1, found that the vapor dispersed separately, and the remaining liquid pool vaporized only as heat was supplied from the ground and incoming air. Both of these studies were on simple limiting cases, and in one study, for a relatively small system. There are several other factors that can be involved in the split between vapor and liquid in a rupture. For instance, in the case of a pipeline rupture, the terrain upstream from the rupture will separately affect the rates of supply of vapor and liquid after the initial burst of fluid. Also, in connection with the terrain, it should be noted that LPG vapor is denser than the LNG vapor on which more vapor dispersion data are available. Thus, there is a greater tendency for the LPG vapor to pool in low areas, and to follow the natural terrain. Further, there seems to be some tendency for the vapor cloud to damp out normal turbulence in the wind.

The following subsection reviews deflagration and detonation phenomena as they pertain to potential explosions in an unconfined atmosphere.

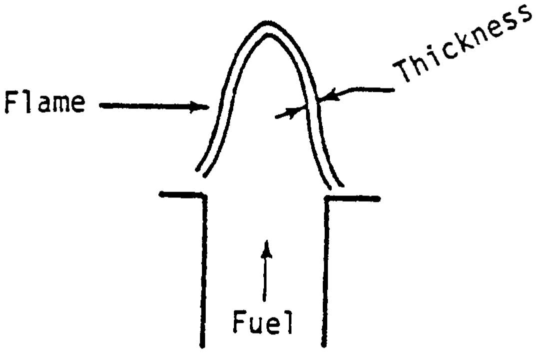
8.3.1 Physical Descriptions of Explosion Phenomena

Four basic phenomena are discussed: deflagration, detonation, fireball formation and flameless explosions.

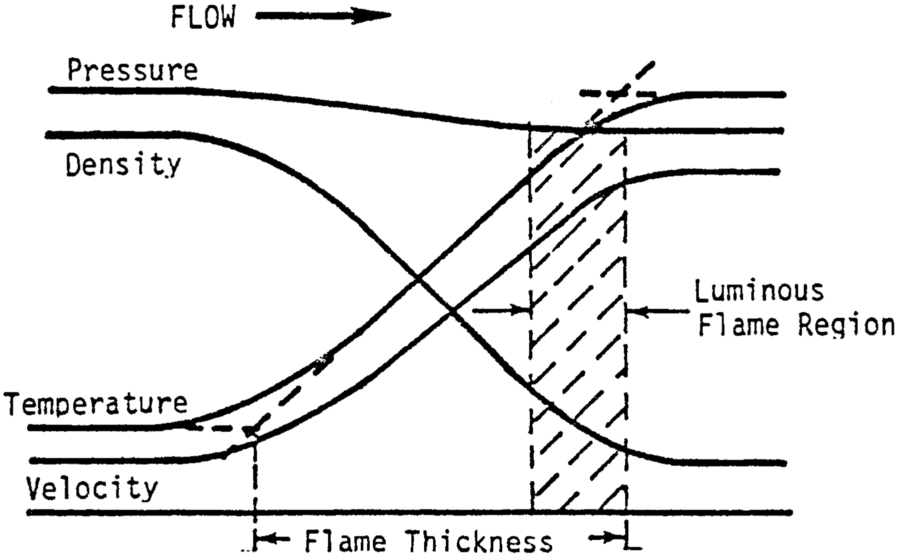
Deflagration

Deflagration is the most common and well-known type of combustion. Flames in gas-fired stoves and residential furnaces present examples mainly of laminar deflagration waves. Industrial gas-fired burners illustrate more often the case of turbulent deflagration waves. Figure 8.1(a) is a sketch of a Bunsen-burner flame and Figure 8.1(b) shows the conditions for such an example of an idealized laminar deflagration. This latter figure shows that as the fuel-air mixture is consumed while passing through the flame front and as the gases heat up, the density decreases considerably, and the flow volume increases correspondingly. There is a small, often neglected, drop in pressure. Maximum laminar burning velocities of common hydrocarbon fuels range from about 1.3 to 10.2 ft/s, that is, up to a Mach number of about 0.1. Propane and butane have peak burning velocities of about 1.6 ft/s.

When the combustible mixture is turbulent, the flame front is highly distorted, and, in fact, may become diffuse in nature, in contrast with the sharply defined outlines of laminar flames. As a result, both the burning rate per unit frontal area and the flame speed increase. The turbulent flame speed may be many times the laminar burning velocity of a mixture, but is finally limited by the rapid stretching and tearing of the flame front. Other than qualitatively, the entire phenomenon of turbulent deflagration is not well understood even after extensive research, and is thus a controversial subject in the literature. Also, since a turbulent deflagration propagating at high velocity can produce pressure waves of significant amplitude, this phenomenon is sometimes mistakenly called a detonation.



(a) Bunsen Burner



(b) Structure of Ideal One-Dimensional Stationary Laminar Deflagration

FIGURE 8.1. Sketches of Simple Flame

Detonation

Detonations are characterized by very rapid consumption of the combustible mixture. In fact, in a detonation, "the flame front is propagated by a shock wave traveling with the speed of sound under the high temperature and pressure conditions of the shock wave" (Gaydon and Wolfhard 1953). Thus, detonation velocities are greater than the velocity of sound at ambient conditions, (about 1080 ft/s). In contrast to the relation shown in Figure 8.1 for deflagration, density increases with the temperature, the pressure increases greatly rather than decreasingly slightly, and the velocity decreases through the thinner (compared to deflagration) flame front. Because of the sudden increase in pressure, strong shock waves are sent out.

One may also distinguish between detonation and deflagration in a more formal and unambiguous manner. The distinction is made on the basis of the Hugoniot diagram, which is a curve showing the after-combustion pressure as a function of specific volume for a given initial set of conditions before combustion. This curve has a forbidden region (resulting from an imaginary solution for mass flux) that separates the detonation phenomenon from the deflagration phenomenon (Williams 1965).

A common way to study the detonation of a mixture is to ignite it at the closed end of a long duct. The easier the mixture is to detonate, the smaller in cross section the duct (one-dimensional case) can be and the shorter the distance before a spark initiated deflagration initiates the detonation. Turbulence generators in the duct will shorten this distance. Detonations are harder to initiate by a flame spreading out from the axis of a cylinder (two-dimensional case) and still more difficult to initiate by a flame spreading from the center of a sphere (three-dimensional case). The latter represents the unconfined condition in the atmosphere, and relates to the present question as to the possibility of detonation of an LPG vapor cloud in unconfined or partly confined conditions.

Fireball

A fireball is a luminous volume of hot gases resulting from a sudden release of heat. A well-known example is the mushroom-like cloud resulting from an atomic explosion, but the rapid propagation of a flame through a locally combustible mixture (vapor cloud explosion) can produce a similar visual effect. The buoyant effect of suddenly heated gases produces an upward-directed jet. A rolling vortex forms around the head of the jet, between it and the stagnant air. The leading gases are rolled into the vortex, producing mixing in the hot, buoyant, ball-shaped vortex structure. If the supply of fuel is limited, this vortex continues to move upward as an entity. A similar structure may form at the start of a fire in a continuous supply of fuel, such as from an ignited pool. When the various parameters are in the correct relation, a flame may be unstable and consist of a lightly connected series of fireballs. In the case of a combustible mixture in a wind-blown plume from an LPG spill, one can conceive of a situation in which a local ignition will produce sufficiently rapid inflow that a flame cannot propagate outward. Rather, combustion occurs in the local area, forcing a fireball. Another possible path leading to the formation of a fireball could start with the ignition of a local pocket of fuel and air with a composition near, or on the fuel-rich side of stoichiometry. In this case, the combustion process could be either a deflagration or detonation. It seems that the fireball, as related to this study, is a special fluid dynamic configuration that can occur under a range of conditions in a mixture of fuel and air. Pressure effects of fireballs involving normally expected flame speeds, as compared to those from detonations, are usually small, or even negligible (Hardee et al. 1978). Radiation effects of large fireballs are not well documented, but reasonable computational approaches are available (Hardee and Lee 1973, Hardee et al. 1978). We might also note that a detonation in a vapor cloud could result in a form of fireball from the buoyant lifting of the resulting hot, radiant volume of gas.

Flameless Explosion

Under the proper set of conditions, a drop of liquid can be superheated above its normal boiling point. Excessive superheat, or a sudden shock, will then result in a very rapid conversion of the droplet to vapor, with a resulting mini-explosion. As an example, a cold liquid drop can be released at the bottom of a denser column of a second, immiscible liquid in which the temperature increases with height. The test liquid drop will move upward from buoyancy effects and heat up to and then beyond its normal boiling point, until it finally explodes. Accompanying this explosion will be a sharp sound and a displacement of the surrounding fluid.

In a case where a large number of superheated droplets are forming, the shock wave from the explosion of one droplet can trigger the explosion of neighboring drops, thus leading possibly to the nearly simultaneous explosion of a significant quantity of liquid. Experimental studies have shown that propane with water as the second fluid may produce this phenomenon under the proper set of conditions. The critical conditions do not appear likely to occur in the case of propane, butane, or LPG, however.

8.3.2 Analytical Modeling Studies

Pressure waves can arise from a deflagration wave propagating from an ignition source in a combustible vapor cloud. Furthermore, the possibility exists for detonation waves with accompanying high pressure to be initiated in such clouds. In both cases, radiation phenomena will also be present, and in the first instance may be the more severe potential hazard. Because of the environmental features which are similar, deflagration and detonation will be considered together here.

Character of Vapor Clouds

Vapor generation and dispersion studies can be used to predict the mean and statistically varying composition in 1) the initial burst of vapor from a rupture, and 2) the plume downwind of an LPG spill, as a function of location and time. The objective of mathematical studies of explosions is to predict the fire and overpressure hazards resulting from an ignition at any point in the LPG vapor cloud on the basis of this input information. The conditions of interest cover the full range of possible spill sizes, terrain, and meteorological conditions.

If the LPG vapor-air mixture resulting from an LPG spill is not ignited at the source of the spill, it may be ignited at any downwind point where it still persists as a combustible mixture. Of primary interest is the probability of an ignition at any point downwind resulting in a flame traveling through the cloud of varying concentration back to the pool. Because a variation of concentration is inherent in the vapor dispersion phenomenon, the effect of wind conditions and distance on the ratio of peak-to-mean concentration can result in a combustible flame-propagating region where the average concentration is below the ignition limit. Thus the farthest downwind distance of interest, the average concentration of the combustible gas will be low, below the ignition limit.

In the case of LNG, to compute this furthest downwind distance of interest, the axis of maximum concentration has been assumed in the literature to be governed by the same meteorological relation as a neutral density gas (Duffy et al. 1974). In effect, it is assumed that the extra effects of increased horizontal spreading and decreased vertical spreading cancel out. At least for the normal turbulent wind conditions, this approach appears to yield reasonably accurate results. However, LPG vapor is significantly denser than LNG vapor, and the LPG vapor cloud is also much denser than air in spite of heat input from the ground. As a result, the details of the ground contour are significant, especially near a spill. Since LPG vapor will tend to settle in lower areas of ground contours. The vapor can also move upwind with incremental mixing at the interface with air. It is not clear how the difference in density between LNG and LPG vapors will affect the overall cloud areas susceptible to ignition.

Combustion of Vapor Clouds

Assuming that a combustible region is ignited, alternative phenomena can be considered. The LPG vapor cloud, because of severe negative-buoyancy spreading, will tend to become somewhat pancake-shaped if not confined in a valley. In any case, the combustible region will be unconfined at the top and at the sides. In a closed system, the flame propagation (deflagration) velocity might be expected to accelerate as the flame spreads; while in this type of open system, the flame propagation velocity is expected to reach a constant value. Raj and Emmons (1975), using scattered data on LNG flames, relate the flame speed linearly to the wind velocity, but Koch and Bokemeier (1977) and Pfortner (1977) put absolute maxima on the value for common hydrocarbon fuels. This propagating flame will send out pressure waves. In a closed system, or with a rapidly accelerating flame, a detonation wave can develop either from ignition caused by shock wave heating or from a very rapidly moving deflagration wave (Strehlow 1973). However, there is considerable question as to whether a detonation can develop in an open system in a cloud of LPG vapor in air. Koch and Bokemeier (1977), Lee et al. (1977) and Pfortner (1977) indicate that it is not possible. However, Pfortner feels that sufficiently rapid deflagration may be developed under the proper conditions with some hydrocarbon fuels so that pressure of the order of 1/3 of an atmosphere may be generated, with damage possibly as intense as from a detonation because of the greater time duration of the pressure.

It might be noted at this point that if a detonation is possible in a fuel vapor cloud, it might be initiated as follows. A deflagration starts from an ignition source in a closed system, such as a house; the deflagration in the closed system accelerates to a point that a detonation is initiated; the detonation breaks out through an opening into the surrounding combustible cloud and continues through the cloud. While, for LNG, the required area of breakout would appear too impractical to consider, for LPG this is probably not true.

The theory for both deflagration and possible detonation of LPG vapors in turbulent atmospheric conditions is far from definitive. This stems primarily from uncertainties as to how a flame would spread through a non-uniform mixture of fuel and air in a turbulent atmosphere. Raj and Emmons (1975) approached the problem by assuming a two-dimensional flame-front geometry, a turbulent, natural convection-controlled burning process, a thin, fuel-rich vapor cloud,(a) and a constant flame-front velocity. They computed flame size therefrom so that they could estimate radiation output. Their theory corresponds to the turbulent buoyant flame theories discussed elsewhere, and no light is cast on the method of predicting the speed of the flame front.

(a) Variations in point-to-point composition do not enter into this method of solution.

Kuhl et al. (1973) discussed the pressure rise caused by constant velocity hydrocarbon flames expanding in planar, cylindrical, or spherical fronts. With a knowledge of the mixture properties critical to detonation initiation, one could presumably determine whether a detonation could develop from a deflagration. However, the turbulent flame speed relative to the unburned mixture would also be required as an input. Strehlow (1975) simplified this self-similar approach by assuming a constant pressure between the flame front and the shock wave. This resulted in some underestimation of pressure rise for a given flame speed. Williams (1975) and Lind (1974) carried this approach even farther, to nonsymmetrical systems, and also considered the requirements for transition to detonation by a combustion wave overtaking the shock wave or by sufficient time and temperature being available for autoignition. While these studies considered LNG specifically, the results appear to apply equally to LPG. They concluded that a "nonideal" explosion could produce a damaging pressure wave over a distance of the order of magnitude of the cloud height. Unless a transition to detonation occurred, severe damage would not extend to large distances, and shocks should be quite weak at distances greater than 10 times the height. In considering the possibility of transition to detonation from the nonideal explosion, Williams and Lind point out that decay due to upward relief lessens the pressure wave significantly at distances that are large compared to the cloud height. Further, the flame speeds tend to approach a constant value, whereas substantial flame acceleration is needed to develop a detonation. Finally these studies point out the importance of confinement in producing a detonation, and conclude that it would be most difficult for a detonation to develop in an unconfined LNG cloud.

8.3.3 Experimental Work

Experimental studies are reviewed as they relate to the basic phenomena described in subsection 8.3.1.

Deflagration

As in the case of mathematical modeling, there is little experimental work directly related to LPG vapor explosions in the atmosphere. Raj and

Emmons (1975) reviewed the work of Gaz de France (1972), TRW (1968), and a portion of the AGA work (Duffy et al. 1974) on flame propagation through LNG vapor clouds, and plotted the results shown in Figure 8.2. They noted that in cases where the flame was ignited downwind, it propagated relative to the gas cloud at about twice the wind velocity, with a thickening flame front. Thus, for a given wind velocity, the flame propagated at essentially a uniform velocity against the wind back to the source. For high gas concentrations, a tall turbulent plume developed. A similar result would be expected with LPG.

In addition to the observation of LNG flame propagation rate in vapor clouds in the open, there have been other experimental studies pertinent to flame propagation of LPG in the atmosphere. For instance, in studies of the propagation of a stoichiometric propane/air flame from an ignition source at the base of a hemispherical balloon of 16-ft radius, Lind and Strehlow (1975) report an upward flame speed of 36 ft/s and a horizontal flame speed of 27 ft/s.

Liebman et al. (1970) reported data on the propagation of flame through a quiescent, layered, methane-air system in a horizontal tunnel open at the bottom to relieve the pressure. The flame propagated at a rate close to the product of the expansion ratio on burning of the stoichiometric mixture and the corresponding laminar burning velocity. The rate was a mild function of the ratio of the fuel contained within the limits of combustion to the total amount of fuel, and a mild function of a Reynolds number based on burning velocity and width of the combustible mixture zone. This latter result is pertinent to the propagation of a flame through a combustible mixture of either LNG or LPG in the atmosphere, but the full significance is not yet clear.

Karim and Tsang (1975) reported a short study in which flames were propagated vertically through various nonhomogeneous methane-air mixtures in a 0.3-ft diameter by 7.9-ft tube, open at the end opposite to the ignition end. The data indicated a quasi-homogeneous flame velocity when the flame was accelerating because of composition change; i.e., the flame velocity

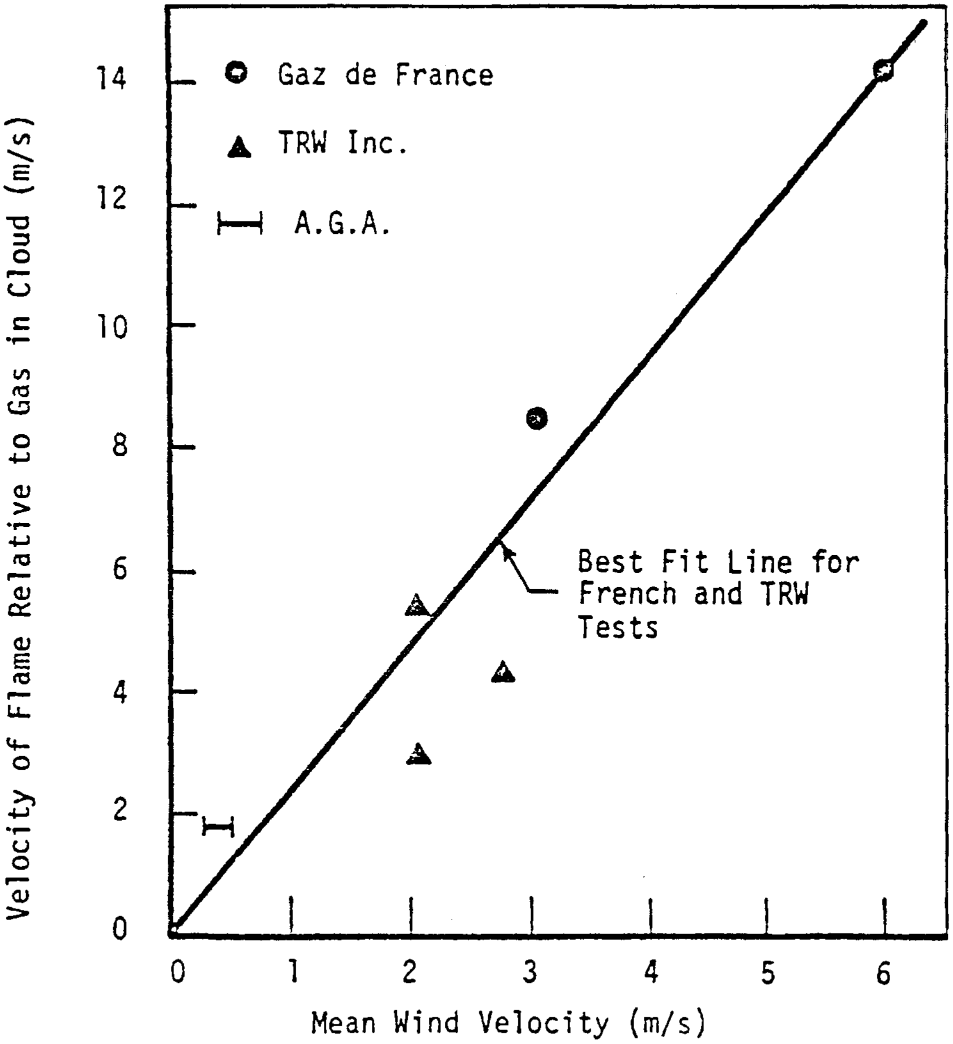


FIGURE 8.2. Experimental Data on Turbulent Flame Velocity Relative to Gas in Vapor Cloud as a Function of Mean Wind Velocity

depends on local composition. However, when the flame was decelerating, the flame velocity was higher than the quasi-homogeneous value (i.e., the flame tended to retain its velocity).

Burgess and Zabetakis (1973) describe in detail the results of the destruction produced by the ignition of a flammable cloud of propane in air following a pipeline break. In this explosion, the source of the ignition was about 980 ft from the break. Burgess and Zabetakis indicated that an open air detonation in the pancake-shaped cloud occurred, probably from a violent explosion exiting from the concrete block warehouse in which an ignition source was present. However, the discussion in this section has indicated the possibilities that the explosive outbreak from the source could have resulted in a violent deflagration occurring in the initial fire (fireball?), which could also have resulted in extensive damage. Thus, no definite conclusion can be drawn on the type of combustion propagation that resulted in extensive damage to structures. It was noted that a more normal deflagration apparently swept through the remainder of the cloud after the explosion.

Detonation

Detonation can be readily produced in propane/air mixtures in tubes. For instance, in a series of studies of stoichiometric mixtures of propane with various O_2/N_2 compositions, Manson et al. (1963) indicated a minimum diameter with air of somewhat over 2 inches. However, this does not prove that detonations can occur in an open cloud of LPG in air, or even that detonation can occur from a source in the center of an open premix of propane and air. Propane could behave like methane, in which a turbulent flame can accelerate in a sufficiently large tube and eventually produce a detonation, but cannot produce a detonation in the open without an excessively large initial explosion being required.

As an example, one can consider the results of experiments (Lind and Strehlow 1975) to study flame front propagation and pressure development in bag tests. In these experiments, 5% propane-air, and 10% methane/air mixtures were ignited with a weak ignition source in hemispherical balloons with 16.4-ft radii. No detonations or overpressures above 0.1 atmosphere were observed. It was noted that vertical flame speeds were somewhat greater than horizontal ones, which may mean that the thickness of vapor clouds is important (Lind and Strehlow 1975). For propane, the upward flame speed was 36 ft/s.

Kogarko, Adushkin and Lyamin (1966) reported that propane/air mixtures were detonated by massive explosive charges in a 9.8-ft radius balloon. These investigators also report the maximum pressure obtained at various distances from the ignition source (Lee et al. 1977, Kogarko, Adushkin and Lyamin 1966).

Bull, Ellsworth and Hooper (1978) determined, for a series of stoichiometric mixtures of hydrocarbon gas with air, the minimum amount of Tetryl explosive charge necessary to produce a detonation in 10-ft x 5-ft (uninflated) polyethylene bags. The values are given in Table 8.1, along with the kinetic induction period at 3140°F, the only one of the property values examined that appeared to show a close correlation.

TABLE 8.1. Detonation Limits and Kinetic Data

Fuel	Methane	Ethane	Propane	n-Butane	Iso-butane	Ethylene
Minimum Tetryl detonation charge, (lbs)	48.5(a)	0.09	0.18	0.18	0.22	0.033
Maximum Tetryl nondetonation charge, (lbs)	--	0.07	0.11	0.11	0.18	0.22
Induction period, μs	3.30	0.17	0.39	0.31	--	0.062

(a) Extrapolated from Bull et al. (1976)

The investigators indicate that "the correlation between high temperature oxidation kinetic rates and detonability measurements of fuel in air is identified but does not accord strictly with available theories." A consideration of the values of Table 8.1 shows that the data are linearly related, except for methane, where the high value for methane seems to imply a changing mechanism, inhibiting detonation even further than expected. An interesting point, however, is that when Boni et al. (1978) used Bull's (1976) data on methane and kinetic data from the literature to extrapolate to the unconfined air condition, they obtained a value of 22,000 to 2.2 million lbs Tetryl rather than about 50 lbs.

Edwards, Hooper and Morgan (1976) earlier had obtained two data points on C_3H_8-$5O_2$-ZN_2 mixtures (Z variable) using the same type of experimental system. These data and a single point (at 0.34 lbs of an explosive generally considered less strong than Tetryl) reported by Kogarko, Adushkin and Lyamin (1966) fell in line. Even though above the value of Bull, Ellsworth and Hooper (1978) the differences are not significant considering the variances in explosive and technique. However, using the extrapolation of Edwards' data suggested by Boni et al. (1978), a value of 0.09 lb is obtained for propane.

There are few data on the size of charge necessary to produce a detonation in a stoichiometric mixture of propane or butane in air, and the available methods to extrapolate data, obtained with various oxygen-nitrogen mixtures to air are not reliable. Also, the energy requirement to produce detonation in propane-air mixtures is far less than to produce detonation in natural gas-air mixtures.

Fireballs

Hardee and Lee (1973) present experimental data showing reaction times of about 2 and 4 seconds for two propane fireballs of 3000 lb fuel (plus stoichiometric air) and 50,000 lb fuel (plus stoichiometric air) to reach the maximum sizes of the order of 40-ft and 98-ft radius, respectively. The data fit a relation derived in an earlier study of liquid propellant fireballs (Bader et al. 1971). For comparison, Lind and Strehlow (1975) report 36-ft/s vertical flame speed and 27-ft/s horizontal flame speed in their studies of premixed propane/air explosions in a hemispherical volume of

16-ft radius. Times computed from these values generally agree with those of Hardee and Lee, even though the assumption of a constant velocity would result in a different relation for fitting the data. Hardee and Lee's data for one mixture indicated a growth of the radius with the cubic root of time, which corresponds to a constant rate of fuel consumption. Even though this does not seem compatible with the assumption of a point ignition source, it apparently conforms to the type of ignition source that these investigators used for a turbulent mixture.

Hardee and Lee found that their prediction of total radiation for fireballs from mixtures of about 2.2 lb, 64 lb and 990 lb propane with 100% excess air to targets on the center, on the final flame surface, and 1-1/2 final radius from the center were in good agreement with observations. In their discussion of LNG fireballs (Hardee et al. 1978) they compare results with premixed fireballs and diffusion flame fireballs, and find that the latter take about twice as long to develop. Presumably, a similar result would be expected for propane fireballs.

Fay and Lewis (1977) reported experimental data on the fireballs formed on ignition of methane, ethane, and propane in soap bubbles from 1.2 to 11.6 in.^{3} in volume. They found that the maximum fireball diameter varied with the cubic root of the fuel vapor volume, and the burning time varied with the one sixth power of the vapor volume. These results agree with the experimental results of Hardee et al. (1978), mentioned above, in spite of the great difference in scale of experiments.

In later experiments Fay et al. (1978) reported that the rate of growth of the non-premixed fireballs was given by the form

$$\delta = \alpha\eta^{\beta} \tag{24}$$

where δ is a dimensionless diameter, η is a dimensionless time; α and β are constants which vary with the fuel but are near unity. Fay et al. (1979) reported on laboratory scale experiments with fireballs of methane, ethane and propane clouds. They found that the radiation characteristics from the fireballs of these three fuels were similar and that they behaved essentially like a grey body. The time-integrated radiant energy flux was always estimated to be in the order of 10% of the initial heating value contained in the experimental sample. The researchers were not sure, however, whether their small scale results could be extrapolated to very large fireballs. The authors note the values for methane do not extrapolate to those found by Hardee et al. (1978).

Thus, even though data available on moderate and large premixed fuel fireballs as well as on small and moderate diffusion flame fireballs seem to correlate well on change in size and time to maximum size, the theories are not well developed and display some questionable aspect. Radiation data on premixed flames agree with the theory quite well, but radiation data on the much smaller diffusion flames leave their extrapolation to large spills in doubt. Raj (1977) ascribed such scaling problems with fireball size, to changes in ignition time, turbulence level, and type of mixing.

Flameless Explosion

A flameless vapor explosion of LNG upon its contacting water was reported by the Bureau of Mines in 1970 (Burgess et al. 1970a,b). The incident was unusual in that the Bureau had made many previous spills of LNG on water without observing any such explosions. In this case, after 70 gallons of LNG were spilled, there was an explosion, delayed by about 1/16 second, which observers compared to the force or sound of a stick of dynamite. Subsequent to this report, the Bureau and many others conducted research into this unexpected phenomenon (Burgess et al. 1972, Nakanishi and Reid 1971, Garland and Atkinson 1971, Katz and Sliepcevich 1971, Enger and Hartman 1972, Rausch and Levine 1973, Porteous and Blander 1975, Porteous and Reid 1976). Shell Pipeline Corporation Laboratory (Enger and Hartman 1972) conducted extensive spill research in which flameless explosions of LNG occurred only when the LNG contained 40 mole percent or less of methane. The Shell paper stated that the explosion is caused by the rapid phase transformation from liquid to gas following superheating of a thin layer of LNG at the water surface. It was concluded that sufficient superheating can occur only under conditions suitable for transition boiling. This basic theory is supported by several other authors (Nakanishi and Reid 1971, Garland and Atkinson 1971, Katz and Sliepcevich 1971, Enger and Hartman 1972, Rausch and Levine 1973, Porteous and Blander 1975, Porteous and Reid 1976). Reid (1976) presents an excellent review of superheat phenomena. Pressures in the air or in the water near these LNG/water explosions appear to have been low order, sufficient only to cause minor damage, such as breakage of glass panes. It appears that possible damage from explosions following large spills will be limited by the available energy release (estimated to be in the range of 2.2 Btu/ft^2), by the limited interface area which can be generated within the delay period, and by the requirements of LNG composition for explosions (Katz and Sliepcevich 1971).

Recent work at the M.I.T. LNG Research Center (Reid et al. 1977) has focused on spills of several pure fuels, including propane and n-butane, on water and other fuels, and several binary fuels on water. For propane, the critical minimum water temperature is 127° to 142°F (Porteous and Reid 1976). The effects of injection velocity of hot liquids on various fuels have also been studied, including water into propane. The authors state that cryogens will show a flameless vapor explosion when spilled on a substrate at the appropriate temperature providing the substrate does not freeze or become highly viscous, and has a vapor pressure less than atmospheric. Nucleate boiling takes place at lower temperatures. Stable film boiling occurs at high temperatures. For impacts of water jets on the cryogen, the range of temperatures is increased upwards. As a result, it would appear that there is no danger of flameless explosion of LPG in any practical situation.

8.3.4 Combustion and Detonation Studies in the DOE Program

In order to gain a better understanding of LNG and LPG fire behavior, it is first necessary to understand the basic combustion phenomena. For this purpose, the DOE Program is supporting a substantial effort at the Lawrence Livermore National Laboratory (LLNL) to model the combustion processes by computer codes. Haselman and Chase (1980) have given an overview of the type of programs which are available and which are still under development. The codes simulate primarily unsteady gaseous combustion kinetics and fluid dynamic processes.

As one example of the type of subroutines which enter into these combustion codes, Westbrook (1980) describes a chemical kinetics submodel which is based on detailed, but still simplified, kinetics of the combustion of methane in air.

Since it is not possible to definitively rule out the occurrence of a detonation in LNG or LPG vapor clouds, potential transitions from deflagration to detonation are still under intense study. Westbrook and Haselman (1979) describe a detailed reaction mechanism for the chemical-kinetic evolution of methane and ethane mixtures with oxygen. Their model indicates that increasing amounts of ethane in the mixture could reduce the induction time for the combustion processes, thus making transition to detonation more likely. Since the behavior of propane is similar to that of ethane, LPG clouds can be expected to be more prone to detonation than LNG clouds.

In a subsequent report, the same authors (Westbrook and Haselman 1980) combine their theoretical model with experimental data to predict the amount of high explosive required to initiate a detonation of a stoichiometric mixture of methane and air. For pure methane, they arrive at about 110 to 220 lbs of high explosives (for a spherical geometry). They are also able to describe the mechanism by which minor amounts of other constituents (e.g. ethane) would "kinetically sensitize" the methane.

Urtiew and Tarver (1980) shed some light on the basic gas-dynamic reasons why detonation waves may gain or lose in strength and decay into simple deflagration waves. By determining the characteristic cell size behind a detonation front, they were able to predict the critical energy needed for initiating and maintaining a detonation process for various fuel types (including methane, ethane, propane, etc.

propane, etc.).

The DOE Program supported small-scale laboratory experiments conducted by LLNL in an open semi-confined test chamber to investigate flame acceleration phenomena (Urtiew 1981). A comparison of ionization probe data with Schlieren records led to the important observation that the actual velocity of the flame front may be less significant than the rate at which the total mass of original mixture is transformed into combustion products. The introduction of obstacles into the flow resulted in higher flame velocities. Several modifications made progressively during the test series caused unanticipated increases in flame acceleration. Schlieren photographs revealed the real cause of sudden accelerations, e.g., the flame getting through the slot under obstacles and starting the burnup process from the bottom up.

The general lack of similar studies by other investigators makes it difficult to compare or correlate these results. A table is presented showing final flame velocities and over pressures measured in different geometries and scales comparing results of these experiments with those of other investigators.

9.0 LPG RELEASE PREVENTION AND CONTROL

Under normal conditions, LPG is a safe and reliable fuel at all stages of handling from production to end use. Its hazardous character is encountered principally when accidentally released from containment or otherwise mishandled. Release prevention and control practices and technology have been developed to minimize the chances and consequences of LPG spills. Section 9 reviews the current state-of-the-art in LPG release prevention and control.

Release prevention encompasses all activities that tend to prevent the occurrence of uncontrolled LPG spills. These activities include efforts by designers, constructors, installers and regulators to assure the maximum integrity of all containers and appurtenances, and efforts by appropriate segments of industry and government to develop safety procedures and to communicate safety information to all persons who handle or use LPG.

When release prevention fails, release control techniques are employed to minimize all consequences of the spill. A complete consequence mitigation strategy includes release detection, spill control, fire prevention, fire detection, fire control and damage control. Release control includes design, planning and technology that limit spill size; and training programs, information, response procedures and equipment that deal with hazards after spills occur. These activities involve designers, constructors, installers, owner/operators, regulators, consumers and the public, especially those who live near pipeline rights-of-way or near LPG facilities.

9.1 RELEASE PREVENTION AND CONTROL IN LPG PRODUCTION FACILITIES

LPG releases in production facilities that cause injuries and property damage have been relatively less frequent than those associated with LPG transportation and use. News media coverage of production accidents is also less because the facilities are typically remote from public access and the effects of the releases are generally contained within the facility boundaries. Appendix H includes some examples of LPG accidents in hydrocarbon processing plants to illustrate the nature of these incidents.

The LPG industry has developed a broad range of safety codes, standards, designs and practices to minimize accident consequences. Safety and environmental considerations in LPG production are similar to those of other hydrocarbon processing plants. The most critical characteristics of hydrocarbons in general, and LPG in particular, are their high flammabilities. Leaking LPG forms vapor clouds which in turn can produce flash fires and blasts over wide ranges of severity. An accident of this kind can be very devastating in a processing plant environment where people are working around the clock and where much complex and expensive equipment is assembled in one area.

9.1.1 Plant Layout and Component Separation

Of the many factors which have an influence on LPG plant safety, those relating to physical plant design play, perhaps, the most important role. These include plant layout, structure design, process design, and fire and explosion protection. Plant layouts with adequate separation distances between components are specified or required by local and state building codes, by consensus standards such as the flammable and combustible liquid code and standards for the storage and handling of liquefied petroleum gases (NFPA 1980). These standards tend to be minimal requirements. Frequently insurers have more stringent standards (Industrial Risk Insurance, No. 631) which reflect their underwriting commitment and rate structure (Nelson 1980).

Concerns about vapor cloud explosion effects are the primary basis for these spacing guides (Gugan 1979). In recent years, vapor cloud explosions have been the predominant cause of the largest losses in the chemical and petrochemical industries (Davenport 1977). Because of trends towards plants of larger capacity, higher pressures, higher temperatures and greater product inventory, these losses have been increasing both in frequency and severity. Safety concerns have increased considerably since the devastating vapor cloud explosion at Flixborough, in England. Both in England and in the U.S., major chemical companies are sponsoring research on explosion prevention and effects investigation. The results of the British efforts are reported by Gugan (1979). U.S. investigations are still in progress, and no results are yet publicly available. It appears that industry is continuing to reduce potential hazards by systematically developing lower risk plant designs (Kuhl et al. 1973).

There are several methods available for roughly calculating the potential effects of a vapor cloud explosion resulting from a spill. A reasonably simple one used by industry is the TNT equivalency method. Spill size has been defined by Davenport (1977) as "a maximum credible spill which is equal to the contents of the largest process vessel or train of vessels not readily isolated. Storage vessels and major supply of fuel pipelines are not considered." The diameter D_c (feet) of a vapor cloud resulting from an instantaneous spill of W pounds is given by

$$D_c = 22.2 \sqrt{\frac{W}{hMv}}$$

where h is the height in feet of the cloud cover, M is the molecular weight of the material, and v is the fraction of the cloud volume represented by vapor when the entire cloud volume is at the midpoint of the explosive range (Nelson 1981).

The energy (f) released in the explosion of the vapor cloud is estimated by the weight W_c (in tons) of an equivalent amount of TNT:

$$W_c = \sqrt{\frac{W \Delta H_c f}{4 \times 10^6}}$$

where ΔH_c is the heat of combustion of the material (about 21,000 BTU per pound for LPG, 2,000 BTU per pount for TNT) and f is the explosive yield factor.

Nelson reports the analysis of recent chemical vapor cloud explosions showing that f ranges from about 0.01 to 0.05. In establishing separation distances in hydrocarbon processing plants, it is customary to use $f = 0.02$.(a)

By knowing the TNT equivalent of an exploding vapor cloud, it is then possible to determine the "safe" separation distance from empirically known relationships between the yield of TNT explosions and the resulting overpressures at given distances. No structures are to be located within the distance corresponding to 3 psi overpressure from the potential center of a maximum size vapor cloud.

9.1.2 Inventory Management

Another approach for improving the safety of hydrocarbon processing plants is to keep the inventory of flammable products as low as possible (Kletz 1981). This can be accomplished by design, operational procedures and shipping out flammable products as fast as possible. Economies of scale are encouraging a trend towards larger hydrocarbon processing plants and storage and transfer facilities. This has placed correspondingly increased demands on fire safety systems and challenges the ability of safety design engineers to provide adequate hazard and fire control systems.

9.1.3 Automatic Protection Systems

Fail-safe features including cut-off valves to limit the escape of hazardous materials and nitrogen filled "dump tanks" which operate automatically in the event of sudden loss of pressure in the plant, are now incorporated extensively, and sometimes redundantly, in modern hydrocarbon

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(a) To what extent this customary factor f is safe still appears to be a wide-open question. It is shown by Gugan (1979) that the yield efficiency can vary over a much wider range (about three orders of magnitude) than indicated here. There are also indications that LPG explosions might be generally more efficient than those of other hydrocarbons.

processing plants (Vervalin 1981). Plants are frequently retrofitted or modified to improve their safety and reliability. Structural designs include fireproofing features designed for two or three hours resistance. However, their explosion resistance may vary over a large range (Stephens 1970). Structures have to be designed to survive earthquakes if located in seismically active regions.(a)

Fire protection systems in hydrocarbon processing plants must be highly reliable (West and Brown 1981). Automatic pumps and sprinklers are important features, now commonly adopted in all modern plans. A water supply of sufficient quantity must be guaranteed at all times. Water piping is usually well protected against potential blast effects and missiles and in consequence is placed underground whenever possible. Water deluge systems are often installed to provide a means for keeping storage tanks and other equipment from being damaged by fire exposure. A few water systems are also designed to provide a means of vapor cloud dispersal. Automatic detectors are used to give early warning of vapor leakage. A combustible gas analyzer system is almost always installed in an LPG plant. It is important, however, that the detectors are located at the right places at sufficiently close spacings. Because many types of gas detection systems are commercially available, systems that provide optimal coverage for a particular plant can be assembled (John 1981).

9.1.4 Risk Assessment

The safety consciousness of many large corporations has improved in recent years. This is particularly true in the hydrocarbon processing industry. In-depth quantitative analyses of risk exposure patterns have become commonplace (Kolodnev 1981). They include fault tree analyses, risk-benefit evaluations, and quantitative methods for assessing the probability and criticality of potential system failures.

(a) At least one major LPG plant fire has been caused by an earthquake. On July 21, 1952, two large butane storage tanks in a refinery were ruptured by an earthquake, resulting in a massive spill. There was no explosion, but a large fire ensued which caused considerable damage (NFPA 1961).

9.1.5 Human Factors

It has been the experience of safety engineers that in practically all technologies the human link has been the weakest and that most accidents are caused by human factors. A most important consideration in a facility as complex as an LPG production plant is the proper training of operating personnel. Comprehensive general safety training is provided with respect to the characteristics of the particular plant. Emergency procedures to be taken in cases such as power failure, steam failure and gas leaks are planned and practiced in detail. It is important that each individual has a clear understanding of what he is supposed to do in emergency situations.

Individual operators provide product release and fire control training at their facilities. In addition, educational institutions offer fire and hazard control courses covering LPG, LNG and petroleum products. Local fire departments are also involved in safety exercises. Communications between the operating personnel are vital and are provided by modern equipment such as portable two-way radios, pocket pagers and communicators.

Plant management is required at regular intervals to verify that all safety precautions are in effect and to rectify any shortcomings that can be identified (Jones 1981).

9.2 RELEASE PREVENTION AND CONTROL IN LPG TRANSPORTATION

Transportation is the most critical step in LPG distribution because bulk quantities are carried through areas used by and accessible to the public. Release prevention and control practices in LPG transportation are discussed below.

9.2.1 Release Prevention in Pipeline Transportation

Pipelines transport over 95 percent of the LPG consumed in the United States. On the basis of accident reports filed with the Department of Transportation (DOT), Office of Pipeline Safety Regulations (OPSR) by the operating companies, pipeline breaks in the ten-year period 1968-1977 have resulted in 40 deaths, 69 seriously injured persons, and more than $5 million in property damage and the loss of 651,000 barrels of LPG product (Materials Transportation Bureau 1978). Some typical examples of pipeline releases are documented in Appendix H. It can be argued that, relative to the number of miles of pipeline and the total LPG throughput, the accident record of the LPG industry is good. However, when compared to other liquids carried in pipelines (crude, gasoline, middle distillates, and ammonia), LPG accounts for two-thirds of the deaths and half of the serious injuries, although it is involved in only 10% of all pipeline accidents. Also, although there appears to be a trend toward fewer accidents per year (183 in 1968-1972 versus 139 in 1973-1977), the dollar loss reported in the latter five-year period is double that in the previous period, and the product loss is greater by 40%. Release prevention practices, starting with standards for pipeline design and construction, are being improved continuously to help reduce the frequency of LPG spills.

Design and Construction Practices

Design of an LPG pipeline involves essentially two steps:

1. Hydraulic design of the line (i.e., fluid flow requirements versus pipe diameter, pipe length, elevation changes, etc.)
2. Mechanical design in terms of steel pipe material, pipe wall thickness and other factors important in handling the pressures determined from the hydraulic design.

Basic inputs to the hydraulic design of an LPG pipeline are 1) the volumetric rate of flow, 2) the product characteristics, 3) LPG temperature during flow, and 4) terrain to be traversed. The final design is usually an optimization of the following factors:

1. The required pressure profile along the pipeline calculated in terms of the pressure gradient, changes of elevation along the pipeline, and the vapor pressure of the LPG at the assumed flow temperature (the internal pressure within the pipeline must be consistently greater than the vapor pressure to prevent vaporization of the LPG).

2. The required pressure gradient along the pipeline considering pipe friction, volume flow rate, and pipe diameter (at an assumed flow temperature).
3. The estimated flow temperature calculated from the initial LPG temperature, energy input due to flow (pipe friction and viscosity) and other sources or sinks of heat (i.e., transfer of heat to or from the earth).

Interrelationships among flow temperature, vapor pressure, viscosity, and pipe friction affect the determination of the pressure profile and the calculated flow temperature; therefore, an iterative solution is required to achieve an acceptable design. In practice, several designs calling for various pipe diameters are determined so cost and performance can be optimized. The final choice of pipe diameter and corresponding pressure profile will be affected by the availability and performance characteristics of valves, pumps, drivers, and other necessary appurtenances to the pipeline.

Prior to the establishment of mandatory federal regulations, industry standards and specifications provided a voluntary standard of practice for liquid pipelines. For example, the American Petroleum Institute (API) sponsored the development of specifications for the manufacture and testing of line pipe (API 5L, 5LX, 5LS specifications); the American National Standards Institute (ANSI) and the American Society of Mechanical Engineers (ASME) developed standards to guide the design and the construction of safe and serviceable pipelines. Numerous items of industry standards have been incorporated into federal regulations.

LPG pipelines are subject to regulations detailed in the Code of Federal Regulations, Title 49, Part 195 (49CFR195) - Transportation of Liquids by Pipeline (Government Printing Office 1980). Part 195 "prescribes rules governing the transportation by pipeline in interstate and foreign commerce of hazardous materials that are subject to Parts 172 and 173 of this chapter, petroleum, and petroleum products." Part 195 specifies requirements for numerous details of material properties, design, construction, hydrostatic testing, operations and maintenance, and accident reporting. Many requirements are stated with reference to particular industry standards.

Perhaps the most significant items (with respect to release prevention) specified by Part 195 and by industry codes, e.g., ANSI B31.4, (American Society of Mechanical Engineers 1974) include the following:

- Requirement for hydrostatic test of each new pipeline to a stress level in the pipe wall which is 125% of specified minimum yield stress (SMYS) (both 49CFR195 and ANSI B31.4). This requirement seems justified by the record that, within thousands of miles of pipelines operated within this margin of safety, there have been no ruptures resulting from original manufacturing or construction defects (Bergman 1974). Additionally, DOT recently published a notice of proposed rulemaking to amend Part 195 by requiring a hydrostatic test on all onshore pipelines carrying highly volatile liquids, which have not been previously tested to at least 1.25 times their maximum operating pressure (Federal Register 1978a).
- Requirement for coatings and cathodic protection on buried pipelines to limit corrosion damage (both Part 195 and B31.4). Although coatings and cathodic protection can be used separately, experience has shown that by far the best protection is provided by use of a combination of the two. The accident report records show that the number of accidents per year attributed to corrosion has declined from 229 in 1968 to 48 in 1976. This favorable trend reflects in large measure the requirements of Part 195 promulgated in 1967 with respect to cathodic protection, inspections for corrosion, and reductions in operating pressure or replacement of pipe in cases of severe corrosion.

Although Part 195 and the industry codes prescribe many details of design, materials testing, construction, and quality control, the ultimate item of quality control is that of the field hydrostatic test indicated above. The requirement for coating and cathodic protection of buried pipelines obviously is an effort to preserve the integrity of a pipeline initially qualified by design to the codes, including the field hydrostatic test, against the ravages of corrosion over many years of service.

Other code requirements for LPG pipelines deserve mention. As already noted, Part 195 stipulates that no liquid pipeline may be operated normally at a maximum operating pressure (MOP) greater than 80% of the hydrostatic pressure (this corresponds to the margin of 125% mentioned above). In the operation of a liquid pipeline, it is expected that internal pressures will occasionally exceed the normal operating pressure; these variations will be caused by the opening or closing of valves, starting or stopping of pumps, etc., which create surges of pressure. Part 195 (paragraph 195.406) stipulates that surges or other variations from normal must not be permitted to exceed 110% of MOP. This requirement can be met by the design and installation of surge suppressors, if needed.

It is to be especially noted that 49CFR195 refers to "Transportation of <u>Liquids</u> ...," making no distinction among the several categories of products carried by pipelines with respect to their characteristics or the hazards presented by them when released accidentally. In a Final Rule published on July 16, 1979, DOT/MTB defined the term "highly volatile liquid" as follows: "Highly volatile liquid" means a liquid which has an absolute vapor pressure of 50 psia or more at 100°F (Federal Register 1979). This notice and other notices of rulemaking provide a mechanism for, and recognize the need for, special requirements for pipelines carrying highly volatile liquids.

Construction of steel pipelines for transportation of liquids is covered by 49CFR195, Subpart D, and by ANSI B31.4-1974, Chapter V. Generally, these regulations and standards deal with inspection of pipe and other components, welds, coatings, qualification of welders, pipe cover, construction records, etc. Part 195 requires that pipelines be constructed in accordance with comprehensive written specifications. Nondestructive inspection is required by the Federal Regulations on 10% of all welds (100% in some critical locations).

The DOT/MTB published a final rule effective March 17, 1978, permitting and specifying requirements for conversion of existing pipelines to liquid service (Federal Register 1978b). The announcement calls attention to the previous requirement that pipelines placed in liquid service after March 31, 1970, were to be designed and constructed in accordance with applicable Federal safety standards, and that the previous requirement was more stringent than necessary when applied to conversions subject to Part 195. Under the new requirements, a carrier prepares and follows a written conversion procedure, which must provide for visual inspection and historical review of the pipeline to identify actual or potential sources of failures. The review must be supplemented with appropriate tests where historical records are insufficient to judge the line's condition. Problem areas must be corrected by repair, replacement, or other alterations. A pressure test must be performed to demonstrate that the structural integrity of the pipeline is sufficient for safe operation. Applicable corrosion control requirements must be met within 12 months after the pipeline is placed in service. Records must be kept of the investigations. tests, and remedial measures conducted on the pipeline.

Operations and Maintenance

Part 195, in addition to requirements already identified, has established minimum safety requirements for pumping equipment, and written procedures for normal operation, maintenance, abnormal operations and emergencies. Apparently previous requirements of Part 195 were not sufficiently specific in some respects, because changes to the regulations have been made final, to clarify intent and facilitate enforcement (Federal Register 1979). According to new rules, it is required that written procedures be incorporated in a manual, and the manual reviewed by the MTB; if in the MTB's view the procedures are inadequate, corrections will be required (paragraph 195.402).

System Control. The overall control of liquid pipelines, as reported by Koch and Lutz (1969) and verified by a survey (Bearint et al. 1976), is from a central location. The type of control ranges from manual call-in of parameters, which are then usually displayed on a control panel, to complete computer control. Over half of the 29 pipelines (LPG and NH_3) surveyed by Bearint et al. (representing over 19,000 miles of pipeline) had computers in their operations, but in varying degrees or modes. Some lines reported using a main computer in conjunction with various types of peripheral equipment such as hard wired controllers. Two computers of the same type, one on standby, are also used. A large number of pipelines also use a manually operated control panel with various types of peripheral equipment.

There appears to be a trend toward incorporating centralized control systems in pipeline systems transporting highly volatile liquids, but these are not truly computer-controlled systems (Berint et al. 1976). The computer is used in an adjunct capacity, coding and decoding data, checking validity of transmitted data, calculating, receiving orders, and supplying information to the operator. The operator makes the key decisions.

In order to maintain throughput of product, keep a continuous account of the amount pumped, and protect against pipeline and equipment damage, numerous pipeline and equipment variables are monitored and logged, both at the pump stations and at central control. The specific variables used and the manner of use vary from pipeline to pipeline, but generally include pump suction and discharge pressures, commodity flow rate, temperature, and density, pump and drive running status, outputs of combustible vapor and flame sensors, and positions of pump station and mainline block valves. The transducers in a liquid pipeline control system are monitored by controllers which compare the instantaneous value against a set value. The controllers take action to minimize the differences; however, if the differences exceed certain limits, the system is shut down.

Operating variables that are used to shut down the pump station are mainly variables of the pump, such as: bearing failure, seal failure, case pressure, and case temperature. Some of the variables monitored for shutdown of the pipeline system are discharge pressure, suction pressure, and flow rate. The latter variables are good indicators of the condition of the pipeline, but only for a short distance down the line. Other transducers are needed along the line in order to monitor the condition of an entire line section.

Pump Station Control. In order to maintain operating parameters or to initiate a status change in the pipeline, controllers and actuators at the pumping stations are used. The types of pump station actuators reported in use are mainly, in order of the number reported, electronic, pneumatic, and hydraulic (Bearint et al. 1976). The operating set points on these controllers are usually manually set at the remote sites. Most controllers being used are electronic because they are the most economical type that can provide the tight control necessary to run with the control valve normally wide open (Brainerd 1964) A good system is needed to prevent pressure overshoots, and some controllers, such as pneumatic types, may not respond fast enough if the running pressure is close to the maximum operating pressure.

In the survey of pipelines carrying highly volatile liquids, almost all reported having pump stations with both automatic and remote shutdown capabilities. The local automatic shutdown capability protects the primary pumps from damage and prevents overpressuring of both pump station and trunkline pipeline.

The time required to isolate a pump station or shut down the pipeline system varies with the degree of automatic controls. Most pipelines require 2 to 4 minutes to shut down a pump station either automatically or remotely. It may take up to 2 hours or more to manually shut down a pipeline section due to the fact that in general all of the block valves are not automatic and some must be manually closed to isolate a pipeline section.

With regard to the shutdown capability of pump stations, 43% of those survey had an automatic shutdown capability and 87% had a remote control capability. Many pump stations have dual or triple means of shutting down the pump stations (Bearint et al. 1976).

Paragraph 195.262 of 49CFR195 requires that each pump station must be provided with 1) safety devices to prevent overpressuring, including auxiliary pumping equipment within the station, 2) a device for the emergency shutdown of each station, and 3) and auxiliary power supply, if power is required to actuate the safety device.

Third Party Damage. A large fraction of incidents involving release of products from pipelines involves damage by excavating equipment ("third party damage"). In a study of third party damage incidents on natural gas pipelines, it was found that farmers were responsible for 35% of the incidents, government units 26%, utilities 11% and developers 10% (Connell and Smith 1968). One type of effort directed toward reduction or elimination of third party damage is the institution of communication systems in particular areas, variously called "cooperative area notification plans" or "one-call systems." The idea is that, when any excavator plans to dig in a certain location, he can call one telephone number some time beforehand, say 48 hours, and all utilities having facilities at or near the site will be notified. An NTSB report indicates a markedly downward trend in damages in areas covered by one-call systems (NTSB 1971).

Maintenance Procedures. As noted above, federal regulations require that operators have written procedures for maintenance, the scope of which includes (1) maintaining maps and records, (2) maintaining line markers, (3) inspection of right-of-way, (4) external and internal corrosion control, (5) valve maintenance, (6) inspection and testing of overpressure devices and relief valves.

Accident Reporting. A requirement of Part 195 already noted is that pipeline operators must report to the DOT/OPSR releases of commodity according to specified rules and on a specified form (Form 7000 1). The accident data so gathered could provide a basis for observing trends, analyzing causes and determining whether changes in rules have desired effects. However, at least until very recently, the DOT/OPSR has not analyzed the data gathered since 1968 other than to summarize them, and apparently has not monitored the operators' reports for completeness and accuracy (NTSB 1978a).

9.2.2 Release Control in Pipeline Transportation

Because a pipeline can fail at any point along its length, it is more likely to be discovered by people other than pipeline company personnel and to affect anyone who happens to be near the break at the time. These may include residents near the right-of-way, travelers on nearby roads, farmers and others operating machinery or vehicles. Accident histories indicate that people can become engulfed in a flammable vapor cloud without recognizing the danger, and others are injured because their own activity has broken a pipeline (third party damage). In any case, control of the quantity of spill depends on several factors:

- Early detection of leak or break
- Shutdown time of pipeline section
- Shutdown time of pump stations
- Quantity of liquid between block valves.

The last item--the quantity of liquid between block valves--is, of course, determined during design and construction of the pipeline.

Shutdown Response

In a survey of LPG and NH_3 pipeline operators, it was reported that almost all pipelines have both automatic and remote shut down capabilities for the pump stations. The pump stations can be shut down in 2 to 4 minutes either automatically or remotely (Bearint et al. 1976). (The local automatic shutdown capability protects the primary pumps from damage and prevents the overpressuring of pump station and trunk line piping.) The survey posed a question to operators concerning their policy regarding continued pumping of LPG or other hazardous liquid when a failure of significant magnitude exists at a known location on the right-of-way. All answers to the question were that the system would be shut down and repairs made. An exception to this procedure was indicated by only one operator. This pipeline company, which batches propane, indicated that it would continue to pump, provided the leakage was small and the site isolated. The pumping would continue until a less volatile product was at the site of the leak. Responses indicated generally that emergency procedures require shutting down up-stream pump stations as soon as possible, diverting or stopping the up-stream flow, while continuing to pump on the down-stream side of the leak as long as possible.

Shutdown Response Time

The time required to shut down a pipeline section, i.e., to close the block valves, depends on whether they can be operated remotely or manually. Manual operation requires that personnel go to the valve, which may require trips of many miles, sometimes over back roads or in rough country. Typically these trips require appreciable time, up to 2 hours, or even more. Unfortunately, the vast majority (>80%) of block valves on LPG pipelines are manually operated (Bearint et al. 1976).

Leak Detection

Early detection of a leak can be accomplished by pressure or flow transducers closely spaced along the pipeline; however, at present such equipment is limited almost entirely to pump stations. Shutdown of a pipeline is based largely on the experience of the operator, using the data from pressure transducers and over/short(a) instrumentation at stations. The latter consist of turbine or positive displacement meters, plus integration equipment. Bearint et al. (1976) showed that only 50% of pipeline operators reported having over/short equipment and for these the intervals over which they made checks ranged from 2 minutes to monthly. Three of the most interesting systems reported were: 1) comparative metering, ± 2% of flow on an hourly basis, 2) hourly check, leak detection to 1% or 10 barrels per hour, and 3) average compensated flow rate with 2-minute-interval checks, sensitive to 1 barrel per hour with the alarm set at 5 barrels per hour. Due to the compressibility of the product, leak detection systems cannot detect small losses. Operators commented that leak detection systems are not sensitive enough.

Information, Procedures, and Communications

The above discussion of release control has dealt mainly with pipeline design and system characteristics. Other aspects of hazard control with respect to accidental LPG releases involve information systems, procedures and communications. As in any area where safety is an important concern, it is axiomatic that safety will be promoted and injuries and damage reduced if people know something specifically about the particular hazards and how to cope with them, and if detailed procedures have been developed, frequently

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(a) Over/short. In a pipeline gathering system O&S refers to the perennial imbalance between calculated product on hand and the actual product on hand. This is owing to contraction, evaporation, improper measuring of tanks, and losses through undetected leaks. Product is paid for on the basis of the amount shown in the lease tanks. By the time the product is received at the central gathering station, the amounts invariably are short, which represents a loss to the pipeline system.

reviewed, and released. There is abundant evidence, based on reports and analyses of LPG pipeline accidental releases, that written procedures have often been non-existent or inadequate. Procedures had, in some cases, not been reviewed recently, personnel had been inadequately trained, and the public who reside or work near rights-of-way had not been sufficiently informed as to the product, its hazards, or how to respond to a pipeline break. Even emergency personnel such as police and fire departments in many cases had no experience in dealing with a large conflagration typical of large pipeline breaks, or training in appropriate procedures. In order to improve this state of affairs, 49CFR195 was amended recently to increase the requirements for normal, maintenance, and emergency procedures (Federal Register 1978f and 1979). The rules also establish new requirements governing the training of carrier personnel and educating public agencies and the general public about the hazards of liquids being transported. This final rule also establishes a definition of "highly volatile liquid" as one having a vapor pressure of 40 psia or greater at 100°F; LPG is by definition a "highly volatile liquid."

Emergency Response

Actions which should be taken for safety at the actual site of a pipeline break include, most importantly, 1) stopping traffic or otherwise limiting approach of the public to the site, taking into account the wind and possible or likely changes in wind direction, 2) if possible, the installation of temporary plugs in the pipeline to limit the quantity of liquid which can be released, and 3) to consider intentional ignition of the vapor cloud, taking into account its present extent and weighing the damage which will be done by the combustion of the cloud and continued combustion of the LPG at the break, versus the risks of allowing the cloud to continue unignited.(a) If fire has already started, it will generally be better not to attempt extinguishment. Probably all liquid pipeline operators have plugging equipment available, or have arrangements with contractors to provide plugging service on a normal or an emergency basis; the techniques for temporarily plugging and by-passing a section which has broken or which is to be repaired are well

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(a) This option is very difficult to perform safely, primarily because response times are usually somewhat delayed. There also may be some legal problems encountered with this alternative.

developed and are incorporated in standard operating procedures. The procedures require excavation, welding, insertion of plugs, and laying of a by-pass pipe, all of which will normally require several hours, even if the equipment is readily available at the site.

9.2.3 Release Prevention and Control R&D for LPG Pipelines

Most research and development which contributes to release prevention and control, directly or indirectly, can be considered in several categories: 1) improvement of pipe steels, welding materials or procedures, coatings and other means of limiting corrosion, and appurtenances; 2) understanding causes of accidental releases; 3) better operating procedures or practices; 4) improved emergency response capabilities; and 5) better training and hazard awareness.

Material Development

Hardware-oriented R&D will not be discussed in detail here. However, it can be said in general that any pipeline can be designed and constructed so that it will have a high degree of structural integrity. Incentives to provide an adequate degree of integrity include the safety of the general public, the safety of employees, avoiding loss of products and damage to company and non-company property, and preventing costly shut downs. Toward this objective, as previously noted, the pipeline industry, represented by the API, ANSI, ASME, etc., and the Federal government have established specifications and regulations which are intended to produce the desired integrity and serviceability of pipelines. Committees which consider and promulgate changes in industry codes include a broad range of representatives, of manufacturers (steel, coatings, components, etc.), and of the pipeline operators, and various experts and researchers. A vast amount of research has been conducted over the years toward better pipe materials; better welding materials and techniques; more knowledge concerning fracture of pipelines; the significance of chemistry and metallurgy, residual stresses, and operating stresses in fracture behavior; and better understanding of the function and value of hydrostatic testing and other methods of quality control on reliability. To put the question of materials in pipeline systems in perspective, the results of a study (NTSB 1978a) on liquid pipeline accidents indicate that the causes of 2294 accidents (1968-1976) for all liquid pipelines ranked as follows:

Corrosion	45%
Equipment Rupturing Line	28%
Defective Pipe	7%
Defective Welds	5%
Incorrect Operations	2%
Other Causes	13%

Some 12% of all accidents were attributed to defects in the pipelines themselves; these defects might have included manufacturing or field welding defects or damage to the pipeline due to handling during laying into the ditch or due to backfilling.

Knowledge of Accident Causes

Pipeline accident reports filed with the OPSR by pipeline operators list the general causes of pipeline accidents. The form provides the possibility of indicating as cause 1) corrosion, 2) defective pipe, 3) defective weld, 4) incorrect operation by carrier personnel, 5) equipment rupturing line, or 6) other (specify). These accident reports are summarized annually by OPSR and are analyzed by the National Transportation Safety Board (NTSB). Some recent conclusions reached by the NTSB (1978a) include the following:

- Among the various possible causes of accidents, that of equipment rupturing line was the leader in losses of product as well as total casualties from 1968 to 1976.
- Michigan did not have any reported equipment-caused ruptures in 1974, 1975, or 1976. NTSB attributes this in part to the state-wide "one-call system."(a)
- Instructions for filling out the DOT Form 7000 1 are not adequate to insure consistency and thoroughness. Also, the form does not request

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(a) All operators of underground utilities are required by State law to participate in this system.

sufficient kinds of data to support the formulation of safety regulations.

- There is no way to predict the safe service life of a liquid pipeline using the currently available data.

The NTSB issued several recommendations on the basis of its study of liquid pipeline accident data:

- The OPSR should publish a plan describing how it will use accident report data to formulate safety regulations and to develop a safe service life model for pipelines.
- The OPSR should redesign the Form 7000 1 to include data similar to that collected for natural gas pipelines.
- The OPSR should computerize the liquid pipeline accident reporting system to permit analysis and determination of meaningful relationships among leak rates per mile and various pipeline/product parameters.
- The OPSR should expedite rulemaking concerning LPG pipelines.
- The API should urge member pipeline companies to participate in and encourage improvement in any one-call systems where they operate, and help organize systems where they do not exist. The Governors of Texas and Oklahoma were urged to take action to implement state wide one-call systems in their states.

Another source of information on release causes is the reports of investigations of particular accidents by the NTSB and the Bureau of Mines (NTSB 1973c; NTSB 1978b; NTSB 1976a; NTSB 1976b; NTSB 1972; and Burgess and Zabetakis 1973). In addition to calling attention to the magnitude of combustible vapor clouds which can form and the damage which can be done by combustion of such clouds when an LPG/NGL line ruptures, the reports also indicate basic causes and make suggestions for improving pipeline safety. The following are examples from these six cited reports:

- In the case of a pipeline which ruptured in Missouri in 1970, 10 persons were injured in an explosion estimated to be equivalent to 50 tons of TNT. This pipeline had ruptured 12 times in the previous 6-year period, all due to failure of the longitudinal welds.

- Another pipeline, in Texas, ruptured in the longitudinal weld in February, 1976, killing 5 people. This pipeline had suffered a series of 14 longitudinal pipe seam failures in the period 1968 to 1976.
- In January, 1976, in an accident at an LPG compressor station in Oklahoma, 5 men were killed and 2 were burned seriously when NGL was released into a ditch by an error in procedure.
- Another accident occurred in Michigan in August, 1975, when an 8-inch propane pipeline ruptured due to previous mechanical damage. Nine persons were burned.

Improved Operations and Practices

The feasibility of limiting of pressure to prevent pipeline failure due to overpressure was investigated for the DOT, as part of a larger study (Platus et al. 1974). This study identified a variety of systems and devices which have been used for decades to limit pressures on pipelines. However, the study identified as a relatively new development the existence of supervisory control systems which link the various pipeline components, including pressure-limiting devices, into a coherent system. The report indicated that almost all liquid lines have some means of addressing the problem of pressure transients (surges), but the results of a questionnaire showed that surges are not a significant problem in liquid pipelines.(a)

In another project, "Transportation of Highly Volatile, Toxic, or Corrosive Liquids by Pipeline," sponsored by the OPSR, the objective was to identify and catalog the specific design, construction, operation, and maintenance practices of pipeline operators, and the problems associated with the transportation of highly volatile, toxic, or corrosive liquids (Bearint et al. 1976). Information was gathered from questionnaires sent to operators of LPG and NH_3 pipelines, personal contacts, and survey of pertinent literature. This information was gathered and analyzed in order to provide a data base for DOT rulemaking. The most significant conclusions and recommendations reached in this project relating to release prevention and control are:

(a) This conclusion is questionable. Surges may, in fact, be a prime cause of pipeline rupture.

- Pipe quality and liquid pipeline safety can be enhanced at the pipe mill by control of the "carbon equivalent" to improve weldability, and by normalizing ERW longitudinal seam welds to improve the toughness of the heat-affected-zone.
- Federal regulations (in Part 195) should continue calling for field tests to 125% of MOP on new construction and existing systems that have been relocated, replaced, or otherwise changed. Also, reconsider requiring such a test on existing pipelines that have not been tested in this manner.
- The OPSR should consider modifying Part 195 to make it mandatory that any pipeline that is to be converted to highly volatile liquid service must have
  1. The original design of the line reviewed
  2. A hydrostatic test to 125% of MOP
  3. Corrosion, construction, and maintenance records carefully examined before the system is put into service.
- The OPS should specify a review interval for the written procedures covering normal shutdown, emergency shutdown, and an accidental release of liquid.
- All pipelines, and in particular those transporting LPG or NH_3, should be encouraged to consider the safety benefits of cooperative notification plans.

As discussed above, recent amendments of 49CFR195 have taken many of these conclusions and recommendations into account.

Emergency Response and Hazard Awareness

Activity in this area has been mainly regulatory, as described above. The NFPA has recently developed, under contract to MTB, a training course titled "Handling Pipeline Transportation Emergencies," which consists of 139 slides, 2 cassette tapes, a course work guide, and a performance manual for student use (NFPA 1979a).

9.2.4 Release Prevention in LPG Transportation by Truck

LPG releases associated with truck transportation occur mainly in three situations: 1) during loading and unloading operations, 2) in repair garages and 3) in collisions or mishaps on the road. Representative accident case histories are included in Appendix H. As with LPG pipelines, the principal approach in release prevention is compliance with government regulations and industry codes and standards. These are summarized below as they apply to design and construction practices, general and transfer operation, maintenance and repair.

Design and Construction Practices

LPG trucking is regulated by the DOT. General industry guidelines are also provided by NFPA 58. These standards are often adopted as law by many states. The general objective of LPG regulations, therefore, is the prevention of accidental LPG releases and the consequent production of a flammable cloud. The more important aspects of the regulations applicable to LPG transportation by truck are presented below.

Interstate LPG trucking is regulated by the DOT under 49 CFR, Parts 172, 173, 177, and 178 (GPO 1980). The general scope of these parts is as follows:

- Part 172 contains a Table of Hazardous Materials (paragraph 172.101) identifying materials by name, the hazard classification of each, labels required on packages, and specific packaging requirements. In this part, LPG is classed as a "flammable gas," and must be labeled so, unless excepted. This part also deals with 1) shipping papers and information required on them; 2) marking requirements, which refers to the shipping name (in this case "liquefied petroleum gas"); color, size, location, etc., of the marking; 3) requirements for labeling; and 4) placarding requirements.
- Part 173 defines hazardous materials for transportation purposes and prescribes certain requirements to be observed in preparing them for shipment. "Flammable gas" is defined in paragraph 173.300. Subparts

pertinent to LPG include 1) Subpart A - General, 2) Subpart B - Preparation of Hazardous Materials for Transportation, and 3) Subpart G - Compressed Gases Definition and Preparation. This part also covers specifically (in 173.33) the qualification, maintenance, and use of cargo tanks.

- Part 177 deals specifically with transportation by highway, and generally pertains to the vehicle and operations, rather than containers.
- Part 178 prescribes the manufacturing and testing specifications for containers. Subpart C gives specifications for cylinders, Subpart H for portable tanks, and Subpart J for containers for motor vehicle transportation (paragraph 178.337). "Portable tank" is defined as any packaging (except a cylinder having a 1000-pound or less water weight capacity) over 110 U.S. gallons capacity and designed primarily to be loaded into, or on, or temporarily attached to, a transport vehicle, and equipped with skids, mounting, or accessories to facilitate handling of the tank by mechanical means (paragraph 171.8).

Intrastate trucking of LPG is subject to NFPA 58, specifically Chapter 6, "Truck Transportation of LP Gas" (NFPA 1979). Paragraph 6002 of NFPA 58 indicates that "many of the provisions of Chapter 6 are identical or similar to DOT regulations and are intended to extend these provisions to areas not subject to DOT regulation," i.e., intrastate operations. The design and construction of containers is covered in Chapter 2, and filling of containers in Chapter 4, of NFPA 58.

49CFR173.315 stipulates that LPG shall be shipped only in cargo tanks (Specification MC-330 or 331) or in portable containers (Specification DOT-51), or as provided in 173.32 and 173.33. (Liquefied gases may also be charged into and transported in cylinders as indicated in paragraph 173.304; cylinders will be discussed in another section). Specification MC-330 is obsolete, and tanks have not been built to this specification since 1967; however, MC-330 tanks continue in use. Specification MC-331 is described in paragraph 178.337, and DOT-51 in paragraph 178.245. Both MC-331 and DOT-51 tanks must be built to the same ASME code (ASME 1976).

The designs of vehicles to carry cargo tanks or portable tanks are not specified in detail in the regulations except with respect to requirements for fire extinguishers, chock blocks, exhaust systems, lights and wiring and truck bodies (tie downs for hauling cylinders, etc.). Specification MC-331 (49CFR178.337) includes the following items for LPG containers:

General Requirements. Tanks must be seamless or welded steel construction and designed and constructed in accordance with the ASME Code. The design pressure shall not be less than the vapor pressure of the commodity at 115°F. Excess-pressure relief valves shall be located in the top of the tank or heads. Post-weld heat treatment must be as prescribed in the ASME Code, except that each tank constructed in accordance with Part UH of the Code must be post-weld heat treated. In no event shall the post-weld treatment be at less than 1050°F.

Materials. All materials used for construction of the tank and appurtenances must be suitable for use with the commodities to be transported and must comply with requirements of the Code and of the ASTM. Impact tests are required on steel used in fabrication of each tank constructed in accordance with Part UHT of the ASME Code.

Design Stress Considerations. Specification MC-331 includes requirements on minimum thicknesses of shell, head, etc.; on minimum stresses; and prescribes a formula for calculation of stresses. Webster (1965) discusses these calculations and other aspects of MC-331.

Construction. Tanks are to be assembled generally according to requirements of the ASME Code. These requirements cover welding procedures, welder qualifications and performance tests, preparation of joints, cutting of openings, tolerances for misalignment, etc.

Outlets and Safety Devices. Excess-flow check valves or backflow check valves may be used on vapor lines, but liquid-discharge lines must be fitted with internal, self-closing valves. The latter function so that they are closed during transport, must be held open during unloading operations, and close automatically or manually in event of an emergency. Also, tanks greater than 3500-gallon capacity must have remote-control stations at each end of the tank and diagonally opposite each other for operation of valves. For smaller tanks, at least one remote-control station must be provided.

The internal valves serve as primary shut-off valves, excess-flow valves and as back-pressure valves. The valves are normally in a closed position. A lever must be pulled to open the valve during loading and unloading operations. Each internal valve contains a built-in excess-flow valve. If the flow of LPG through the valve exceeds the rating of the excess-flow spring, the valve closes to reduce the chance of uncontrolled vapor or liquid discharge. The valve's mechanism is self-closing, allowing quick closure whenever the operating lever is released. The valve opening also incorporates a fusible link which is designed to melt in the case of a fire, thus allowing the valve to close.

Basic piping systems in the MC-331 cargo tank includes the sprayfill pipe which enters the tank vertically, near the bottom of the tank. Within the tank, it extends to the top, where it bends forward. LPG is loaded through this pipe and is sprayed forward. As it sprays, it flashes to form vapor and cold liquid. Due to internal pressure and the presence of cold liquid, the excess vapor is condensed back to liquid, eliminating the need to remove excess vapor. A vapor line is available to withdraw vapor from the top of the tank and return it to a storage tank, if necessary. The unloading, or liquid withdrawal line connects at the bottom of the tank. Each of these lines is connected to the tank by an internal valve. An internal valve also connects to the pumping system used on the cargo tank (NTSB 1973a). On most of these tanks, the pump and external piping are located below the tank at approximately mid-length (see Figure 5.3).

The exterior piping and valve system consists of a series of angle valves and globe valves connected by pipe and sealed off with an end cap. Each portion of liquid piping that can be closed off at both ends by valves or end caps must be provided with a hydrostatic relief valve. These valves are typically 3/8 inch in diameter and have a start-to-discharge pressure of about 400 psig (Fisher Controls Company 1976). The packing for the inner valve is made of polytetrafluoroethylene (PTFE). Seals are made of synthetic rubber as are the seat discs (Fisher Controls Company 1978).

The MC-331 cargo tank is also provided with liquid level vent valves. These valves are used to give a positive visual indication of the liquid reaching the maximum allowable fill level within the tank. Liquid level vent valves are used to ensure that LPG tanks are not overfilled. An O-ring seal prevents leakage from around the top of the valves. These valves are set within the side walls of the MC-331 tank, at the level of the maximum legal LPG fill (Fisher Controls Company 1976).

Fittings must be protected against damage from collision with other vehicles or objects. Safety relief valves, in particular, must be protected so that in event of overturn of the vehicle on a hard surface, opening will not be prevented and discharge will not be restricted. There are specific requirements on the design of these protective devices or housings.

Qualification. Every cargo tank used for the transportation of hazardous materials must be an "authorized" container by virtue of testing and inspections as prescribed in 49CFR173.33. A container, less any fittings, must be subjected to a hydrostatic or pneumatic test to a pressure 1.5 times the design pressure.

An amendment to 49CFR178 effective July 1, 1979 prohibits the location of manhole assemblies on the front heat of MC-331 cargo tanks. This is intended to reduce the probability of release of cargo in accidents because of collision damage to the manhole assembly (Federal Register 1978d).

General Operations

As noted above, LPG truck operation and maintenance are regulated mainly by Parts 173 and 177 of Title 49, and by NFPA 58. Parts 390 to 397 of Title 49, "Motor Carrier Safety Regulations" also apply.

Recommendations and procedures for safe operation of truck transport are included in "LP-Gas Safety Handbook" prepared and published by National

LP-Gas Association,(a) and "Handbook of Compressed Gases," a Compressed Gas Association publication (NLPGA 1978 b and Reinhold Publishing Co. 1966).

Interstate transport of LPG by truck is subject to regulation by the U.S. DOT, and thus, to CFR Title 49, Part 177. Insofar as the truck transport of portable containers is concerned, provisions for filling quantities, protection of valves, loading procedures and marking are specified in NFPA No. 58, Chapter 6. Similarly, provisions for cargo vehicles covering safety relief valves, filling connections, the need for remodeling controlled shutoff valves, gaging devices, drainage openings, necessary labeling and protection of container appurtenances are specified.

All pipe, tubing, fittings, valves, hoses and flexible connectors must be inspected daily to ensure that they are in good condition, free from leaks and not damaged in any way.

The regulations of the Federal Highway Administration FHWA, as detailed in 49CFR390-397, pertain to the driver and, generally, to the transport of LPG on public highways. The titles of several Parts indicate the general nature of these regulations:

Part 390 - Federal Motor Carrier Safety Regulations: General
Part 391 - Qualifications of Drivers
Part 392 - Driving of Motor Vehicles
Part 393 - Parts and Accessories Necessary for Safe Operation
Part 394 - Notification, Reporting and Recording of Accidents
Part 395 - Hours of Service of Drivers
Part 396 - Inspection and Maintenance
Part 397 - Transportation of Hazardous Materials; Driving and Parking Rules.

Transfer Operations

LPG is usually transfered by a liquid pump, although propane under some circumstances may not require mechanical assistance. On the other hand, butane in cold climates may require both liquid pump and vapor compressor. If a compressor is used, it draws vapor from the receiving

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(a) National LP-Gas Association, 1301 West 22nd Street, Oak Brook, IL 60521.

vessel, compresses and discharges the vapor into the delivery tank, thus creating a pressure differential and forcing the liquid to flow from delivery to receiving tanks.

When undertaking transfer operations, propane must never be loaded into a vessel designed for butane, although the converse is allowable. Overfilling must be guarded against by constantly monitoring the liquid level. Road trucks must be electrically grounded to prevent static electricity buildup before the transfer operation is initiated. In addition, transfer lines must have electrical continuity across any connectors. Also, chock blocks should be used to prevent rolling of the vehicle.

Maintenance and Repair

All vehicles should undergo a formal written schedule of preventative maintenance as well as a daily routine check by the driver. Any defects noted during inspection or operation should be reported in writing to the driver's supervisor. Tanks should be repaired only by persons qualified in pressure work and must meet requirements of the authority having jurisdiction. Repaired containers cannot be returned to service until their use is certified by a qualified pressure vessel inspector.

An amendment to Part 177, effective January 1, 1979, is intended to reduce the probability of accident involving hazardous materials as a result of maintenance or repair work on the motor vehicle by specifying conditions for such work inside a building (Federal Register 1978e).

9.2.5 Release Control in LPG Transportation by Truck

Section 65 of NFPA 58 requires that LPG be removed from a vehicle and that the tank be purged before moving the vehicle into a public garage for parking or repair. If the garage is owned by the vehicle operator, the vehicle may be moved into the garage without removing the LPG if the provisions of Chapter 7, NFPA 58, are followed. Chapter 7 provides for proper construction ventilation, and heating of structures housing LP-gas systems. Construction requirements include use of non-combustible materials for floors, ceilings, and wall materials, and that the floor not be below grade. The structure shall be ventilated utilizing air inlets and outlets arranged to provide air movement across the floor as uniformly as possible, at specified rates, and outlets should discharge at least five feet away from any other opening into the garage or any other structure. Heating shall be by hot water or steam radiation or other heating transfer medium with the source remotely located. Accidents in garages have been essentially all cases where the LPG had not been removed (Lathrop and Walls 1974). It appears that adherence to the above regulations would prevent or control LPG releases in garages.

Accidents during the transfer of LPG can involve a broad range of human error, equipment failures and malfunctions. These will often involve releases at relatively small rates, with and without fire. The NLPGA has published bulletins which offer guidance on controlling such releases:

- "How to Control LP-Gas Leaks and Fires" (Bulletin 200-73)
- "How to Stage LP-Gas Fire Control Demonstrations" (Bulletin 201-62)
- "How to Handle Small LP-Gas Fires With Portable Fire Extinguishers" (Bulletin 204-76).

These bulletins describe basic precautions in approaching a leak or a fire, when and how to extinguish a fire, protection of a tank from radiation by water spray (to avoid a BLEVE), policing the area, etc. The bulletins are applicable generally to releases from trucks during loading or unloading, or from piping connecting the truck tank to storage, or from the storage tank.

If a truck tank has an accident on a highway which results in leakage, as contrasted to a major rupture and large spill, the same procedures referred to above in connection with releases in loading/unloading operations are applicable.

In any case of leakage or fire at or near a truck tank, there is the possibility of a BLEVE, as with rail tank cars or with stationary tanks, and the NFPA slide-cassette package on Transportation BLEVE's should be applicable and useful (NFPA Package SL-36).

9.2.6 Release Prevention and Control R&D in LPG Truck Transportation

Research and analysis of accident data and particular serious accidents have been conducted to identify causes and formulate prevention and control measures. Necessary data for this analysis are collected by several agencies.

Statistics on LPG truck accidents are gathered by the Department of Transportation/Office of Hazardous Materials, (DOT/OHM), the Federal Highway Administration (FHWA), the Bureau of Motor Carrier Safety (BMCS), agencies in several states, and private trucking organizations. However, the gathering of statistics on truck accidents is a major problem. Accident data are incomplete, sometimes difficult to obtain even when they exist, and are largely not dependable because different criteria are used in collecting data (Davis et al. 1977, Krasner 1970). A particularly significant problem with respect to hazardous materials is that many truck accident data are not differentiated by type of cargo, and there are many times more trucks hauling gasoline, diesel fuel, and other middle distillates than there are hauling LPG. The statistics available from various studies include the following:

- The NLPGA collected and analyzed LPG truck accidents for the period 1972-1976. The data include 178 highway accidents in the 6-year period, of which 96 involved no release of cargo, 50 produced minor spills, and 32 involved fire or significant releases (Drake et al. 1978).
- The DOT/OHM collected and analyzed data for the period 1971-1977, showing a total of 124 spills from all sources except loading/unloading operations (Drake et al. 1978).
- FHWA statistics, although quite incomplete in many categories, show that in the 5-year period 1973-1977 there was an annual average of 79 accidents involving LPG trucks, and an annual average of 9.4 were reported as spills (Drake et al. 1978).
- Factory Mutual analyzed DOT data for hazardous materials for the period July, 1966, to December, 1968, finding that combustible compressed gases accounted for 139 accidents, 9 deaths, and 81 injuries (Krasner 1970).
- There were three separate incidents in 1974 involving LPG tank trucks in repair garages. Fires resulted because the trucks were not emptied of LPG and purged before being moved inside. Several people were burned; there were 4 fatalities and large amounts of property damage (Lathrop and Wells et al. 1974).

In 1974, the Alberta Manpower and Labor Department investigated accidents that occurred in Alberta in the period 1969-1974 and which resulted in fatalities (Rees 1974). There had been a trend of increasing frequency of such accidents (reflected also in the entire Canadian propane industry), and this study was carried out in an effort to determine the causes. It was found that each accident was caused by a human error. In most cases both management and truck drivers appeared to be unaware of safety requirements, of the hazards of putting trucks containing propane in garages, and of the requirements for safe handling and transporting LP-gas. The study made several recommendations, among them that the industry provide for adequate training of employees, ensure proper maintenance of equipment, enforce safety practices, locate large storage tanks (2000 imperial gallons or more) at least 400 feet from all public and residential places, and prohibit the garaging of LP-gas trucks and tanks.

In 1977 the DOT/FHWA sponsored a project to assess accident data relating to cargo tank overturns and to determine whether the existing cargo tank specifications should be revised (Davis et al. 1977). The report lists these recommendations pertinent to MC-331 tanks:

- A systematic effort should be made to collect cargo tank damage information and related data.
- Any truck or tractor used to carry or tow a cargo tank containing hazardous flammable materials should be subject to special performance and compliance requirements. The complete fuel system should be designed for a high level of safety performance more consistent with the risks associated with leakage in possible accidents.
- The requirements for automatic shut-off and excess flow valves should be clarified and improved. Positive shut-off internal to the shell and establishment of sufficient compliance lists and procedures should be specified. Required research includes the development of cost-effective systems and components.
- The requirements for fitting protection should be reviewed and clarified. Any opening in the shell should be closed internally except when transferring cargo, or be adequately protected for most accidents which the basic shell can survive.
- The requirements for shear section performance, (as presently specified in 178.337-12), should be clarified and compliance tests developed.

- The requirements for supporting and anchoring the shell should be reviewed and specific requirements established for overturn conditions.
- Manholes should be located in an area of low probability of accident damage. The design should minimize stress concentration and maximize damage resistance. Further research and testing is required to establish definitive methods of specifying design requirements.
- Baffles, bulkhead, and stiffeners, when used in MC-331 tanks, should be designed to minimize their contribution to adjacent structural stress concentrations. Further research effort is required to define cost-effective methods of spreading out the connecting loads.

Some of these recommendations have been the subject of recent rulemaking.

9.2.7 Release Prevention in LPG Transportation by Rail

LPG is released accidentally from tank cars mostly as a result of train derailments; accidents have also occurred in railroad yards, e.g., in humping operations. The actual mechanisms of release in most cases are rupture of a tank car by impact with other cars, rails, or other objects during derailment, or rupture due to overheating of the tank because of impingement of flames from another already-ruptured tank. A third mechanism, that of delayed rupture, occurred in a derailment at Waverly, Tennessee, in February, 1978. Two LPG cars were dented and overturned, but not leaking. About two days later, while preparations were being made to transfer LPG from the derailed cars, one of them ruptured, the vapor ignited, and the fireball caused 16 deaths and more than 40 injuries (NTSB 1979). This rupture was caused by weakening of the tank wall in the dented area followed by an increase of pressure in the tank in the intervening two days due to moderation (i.e., increased temperature) of the weather. Representative railroad accident descriptions involving LPG are included in Appendix H.

The most serious consequence of LPG releases from railroad tank cars is the BLEVE (Boiling Liquid Expanding Vapor Explosion). A typical event sequence leading to a BLEVE is described below.

BLEVE Event Sequence

Frequently, when a car is punctured during derailment, the volume leak rate is limited; that is, the contents of the tank are released through a relatively small opening, such as would be made by impact of a coupler against a head. In these cases the LPG emerges from the tank with significant velocity because of the pressure in the tank. Due to numerous impacts during the derailment providing ignition sources, the leaking cargo is usually ignited and the flame may impinge on the shell of another tank nearby. If flames impinge on a tank below the liquid level, the liquid temperature will gradually rise, increasing the vapor pressure until eventually (perhaps in a few hours) the relief valve will release vapor. On the other hand, if the flames impinge on the tank above the liquid level, the temperature of the steel will increase rapidly because of poor heat transfer to the vapor, soon (within a fraction of an hour) weakening the steel so that the tank ruptures. In this latter case, the tank usually ruptures into two "tubs" which are accelerated by the LPG vapor pressure and rocketed distances between several and many hundred feet. These BLEVEs can cause great damage, both by impact of the flying pieces of tank car and by thermal radiation from the large fire that generally results.

In the period 1950-1969, NFPA received 18 reports on incidents involving LP gas tanks exposed to fire which resulted in BLEVEs; those 18 fires resulted in 22 deaths and 318 injuries. In the years 1970-1974, 12 BLEVE incidents were reported, resulting in 24 deaths and more than 300 injuries.

Design and Construction Practices

Release prevention is a principal philosophy in the design and construction of railroad tank cars. The railroads developed industry standards and specifications before federal regulations became mandatory. The principal industrial associations providing specifications and service standards for pressure tank cars are the Association of American Railroads (AAR) and the American National Standards Institute (ANSI). The AAR and the ANSI are referenced in the federal regulations, and, in fact, the AAR Committee on Tank Cars has an active role in reviewing and approving designs, materials and construction, conversion or alteration of railcar tanks.

The transportation of LPG by rail is covered by federal regulations under Title 49, in three parts (Government Printing Office 1980):

- Part 173 - Shippers - Federal Requirements for Shipments and Packagings.

  Rail transport of LPG is covered specifically in Subpart B, Preparation of Hazardous Materials for Transportation, paragraph 173.31, qualification, maintenance, and use of tank cars; also in Subpart G, Compressed Gases; Definition and Preparation, paragraph 173.314, Requirements for Compressed Gases in Tank Cars.
- Part 174 - Carriage by Rail.

  This part under various subparts covers General Requirements (Subpart A), such as inspections, handling of astray shipments, etc.; General Operating Requirements (Subpart B), such as shipping papers, orders and billings, labels and placards, etc.; Handling and Loading Requirements (Subpart C). Detailed Requirements for Gases (Subpart F), under paragraph 174.204, "Tank car delivery of gases," specifies that compressed gases may be unloaded only on private tracks, unless certain conditions are fulfilled.
- Part 179 - Specifications for Tank Cars.

Specifications 105, 106, 111, 112 and 114 apply to tank cars suitable for LPG transport. However, specification 106 and 111 tank cars are seldom used for LPG transport because of the low pressure rating of the 111 tank and the inconvenience of the multiple small tank 106 style cars. Each tank car must be designed, constructed, tested, and certified to a specific test pressure. In each case the permissible maximum operating pressure (at a specific temperature, e.g., 105°F) is 75% of the test pressure, as determined by the start-to-open setting of the safety relief valve. The rest pressure is incorporated in the tank car designation; e.g., 105T100 is a car designed to 105T specifications and tested to 100 psi. The particular tank car specification required will depend generally on the composition of the LPG, and specifically on its vapor pressure. Propane, for example, requires cars tested to higher pressure than butane. General design requirements of Part 179 include:

- Requirement for approval of the manner in which tanks are attached to the car structure; also of welding procedures, welders, and fabricators.
- Limitations on capacity and gross weight of tank cars built after November 30, 1979, to 34,500 gallons and 263,000 lb on the rails, respectively. Existing tank cars cannot be converted to exceed these limits.
- All tank cars built after 1970 must be equipped with interlocking automatic couplers that will resist car telescoping and jackknifing in derailments and in emergency stops. These couplers must be approved by the Federal Railroad Administration (FRA).

As a result of a series of serious accidents involving uninsulated tank cars built to 112 and 114 specifications and transporting hazardous materials (especially LPG), the DOT/MTB issued amendments to Parts 173 and 179, effective October 19, 1977 (Federal Register 1977).

- Existing and newly built specifications 112 and 114 tank cars used to transport flammable gasses (including LPG) are required to have both thermal and tank head protection. The thermal protection system must prevent the release of lading (except through the safety relief valve) when the car is subjected to a pool fire for 100 minutes or a torch fire for 30 minutes. The head shields (either separate or integral) were designed to reduce the chance of a head being punctured (e.g., by a coupler during a derailment).
- All 112 and 114 cars, regardless of lading transported, are to be equipped with special couplers designed to resist vertical disengagements. These are referred to as "shelf couplers."
- Three types of 112 and 114 tank cars were established:
  1. 112A and 114A cars are authorized to transport hazardous liquids (such as gasoline) and non-flammable compressed gases other than anhydrous ammonia (note that LPG is classified as "flammable gas").
  2. Newly built 112S and 114S tank cars to transport anhydrous ammonia as well as commodities carried by 112A and 114A. Each must be equipped with head shields and shelf couplers.

3. Newly built 112T, 112J, 114T, and 114J tank cars to transport flammable and non-flammable compressed gases, including anhydrous ammonia, and hazardous liquids. Each is required to be equipped with a thermal protection system, head shields, shelf couplers, and a safety relief valve meeting the requirements of paragraph 179.105-7. This last requirement permits reduction of relief valve capacity on thermally insulated cars.

The meanings of the letters after the specification numbers indicated above are as follows:

"A"	designated 112 or 114 specification cars which have as additional protection only shelf-couplers; these cars can carry non-flammable gases and hazardous liquids.
"S"	designated 112 or 114 specification cars which have both shelf-couplers and head shields; they can carry anhydrous ammonia, non-flammable gases, and hazardous liquids.
"J"	designated 112 or 114 specification cars with thermal protection enclosed in a metal jacket, plus shelf-couplers and head shields; these cars can carry flammable gases (LPG) in addition to products carried by "A" and "S" cars.
"T"	indicates thermal protection without a metal jacket; such cars are otherwise similar to "J" cars and can carry the same products.

The DOT/MTB has issued a list of thermal protection systems which may be used on 112 and 114 tank cars without further test verification to satisfy the retrofit requirements (Federal Register 1978c).

The total safety relief valve discharge capacity must be sufficient to prevent a pressure buildup in excess of 82.5% of tank test pressure, or 10 psig above start-to-discharge pressure, whichever is higher. Typically, the safety valve is set to begin relieving pressure when the internal tank pressure reaches about 280 psig.

All other piping and valve systems are also located within the manway cover dome, or bonnet, that covers the manhole. The manway opening is normally at least 10 inches in diameter. The manway cover gasket is made of asbestos. The tank car valves are mounted on the manway cover. The manway cover itself is 2.4 inches thick.

Tank car valves are not like the internal valves described for the tank trucks. They operate as excess flow valves only, and are located in the loading and unloading pipes that communicate with the interior of the car. These valves automatically close against an outward flow of LPG when an external valve is broken off. However, if the leak rate is insufficient, they will not be effective in stopping the leaks, since a certain minimum flowrate is required to activate the valves. The valves have gaskets made of asbestos or stainless steel.

Operations and Maintenance

Operating and maintenance aspects of rail transportation of LPG are covered by 49CFR173 and 49CFR174, as noted above. Numerous details of practice are also specified in NFPA 58 "Storage and Handling of Liquefied Gases - 1979 (NFPA 1979), and recommended practices have also been developed and promulgated by the National LP-Gas Association, and by the Compressed Gas Association (Reinhold Publishing Co. 1966).

Important operational aspects covered by regulations include inspections of tank cars and appurtenances before loading and before shipping; periodic retest and reinspection, including hydrostatic test of tanks and testing of safety relief valves; withdrawal from service or repair of tanks which have been subject to fire; filling densities, maximum vapor pressure (at specified temperatures) versus the test pressure; use of foreign tank cars in domestic use (tests and certifications required); respective responsibilities of shippers and carriers; proper handling of empty cars offered for transportation; handling of leaking cars; positions of cars carrying compressed flammable gases in trains with respect to cars carrying other hazardous substances.

9.2.8 Release Control in LPG Transportation by Rail

Control of LPG releases from tank cars after derailments consists primarily of applying water to tanks on which flames are impinging in order to prevent a BLEVE, and attempting to extinguish flames from leaking cars. This work must be done from appreciable distances because of the risk of exposing emergency response personnel to a BLEVE, and also because of the chance that a tank not leaking nor subject to flame impingement may be subject to delayed rupture because of damage incurred during the derailment.

The railroads use several techniques for providing information that is important in release control activities. First, Part 172 of the regulations requires that the shipper include the proper shipping name, the class and other describers on the shipping papers. Also, the train crew must have a document indicating the position in the train of each placarded car containing a hazardous material, and a member of the train crew must have a copy of the waybills of the hazardous materials being transported. However, MTB and FRA regulations do not require that railroads have an emergency response preparedness capability. Nevertheless, most major railroads have developed or are developing their own systems, such as accelerated use of the AAR's Bureau of Explosives Standard Transportation Commodity Codes, which are computerized and printed out for each train's consist. These printouts include emergency actions appropriate for each hazardous material on the train. Other techniques used by some railroads include coordination with local officials along rights-of-way, use of specially denoted hazardous materials trains, and special railroad hazardous materials teams. However, the NTSB reports, there have been informational difficulties such as 1) not obtaining prompt notification from railroads of an accident; 2) carrying only one copy of the shipping papers, which is inadequate; and 3) emergency manuals not located on trains (NTSB 1978c).

As mentioned previously, the NFPA has developed an audiovisual package, "Transportation BLEVE's -- Causes, Effects, Guidelines," to provide release control training. This package consists of 140 color slides, a cassette tape, an instructor's guide, and a student workbook (NFPA Package SL-36).

9.2.9 Release Prevention and Control R&D in LPG Rail Transportation

Following a number of tank car ruptures in 1969, a research project was initiated in 1970, referred to as the RPI-AAR project (Railway Progress Institute-Association of American Railroads) and known officially as "Railroad Tank Car Safety Research and Test Project." In addition to the two organizations already named, sponsors included nine other interested associations and institutes; there were also government observers from the Canadian Transport Commission, the FRA, and NTSB. The program was directed by the Tank Car Research Committee under the Chairmanship of Frank J. Heller. The research was organized in phases, and, except for one continuing project (1978), the research effort has been concluded (Heller 1978; Everett and Phillips 1972). The research phases were as follows (RPI-AAR Project 1979):

Phase 1 - Accident Review
Phase 2 - Derailment Environment Study (including dollar loss)
Phase 3 - Materials Study-Steel (including fracture properties)
Phase 4 - Review of Literature and Related Experience
Phase 5 - Head Study (including head shields)
Phase 6 - Safety Valve in Liquid Study
Phase 7 - Safety Relief Devices-General
Phase 8 - Reduced Scale Model Studies (including derailment studies)
Phase 9 - Design Study-Tanks and Attachments
Phase 10 - Design Study-Car (including couplers and trucks)
Phase 11 - Thermal Effects Study (including thermal shield systems, and torch tests)
Phase 12 - Vessel Failure Research (including fracture and tub rocketing)
Phase 13 - Tank Head Shield Design (this phase conducted under DOT Contract, DOT-FR-00035)
Phase 14 - Stub Sill Buckling Study
Phase 15 - Switchyard Impact Tests.

Some 40 final or phase reports, plus 31 technical progress reports, were issued on the various phases of this project as of February, 1979 (RPI-AAR Project 1979). This program apparently was directed entirely toward release prevention, and has obviously provided the basis, at least in part, for the regulatory actions requiring thermal protection, shelf couplers, and head shields.

The DOT sponsored a 4-month study in 1970 at the Cornell Aeronautical Laboratory (Bullerdiek et al. 1970). The objectives of this research included development of performance specifications and conceptual design and application requirements for safety devices to prevent catastrophic car failure. A number of technical conclusions were reached, and a plan for a research program was developed. Apparently this was the basis in part for initiation of the RPI-AAR project described above, although the latter program was appreciably more comprehensive than envisioned by this study.

The NTSB held a public hearing in April 1978 and issued a report of their analysis and recommendations regarding means of reducing the risks from rail transport of hazardous cargo, including LPG (NTSB 1978c). The NTSB examined safeguard installations for 112 and 114 tank cars, emergency notification and response procedures, the derailment problem, track standards, and other areas. Conclusions by NTSB pertaining to release prevention include the following:

- 112A and 114A tank cars were designed by the tank car and the railroad industry to maximize economics; no specific safety methodology to determine unreasonable risk to the public is employed.
- When the 112A/114A car designs were accepted on special permit, the safety features of thermal insulation and center sills (still found on most DOT 105 cars) were eliminated, and the capacity was increased threefold. No analysis nor full-scale testing was carried out before these designs and equipment were placed into service.
- No adequate safety methodology has been developed by federal agencies to determine risks as a basis for regulation.
- DOT's accident data collection programs for train derailments and hazardous materials provide a limited capability for accident prevention research and countermeasure development.

9.2.10 Release Prevention and Control in LPG Ships and Barges

Release prevention and control in ships and barges that transport LPG are accomplished by combining good vessel design and construction with safe operating practices. A concise summary of current practices and requirements is contained in the U.S. Coast Guard (USCG) publication "Liquefied Natural Gas and Liquefied Petroleum Gas, Views and Practices, Policy and Safety" (USCG 1980).

Ships

The Coast Guard is responsible for the safety of all U.S. flag liquefied gas ships everywhere and all foreign flag liquefied gas ships while in U.S. waters. This responsibility has led to the development of standards for ship design, construction, alteration, repair, maintenance, and operation. The statutory authority for regulation of liquefied gas ships is the United States Code, specifically the Port and Tanker Safety Act of 1978 (46USC391a). The regulations are detailed in the Code of Federal Regulations (CFR). For the design and construction of new LPG carriers, the applicable regulations are found in Title 46 of the Code of Federal Regulations, Part 154 (46CFR154), "Self-Propelled Vessels Carrying Bulk Liquefied Gases." These rules implement virtually all of the Inter-Governmental Maritime Consultative Organization (IMCO) "Code for the Construction and Equipment of Ships Carrying Liquefied Bases in Bulk," Resolution A.328.(IX), referred to as the IMCO Gas Code. Existing LPG carriers are regulated under standards in 46CFR38 and 46CFR154a. All foreign flag vessels entering U.S. waters must meet the same standards as U.S. flag vessels for the cargo containment portion of the ship.

Ship Design and Construction

Ship design is intended to prevent any cargo release whenever possible and to minimize the quantity released when prevention is not achieved. Also tank failures during normal operations are to be prevented. This has led to development of better tank designs that have built-in redundancy, tanks that leak before they fail, or tanks that, through conservative design, can be shown not to fail in ordinary service. Many types of cargo tank systems, including integral tanks, membrane tanks, semi-membrane tanks and independent tanks, are acceptable as long as the level of safety provided is equivalent to, or exceeds, existing requirements. This safety concept is followed throughout the design of a vessel. For example, in minor collisions and groundings, cargo tanks should retain their integrity. For this reason, the tank is set some distance from the hull of the ship. Also it is possible to design a ship to limit damage to the portion of the vessel initially damaged. Passive and active fire protection systems, for example, should be designed to prevent overheating of a tank adjacent to a burning one.

Table 9.1 provides a brief overview of the design, construction, and equipment requirements for LPG carriers (USCG 1980).

Ship Certification

Ship certification requirements are summarized in the reference USCG 1980.

> "Before any U.S. flag vessel operates anywhere in the world, it must undergo a certification procedure to ensure that it meets all applicable design and construction requirements. Both U.S. and foreign flag vessels must meet essentially the same requirements for the cargo containment portion of the vessel. U.S. flag vessels are reviewed by the Coast Guard for compliance in such areas external to the cargo containment system as the design and construction of propulsion and steering equipment, auxiliary systems, accommodations arrangements, lifesaving equipment, and electrical systems.
>
> "Foreign flag vessels are required to meet similar requirements which are found in the Safety of Life at Sea (SOLAS) Convention. Both a review of the ship's plans and a complete inspection are needed before a U.S. flag vessel is certificated and allowed to carry cargo. Thereafter, biennial inspections are required to maintain a valid Certificate of Inspection. For foreign flag vessels, a Letter of Compliance (LOC) is required. Biennial examination of foreign flag vessels is necessary for the renewal of an LOC. Additionally, modifications to the cargo containment system, changes of vessel ownership, or changes of vessel registry invalidate the Letter of Compliance and must be reported to the Coast Guard. Upon completion of a satisfactory reexamination a revised Letter of Compliance is issued."

The following is a listing of design and construction features which would be examined during the certification process for both U.S. and foreign flag carriers (USCG 1980):

- design and arrangement of cargo tanks and cargo piping and vent systems, including the suitability of the cargo containment system materials for the pressure and temperature involved, welder and welding procedure qualification, and nondestructive testing of the cargo tanks and piping
- arrangements and adequacy of installed fire extinguishing system and equipment and structural fire protection
- safety devices and related systems which check the cargo and the surrounding spaces to give warning of leaks or other disorders which could result in a casualty

TABLE 9.1. Overview of Design, Construction, and Equipment Requirements of LPG Carrier

1. Hull Structure - includes requirements for:
    a. Heat transfer calculations that determine the minimum operating temperatures of the contiguous hull structure under the most severe design conditions in order to specify the hull steels suitable for those minimum operating temperatures
    b. The certification that the materials used in the contiguous hull structure meet the specifications required
    c. Weld procedure qualification and weld production testing of contiguous hull structure weldments
    d. Crack arresting steels in the outer hull structure

2. Ship Stability, Survival Capability, and Cargo Tank Location - includes requirements for:
    a. Freeboard and stability
    b. Survival capability after damage and flooding
    c. The location of the cargo tanks with respect to their minimum allowable distance from the outer hull of the vessel

3. Vessel Arrangements - include requirements for:
    a. Segregation of the cargo area from accommodations, service, control, and machinery spaces, chain lockers, stores, and drinking and domestic water tanks
    b. Access to gas-safe and gas-dangerous spaces
    c. Cargo control rooms
    d. Airlocks
    e. Bilge and ballast systems
    f. Special loading arrangements

4. Cargo Containment Systems - includes requirements for:
    a. Membrane, Semi-membrane, and Independent (Types A,B, and C) tanks
    b. Design loads, structural analysis, and allowable stresses for each tank type
    c. Tank supports
    d. Secondary barriers, which for all tanks other than type C tanks are designed to contain the cargo leakage from any envisioned failure of the primary barrier
    e. Tank insulation which prevents excessive cooling of the hull structure and limits cargo boil-off
    f. Construction and non-destructive testing of cargo tanks

5. Cargo and Process Piping Systems - includes requirements for:
    a. Design analysis and allowable stress
    b. Materials of construction
    c. Piping components
    d. Piping fabrication and joining details
    e. Non-destructive and functional testing of piping systems
    f. Remote control piping system

6. Material and Fabrication - includes requirements for:
    a. Materials and their properties for low temperature service including minimum standards for chemical analyses and toughness testing of tank and piping material
    b. Welding procedure tests, weld production tests and non-destructive testing

7. Cargo Pressure and TemperatureControl - includes requirements for:
    a. Refrigeration systems
    b. LNG boil-off burning systems

8. Cargo Venting Systems - includes requirements for:
    a. Pressure relief systems for cargo tanks and interbarrier spaces
    b. Pressure relief valves
    c. Vent piping and vent masts
    d. Vacuum protection systems
    e. Pressure relief valve sizing

9. Atmospheric Control in the Cargo Containment System - includes requirements for:
    a. Inert gas systems
    b. Environmental control of cargo tanks, piping, and hold spaces

10. Electrical Installations - includes requirements for all electrical equipment installed within the hazardous area

11. Fire Protection - includes requirements for:
    a. Structural fire protection
    b. Fire main systems
    c. Waterspray system protection of deckhouses, tank domes, exposed parts of tanks, exposed on deck storage vessels for flammable cargoes, cargo manifolds and their control valves, and boundaries of deckhouses facing the cargo area
    d. Dry chemical firefighting systems
    e. Smothering systems installed in enclosed spaces

12. Mechanical Ventilation within the Cargo Area - includes requirements for:
    a. Spaces that must be ventilated
    b. Amount of ventilation that must be provided
    c. Ventilation fans and their driving motors

13. Instrumentation - includes requirements for:
    a. Tank level indicators and alarms
    b. Tank pressure gauges and alarms
    c. Temperature indicating devices
    d. Gas detection systems

14. Personnel protection - In addition to the requirements in the 1974 Safety of Life at Sea (SOLAS) Convention, includes requirements for:
    a. Protective clothing
    b. Self-contained breathing apparatus
    c. Rescue lines
    d. Explosion proof lamps
    e. Compressed air
    f. Medical and first aid equipment
    g. Portable gas and oxygen detecting equipment

Source: (U.S. Coast Guard, 1980)

- compatibility of the cargo with materials of construction
- hull steel
- cargo pressure and temperature control
- segregation of the cargo spaces from the machinery working spaces and accommodation spaces
- suitability of electrical equipment installed in hazardous areas.

Foreign flag vessels must meet requirements in the following four additional areas where U.S. regulations are more stringent than the IMCO Gas Code (USCG 1980):

- specification of design-ambient temperatures
- requirement of enhanced grades of steel for crack arresting purposes in the deck stringer, sheer strake, and bilge strake
- specification of higher allowable stress factors for independent tanks (Type B and C)
- prohibiting the use of cargo venting as a means of cargo pressure/ temperature control.

Crew Qualifications and Training

The Coast Guard's traditional responsibility of protecting lives and property at sea includes establishment of the qualifications for licensing and certification of merchant marine personnel serving aboard U.S. merchant vessels.

In discussions held in 1975 and 1976, both the Coast Guard and vessel operators agreed that specialized training would be required for the personnel manning liquefied gas vessels. This training would include firefighting, the management of cargo hazards and cargo handling systems, and procedures pertaining to the carriage of liquefied gases.

The USCG has summarized the qualification and training requirements as follows (USCG 1980):

> "Although the United States pioneered LNG and LPG technology, most shipping experience has been gained by foreign flag vessels on voyages not subject to Coast Guard jurisdiction. In the United States there has been a greater emphasis on crew training by formal schooling than by on-the-job training, as practiced in most cases

by foreign flag crews. Nevertheless, the U.S. Coast Guard determined that standards for manning and personnel qualification for U.S. flag LNG and LPG vessels will be applied to foreign flag LNG and LPG vessels operating in U.S. waters.

"Present personnel qualification standards require that the Master, Chief Mate, Chief Engineer, and First Assistant Engineer have satisfactorily completed a course of instruction in the carriage and transfer of liquefied gases as well as firefighting. A company initiated letter indicating the completion of an administration approved training course will serve as evidence of this training.

"Further, the qualification standards require that all officers and ratings with specific duties involving cargo and cargo equipment complete a course of instruction, or onboard training, responsive to the duties performed. All crew members must also have training, or onboard instruction, in the special hazards of LNG and LPG and an awareness of the general safety features of the vessel.

"The Coast Guard Captain of the Port (COTP) will evaluate the certification and training received by officers and crew of foreign flag LNG and LPG vessels. This evaluation will use the Inter-Governmental Maritime Consultative Organization (IMCO) Conference on Standards of Training, Certification and Watchkeeping for Seafarers, 1978, Regulation V/3 and Resolution 12 as a measure of equivalency to U.S. standards."

Ship Operations

The Coast Guard is also responsible for vessel traffic control and safety standards for waterfront facilities to assure protection against fire, explosion, natural disasters and other serious accidents or casualties.

Because of wide variations in harbor characteristics and facility development, each local Captain of the Port (COTP) is responsible for establishing and maintaining safe operating parameters for his port area. The COTP has authority to establish, at his discretion, any special safety measure to be observed by operators of both LPG vessels and waterfront facilities. Operational requirements for transport and handling of LPG in a port are established by the COTP and are set forth in an Operations Plan. The COTP Operations Plan is a compilation of the facility safe operating procedures and additional requirements deemed appropriate by the COTP. As an example, the COTP may:

1. Require the vessel, on entry to the port, to anchor and to be examined prior to permitting transit of the port area.

2. Require the vessel to be escorted by the Coast Guard. This can be accomplished with either Coast Guard craft or by placing a Coast Guard officer or petty officer aboard another designated escort vessel.
3. Require a minimum number of tugs to be in attendance while the vessel is in transit.
4. Establish a fixed safety zone at the facility and a sliding safety zone around the vessel during transit.
5. Restrict vessel entry and movement to periods when there is good visibility.
6. Require communications to be constantly maintained between the vessel and all escort vessels, including towing vessels.
7. Require that exact transit times be established.
8. Require a report prior to entry that all hazardous material cargo alarms, safety devices, and shut-downs have been tested by ship's personnel and are in good working order.
9. Require a pre-transfer conference between ship and terminal personnel.
10. Require effective communications for cargo transfer.
11. Require tests of terminal alarm devices and emergency shut-down prior to commencing transfer operations.
12. Prohibit other cargo operations during transfer of LPG.
13. Prohibit loading of stores or other activity during transfer of LPG.
14. Prohibit welding, burning, or hotwork while vessel is moored at the terminal.
15. Prohibit bunkering during cargo transfer operations.
16. Require a cryogenic supervisor to be in attendance during cargo transfer operations.
17. Require the vessel to rig towing pendants overside while at the terminal.
18. Restrict operations to daylight only (or nighttime only).

This list is not all-inclusive, but illustrates some possible operational constraints imposed in order to enhance port safety.

In conjunction with the LPG Operations Plan, the COTP will develop an LPG Emergency Contingency Plan for the port area. The Emergency Contingency

Plan is intended to be a guide in handling various emergency situations. The plan will provide for a coordinated approach to an LPG emergency. Preplanning with the maritime industry and with other federal, state, and local agencies will result in a comprehensive response organization with a complement of materials, equipment, and training needed to deal with such emergencies. The plans will generally include a listing of resources and personnel, as well as preplanned courses of action. Emergency scenarios that may be preplanned include:

1. collision involving an LPG vessel
2. fire aboard an LPG vessel
3. grounding of an LPG vessel
4. primary containment failure on an LPG vessel
5. release of LPG aboard the vessel
6. release of LPG at the facility
7. fire at the facility.

Coast Guard regulations regarding LPG facility personnel requirements are contained in 33 CFR Part 126 and specifically refer to the "person in charge" of the shoreside transfer operations. The "person in charge" is defined as the assigned representative of the facility owner or operator. Detailed responsibilities are placed upon this individual to assure the safe transfer of cargo between vessel and facility.

Definite training and qualification requirements are not given in these regulations. It is, however, specified that the "person in charge" of the shoreside transfer operation must be trained in, and capable of performing competently, the necessary operations related to the transfer of the specific cargo.

Local COTP Operations Plans have requirements for facility personnel training and qualifications. While these requirements vary due to the differences in facility operations, basic training in cargo characteristics, safety measures, and firefighting are universally required.

Barges

Barges carrying liquefied flammable gases which include propane and butane are subject to United States Coast Guard regulations specifying design requirements and requiring inspection during and after construction. The regulations provide for integral and independent cargo tanks of gravity and pressure types. An integral tank is defined as a cargo containment envelope which forms a part of the vessel's hull in which it is built, and may be stressed in the same manner and by the same loads which stress the contiguous hull structure. An independent tank is defined as a cargo containment envelope which is not a contiguous part of the hull structure. An independent tank is built and installed so as to eliminate, wherever possible (or, in any event, to minimize) its stressing as a result of stressing or motion of the adjacent hull structure. In general, therefore, motion of parts of the tank relative to the adjacent hull structure is possible. An independent tank is not essential to the structural completeness of its carrying vessel's hull.

Barge operations are relatively simple in nature. LPG is transferred from the barge to storage tanks or a storage cavern where it is held for eventual redistribution. Barges seldom have pumps on board, and product transfer is usually accomplished by reducing barge tank pressure and/or pumping to load and increasing tank pressure to unload.

The pressure vessel design for barge tanks is similar to that used for railroad tanks, truck tanks, and pressurized land storage tanks, all of which have a history of safe operation.

In general barge transport of LPG has a history of safe operation. Therefore, the inspection and regulation of barge traffic are not subjected to the intense effort given to tankers. Barge operations are performed on a smaller scale than tanker operations, but at much higher frequencies. The systems are generally well-engineered and have good safety records (Welker 1980).

9.2.11 Release Prevention and Control R&D for LPG in Marine Transportation

A recently completed report by Martinsen et al. (1980) sponsored by the DOE Program, provides a substantial review of release prevention and control knowledge and risks associated with the marine transportation of LPG. This report contains a summary description of LPG vessel design, operations and fire protection systems. The fire protection philosophy for LPG ships includes fire prevention, control and extinguishment and damage potential reduction. If a fire occurs, it may be controlled, extinguished or allowed to burn out. Only small fires can be extinguished reliably, a dry chemical being the best agent for most situations. Fires may be controlled using high-expansion foams. Generally, the LPG fires should be extinguished only after the spill is finished and the flow of products stopped to avoid the possibility of reignition. Water sprays may be used for cooling as a damage control measure; however, water cannot be used on LPG fires directly because it increases the fire size by increasing boiloff rates.

Fault-tree analysis was used to estimate the probability of LPG releases. The goal of this analysis was to estimate the probability of events that either directly, or as the initiation of a larger event, might endanger the public or operators. The highest risk appears to be when the vessel is in port or transiting an inland waterway. The major emphasis of this analysis is on dockside operation when a tankship is unloaded. Results of the fault-tree analysis, showing spill probability per transfer versus spill size in gallons, are presented for both tanker and barge transfer operations. The spill probabilities for tankers and barges are about the same, a conclusion which is generally supported by other estimates found in the literature. Even when the spill probability is large, it does not necessarily follow that the operation is especially dangerous. Small spills, less than 10 gallons in size, which occur in more than half of all transfers, appear to have very low ignition probabilities. Spills involving more than 100,000 gallons of LPG are essentially certain to be ignited. Ignition probabilities versus spill size are calculated in the absence of collision events. The ship collision contribution is omitted because it makes the fire probability estimates appear higher than they really are at the terminal.

Fire damage and personnel injuries are related to fire size. Fires resulting from spills of less than 10 gallons are unlikely to cause substantial damage or result in fatalities. Fires involving thousands of gallons are nearly certain to cause substantial damage and are quite likely to result in fatalities and serious injuries. The overall probability of an operator fatality is estimated to be in the range of 10^{-6} to 10^{-5} per transfer operation. A fatality probability for the general public is much lower because of the great separation from the transfer area. The BLEVE phenomenon could possibly occur in barge operations because the cargo is pressurized. However, no fires at barge transfer facilities that resulted in BLEVEs have been recorded. BLEVEs cannot occur during operations with fully refrigerated LPG cargoes because all operations involve a liquid that is saturated near atmospheric pressure.

9.3 RELEASE PREVENTION AND CONTROL IN LPG CONSUMER STORAGE

As described in Section 7, the LPG industry serves about 18 million customers. In total, roughly 60 million people are dependent on LPG for one or more uses. Accidents with LPG at the consumer level involve a large fraction caused by human error. The emphasis of release prevention and control is to assure that containers and fittings function reliably without constant expert care and that chances are minimized for human error initiating or aggravating LPG releases.

9.3.1 Release Prevention in LPG Consumer Storage

The design and construction of LPG containers is covered by Chapters 2 and 3 of NFPA 58 (NFPA 1980). Chapter 2 includes the basic provisions for individual components, subassemblies, container assemblies or complete container systems. The field assembly and installation of components, subassemblies, container assemblies, or complete container systems into complete LP-gas systems is covered by Chapter 3. Paragraph 2101 of NFPA 58 says that containers shall be designed, fabricated, tested, and marked in accordance with DOT regulations (Government Printing Office 1980), the ASME Code (ASME 1974), or the API-ASME Code. Construction of containers to the last named code has not been authorized since July 1, 1961. Paragraph 2101 also states that containers fabricated according to earlier editions of any of these codes may continue in use, provided certain requirements described in other paragraphs are met, especially requirements pertaining to requalification or retesting.

The DOT Regulations pertaining to LPG containers are found in Parts 173 and 178 of Title 49, Code of Federal Regulations as follows:

- Part 173 - General Requirements for Shipments and Packagings, particularly in Subpart B - Preparation of Hazardous Materials for Transportation, and in Subpart G - Compressed Gases; Definition and Preparation
- Part 178 - Shipping Container Specifications, particularly in Subpart C - Specifications for Cylinders.

The DOT Regulations govern only containers which are to be used to transport LPG. A container which is to be mounted permanently at a consumer's site, or a skid-mounted container which may be moved when essentially empty, is not covered by DOT Regulations.

The codes are quite voluminous, covering in considerable detail many aspects of design, materials, fabrication methods, testing and retesting, marking, and installation of cylinders, tanks, and their appurtenances. Some aspects are highlighted below.

Design or Service Pressure. The term "service pressure" in DOT terminology designates the authorized pressure marking on the container, and is equal to the maximum allowed vapor pressure in the container at 70°F. The DOT further specifies in paragraph 173.301 that the pressure in a compressed gas (LPG) container at 130°F shall not exceed 5/4 times the service pressure. "Design pressure" in ASME terminology connotes design for the most severe condition of coincident pressure and temperature expected in normal operation. ASME containers for LPG must be designed to a minimum design pressure generally equal to 5/4 of the expected vapor pressure in the container at 100°F (containers for automotive use require a higher factor). See paragraph 2111, NFPA 58.

Design. In addition to specification of a design pressure or service pressure as explained above, the codes also contain requirements on design temperature, on wall thickness of the shell, thickness of the heads, concavity of heads toward the pressure, allowable stresses and methods of calculating stresses. On openings in cylinders the DOT (Paragraph 178.37-12) deals only with the types of threads to be used and some details as to cleanness, lack of defects, number of engaged threads, etc.; the ASME treats the design of openings in much greater detail, including the possible need for reinforcements, the geometric limits of reinforcement, the strength of reinforcements, and methods of attachment of pipe and nozzle necks to vessel walls.

Fabrication. Fabrication is regulated in part by stipulation of type of container (e.g., DOT, seamless cylinder), further by requirements on removal of dirt and scale, on rejection of defective material, repair of defects, etc; also, by stipulation as to when and where welding or brazing are permitted, and as to proper heat treatment of completed vessels. Again, the ASME Code is much more detailed than DOT Regulations, treating such items as permissible out-of-roundness of shell, tolerances for formed heads, details of lugs, fittings attachments, holes for screw stays, etc.

Inspection and Tests. Tests required under specified circumstances include hydrostatic tests of completed containers; tests to measure mechanical, impact, and ductility properties of the materials; leakage tests. Inspections must be made by an independent inspection agency (i.e., an agency not under control of the manufacturer). Required inspections include material inspections, verifications of chemical analyses, verification of compliance of containers with all requirements, inspection of interior before closing, verification of heat treatment, witness of all tests, verification of threads by gauge, reporting of volumetric capacity and tare weight and minimum thickness of wall.

Appurtenances. Both ASME and DOT require positive shut-off valves, internal excess-flow valves, external excess-flow valves, and back-flow check valves on containers, depending on the type, size, and mode of use of the container. External appurtenances must also be protected from physical damage while in transit, storage, when being moved into position for use, and when in use except in residential and commercial installations.

In addition to the appurtenances mentioned above, other devices must be or may be used:

- Liquid level gaging devices must be provided on all containers filled by volume.
- Pressure gages, if used, shall be attached directly to the container opening or to a valve or fitting which is so attached.

Pressure Relief Valves. ASME requires that containers be equipped with one or more safety relief devices designed to relieve vapor. These devices shall be spring-loaded valves set to start-to-discharge at pressures related to the design pressure as below:

	Minimum	Maximum
For ASME codes prior to 1949, and 1949, paragraph U-68 and paragraph U-69	110%	125%
For ASME code 1949 paragraph U-200 and paragraph U-201, and codes after 1949	88%	100%

Hydrostatic Relief Valves. Hydrostatic relief valves are required to relieve pressure in sections of liquid piping which can be completely closed off by valves, and shall have pressure settings not less than 400 psig or more than 500 psig unless installed in systems designed to operate above 350 psig. Hydrostatic relief valves for use in systems designed to operate above 350 psig shall have settings not less than 110% or more than 125% of the system design pressure.

Regulators. Requirements for regulators are described in paragraphs 247 and 314 of NFPA 58. Final-stage regulators shall be equipped on the low pressure side with either a relief valve or a shut-off device that shuts the gas off at the inlet side when the downstream pressure reaches certain limits, or both. The minimum and maximum limits for these devices are given in Table 9.2 reproduced from NFPA 58. The shut-off device just mentioned shall not open automatically to permit flow of gas. Regulators used to control distribution or utilization pressure shall be as close to the container as possible. Generally, first stage regulating equipment must be outside of buildings. LPG vapor may not be piped into buildings at pressures exceeding 20 psig, except under certain prescribed conditions. On regulators installed inside of buildings, the discharge from the safety relief device and from above the regulator and relief valve diaphragms shall be vented to outside air in such a way as to minimize significant diffusion of the vapor back into the building (see paragraph 315, NFPA 58).

TABLE 9.2. Limits for Devices on Final Regulators

Regulator Delivery Pressure in psig	Relief Valve Start-to-Discharge Pressure Setting, % of Regulator Delivery Pressure	
	Minimum	Maximum
1 or less	170%	300%
Above 1, not over 3	140%	250%
Above 3	125%	250%

LP-gas systems must be installed according to the requirements in Chapter 3 of NFPA 58, and in accordance with other national standards which may apply. Paragraph 300 lists several such codes:

1. National Fuel Gas Code, NFPA 54 (ANSI Z 223.1)
2. Stationary Combustion Engines and Gas Turbines, NFPA 37
3. Mobile Home Parks, NFPA 501A (ANSI A 119.3)
4. Mobile Homes, NFPA 501B (ANSI A 119.1)
5. Recreational Vehicles, NFPA 501C (ANSI A 119.2)
6. Removal of Smoke and Grease-Laden Vapors from Commercial Cooking Equipment, NFPA 96
7. Ovens and Furnaces, NFPA 86 A
8. Incinerators and Rubbish Handling, NFPA 82
9. Motor Craft (Pleasure and Commercial), NFPA 302
10. Grain Elevators and Bulk Handling, NFPA 61B (Grain Dryers).

Chapter 3 deals with 1) location of containers, 2) installation of containers and regulators, 3) piping system service limitations, 4) installation of pipe, tubing, fittings, valves, hose, and hydrostatic relief valves, 5) testing piping systems, 6) equipment installation, 7) distribution and industrial LP gas systems, 8) LP gas systems in buildings or on the roofs of buildings, 9) installation of appliances, 10) ignition source control, 11) installation of LP-Gas systems on vehicles, and 12) fire protection.

Operational Aspects. Experience has shown that a large fraction of leaks and spills of LPG have involved improper filling procedures, or equipment, and also defective equipment including hoses and cylinders. Safety involves several aspects: 1) existence of codes or regulations prescribing important items of equipment and procedures, 2) development of detailed procedures by the industry or by individual dealers to implement the codes and regulations, and 3) education and training of employees and of consumers in safe handling of LPG.

Chapter 4 of NFPA 58 titled, "LP-Gas Liquid Transfer" (NFPA 1979) covers many aspects of transfer, including especially the following:

Locations and Precautions. Liquid shall be transferred into containers only outdoors or in structures especially designed for the purpose; tank trucks unloading into storage containers shall be at least 10 feet from the container and positioned so that shut-off valves on both truck and container are readily accessible.

Personnel. Transfer operations shall be conducted by personnel trained in proper handling and operating procedures, and at least one such person shall remain at or near the operation from the time connections are made until lines are disconnected.

Containers to be Filled. Containers shall be filled only by the owner or upon his authorization; containers shall be filled only after determination that they comply with the design, fabrication, inspection, marking and requalification provisions of NFPA 58; disposable containers shall not be refilled; containers shall comply with requirements for service or design pressure with respect to the vapor pressure of the LPG to be transferred.

Venting. LP-gas, either liquid or vapor, normally shall not be vented to atmosphere except under specified conditions.

General Arrangement and Operation of Transfer Systems. It is recommended that transfer hoses be fitted with a shut-off valve at the discharge end so that the hose normally contains liquid.

Control of Ignition Sources. Internal combustion engines within 15 feet of the point of transfer shall be shut down during transfer operation, and other ignition sources such as smoking, open flames, portable electric tools, shall not be permitted.

Quantity of LP Gas in Containers. In order that containers will not become liquid-full at elevated temperatures due to thermal expansion of the liquid, NFPA 58 provides bases for determining container capacity, specifies maximum quantity to be put into a container and methods for verifying this quantity. Similarly, Chapter 2 of NFPA 58, titled "LP-Gas Equipment and Appliances," stipulates requirements for containers; container appurtenances; piping and hose, fittings and valves; and other equipment such as pumps, compressors, vaporizers, etc. Requalification requirements for containers are specified in paragraph 2101 and paragraph 2102 and in Appendix B of NFPA 58. Hence, it appears that both hardware and major aspects of procedures are prescribed by NFPA 58.

The LPG industry has been involved in a continuous effort to promote safe practices by the development and publication of guidelines, detailed procedures, and recommended practices. The NLPGA has published an "LP-Gas Safety Handbook," (NLPGA 1981) which contains some 73 individual bulletins in loose-leaf notebook form, with titles identified in the following categories:

- Distribution Operations
- Emergency Procedures
- Safety Meetings
- Residential
- Agricultural
- Industrial
- Recreational
- General.

Selected titles which indicate the nature of these bulletins are:

- "Safe Practices Around LP-Gas Installations"
- "Plant Inspection Check List"
- "Safety Considerations in Truck Deliveries"
- "Recommended Procedures for Filling Cylinders"
- "How to Control LP-Gas Leaks and Fires"
- "Safe Use of LP-Gas in Industrial Trucks"
- "Safe Camping with LP-Gas on Recreational Vehicles."

Other efforts at educating LPG consumers and employees of LP-gas distributors include publication of articles in trade magazines (e.g., LP-Gas, Butane-Propane News and Gas Industries) on pertinent topics, including accidents which have happened, specific procedures, discussions by experts on hazards and proper procedures. Examples of the latter include such titles as:

- "Colorado Fire Chief Says Dealers Face New Threats to LPG Facilities," LP-Gas, October, 1976, p. 31.
- "Check Lists for Truck Maintenance," Gas Industries, September, 1976, p. 8.
- "The Flaming Inferno," a story of an LPG fire at a distribution facility in Dallas, Texas, January, 1975; "How to Handle an LP-Gas Emergency;" "What Price Tragedy," description of an LPG truck accident in West Virginia and how it might have been prevented; and "The Decatur Disaster," the story of a jumbo LPG rail car explosion at Decatur, Illinois, July 19, 1974. All of these articles in LP-Gas, October, 1975.

Safety information is also provided by numerous other organizations, such as:

- A manual on safety, "LP-Gas, Safe Handling and Use," published by Engineering and Safety Service, American Insurance Association, 85 John Street, New York, New York 10038, 1972, 64 p.
- A film, "BLEVE," produced by and available from NFPA.
- A film, "Handling LP-Gas Emergencies," also from NFPA and available also in slide version.
- A slide series, "LP-Gas Explosion, Kingman, Arizona," July 5, 1973, from NFPA.
- An "RV (Recreational Vehicle) Owners Manual Copybook," a guide for RV manufacturers in making up future owner's manuals, covering electric, LP-gas, and other systems published by Recreational Vehicle Industry Association, PO Box 204, Chantilly, Virginia 22021.

9.3.2 Release Control in Consumer Storage

Attention here is restricted to that part of a consumer's LPG system containing liquid; that is, releases from appliances or from the vapor piping beyond the regulator are excluded from consideration. Causes of LPG spills from consumer storage cover a broad spectrum of possibilities. A large fraction are caused by human error, resulting from insufficient awareness or knowledge of the hazards and behavior of LPG and a lack of communication of safety procedures. Other causes include corrosion of buried tanks or liquid lines, settlement of foundations, nonstandard equipment, installations not according to code, overfilling, fires of other materials impinging on tanks or cylinders, damage to LPG systems by vehicles or by attempts to move or repair tanks containing liquid. Still other causes include leakage during LPG transfer, the use of a propane cylinder to pressurize a paint sprayer inside a building, failure of pressure regulators, leakage through heat exchangers in butane systems, failure of plumbing on vehicles and improper handling of cylinders and portable tanks.

Present methods of control of releases of LPG from consumer storage are generally very similar to those already discussed for truck transportation (Subsection 9.2.2). Probably the first action to be taken, at least for relatively small leaks, is to report the suspected leak to qualified service personnel, who can then determine the exact cause or location of the leak and correct the problem (Wakamiya and Calvano 1977). A bulletin is available from NLPGA dealing specifically with procedures for finding leaks in LP-gas piping systems (NLPGA Bulletin 403-70). For large leaks which obviously pose immediate and serious danger, the procedures described in the NLPGA bulletins 200-73, 201-62 and 204-76 should be applicable (NLPGA 1981). These procedures would generally involve the following precautions:

- If possible, shut off the source of LPG to the leak.
- If there is a fire at the leak, do not attempt to extinguish until the source has been shut off.
- Evacuate people from the general area downwind from the leak where there is risk of flammable mixtures,
- Approach the leak or fire from the upwind side (for purposes of attempting to shut off or to extinguish fire).

These precautions are presented more fully in the referenced bulletins. The NLPGA has also prepared a bulletin for multiple LP-gas customers called "A Guide for Developing Emergency Procedures for Multiple LP-Gas Customers Using a Common Source of Supply" (NLPGA Bulletin 205-78).

9.3.3 Release Prevention and Control R&D in LPG Consumer Storage

Research and development activities directed toward prevention of product releases from consumer storage have been limited mainly to some analysis of accidents to determine causes and to review of code requirements.

Apparently there is no single agency or private organization which has assembled comprehensive statistics over a period of years on accident experience with consumer LPG equipment. The NFPA has published brief stories describing accidents, fires, or explosions involving LPG (NFPA 1952, 1961). Excluding those incidents not consumer-oriented, such as with truck- and rail-transport and also those involving consumer appliances, which are outside the scope of this study, the NFPA presents these statistics:

For the period 1930-1951						
Habitational occupancies	5	incidents	7	killed	17	injured
Mercantile "	12	"	27	"	128	"
Manufacturing "	5	"	15	"	3	"
Storage "	1	"	1	"	3	"
Transportation Equipment	7	"	3	"	50	"
Garages & Service Stations	7	"	12	"	9	"
Miscellaneous	4	"	9	"	164	"
	41	incidents	74	killed	374	injured

For the period 1951-1961						
Residential occupancies	4	incidents	2	killed	11	injured
Public Assembly "	3	"	1	"	16	"
Mercantile "	2	"	0	"	3	"
Automobiles, trucks	5	"	1	"	11	"
Miscellaneous	3	"	2	"	6	"
	17	incidents	6	killed	47	injured

These statistics were extracted from short news stories on the incidents by selecting only those cases where the cause of the incident appeared to involve the portions of systems containing liquid. In other words, numerous incidents involving heating equipment or other appliances, or the vapor pipeline beyond the regulator, have been excluded.

The Consumer Product Safety Commission (CPSC) had on file 105 death certificates related to LPG tanks and fittings as of June, 1976 (Wakamiya and Calvano 1977) (presumably from March, 1967); of these, 64 deaths were caused by explosions, 30 deaths were attributed to inhalation of gas fumes, and 11 deaths occurred in incidents which could not be categorized under specific accident patterns.

The National Bureau of Standards (NBS) investigated incidents involving injuries associated with LP-gas tanks and fittings (Wakamiya and Calvano 1977). Tank sizes were in the range 20 to 100 lb of gas. Information on incidents was obtained through the National Electronic Injury Surveillance System (NEISS) and the Consumer Product Safety Commission (CPSC). The NEISS system reported that 154 injuries related to LPG systems were treated at NEISS hospital emergency rooms from FY 1973 to FY 1976 (4 years). In that period, the distribution of the more frequent injuries was: thermal burns, 38%; lacerations, 12%; contusions and abrasions, 10%. For FY 1976, the mean severity index of LPG-related injuries was 263 as compared to the average mean severity of 101 for the product group "General Household Appliances." Examples of severity values are shown in Table 9.3.

The NBS also has reviewed 31 in-depth investigations done by CPSC and summarized the incidents as in Table 9.4. It is apparent that leakage from tanks or fittings is the primary hazard (20 out of 31 incidents investigated). The NBS also reviewed the standards relevant to the safety of LP-gas containers and concluded from its analysis of the strength of LPG containers that both DOT and ASME containers should withstand the vapor pressure even at high temperatures (up to at least 100°F), if the container is not corroded or otherwise structurally weakened. The NBS also concluded that, generally, code requirements for appurtenances (relief valves, regulators, shut-off valves, plugs, level gages, and pressure gages) appear to be satisfactory. However, they questioned the requirements for the pressure settings of relief valves on ASME tanks, which can result in relief valves opening (for propane) at a temperature as low as 110°F; such temperatures are not uncommon in parts of the U.S. The NBS questioned the standards for requalification of cylinders, particularly by the method of visual inspection, which can be quite subjective.

TABLE 9.3. Mean Severity[a]

Severity Category	Examples of Injuries in Each Category	Severity Value[a]
1	Mild injuries to small areas, for example sprained foot	10
2	Contusion to lower trunk; dislocated arm, hand puncture, non-hospitalized poisoning	12
3	Arm fracture, sprained neck	17
4	Finger crushing, head laceration, punctured eye	31
5	Concussion, fractured neck, ingested foreign object	81
6	Amputation, anoxia, arm crushing, hospitalized poisoning	340
7	All hospitalized category 6's	2,516
8	All deaths	2,516

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[a] Incomplete or otherwise not acceptable data are assigned a severity value of 0 and are not included in calculations of mean severity.

Source: Wakamiya and Calvano 1977.

TABLE 9.4. Summary of LP-Gas Container Incidents

Type of Incident	Reported Cause	Number of Cases
Explosion	Gas leak from tank; ignition by spark or flame	7
Explosion	Gas leak from fuel line; ignition by spark or flame	7
Explosion	Gas leak from unknown source; ignition by spark or flame	6
Explosion	Overpressurization; rusted container; no safety relief valve	1
Explosion	Undetermined	1
Inhalation of gas fumes	Leak; no explosion	1
Other incidents not involving defective containers or fittings		8
	Total	31

The NBS concluded that "there may be a significant risk of serious accidents associated with the use of LP-gas containers," and cited data indicating that explosions were responsible for most of the injuries and deaths. Further, leakage was reported to be the primary cause of explosions. They note as important possible causes of accidents:

- Faulty requalification of cylinders, particularly by visual inspection (as noted above)
- The low setting of start-to-discharge of ASME cylinders
- Poor installation and maintenance.

It was emphasized that detection and immediate reporting of leaks to qualified service personnel would avoid many accidents. The addition of odorant is ineffective in many instances; more positive means of leak detection and education of consumers would be helpful.

10.0 KNOWLEDGE GAPS AND RECOMMENDED R&D

The preceding sections provide an overview of LPG industry operations including production, distribution and storage, and consumer uses. Current knowledge of LPG properties and release phenomena is summarized together with a review of release prevention and control practices. This information supports the perspective that the LPG industry is a mature industry supplying a generally safe, reliable, and, for many consumers, an essential supply of fuel. A vast quantity of experience, technology and operational expertise has been developed in over 50 years of service. The combined influence of industry-generated standards and government regulations promotes safe and environmentally acceptable practices and tends to minimize the incidence and consequences of accidental LPG releases.

Against this favorable background, LPG is involved each year in a few spectacular accidents that receive widespread news media coverage. It is also spilled and sometimes catches fire in more numerous but smaller incidents that receive less publicity. News media coverage has emphasized the casualties, damage and potential hazards of these events generally out of proportion to the true risks involved. Today, as a result of issues raised by the media and a small group of the interested public, there is pressure for government at all levels to review current LPG safety and environmental control practices in preparation for possible regulatory change and other appropriate action. The assessments contained in Sections 8 and 9 indicate that there are some notable gaps in current knowledge of LPG release phenomena and weaknesses in some aspects of LPG release prevention and control. These indications suggest that additional information is needed as a basis for regulatory decision-making.

The DOE Liquefied Gaseous Fuels Safety and Environmental Control Assessment Program has the purpose of expanding the knowledge and data bases that support such decision-making activities. In the last four years, the DOE Program has focused on LNG. This study was commissioned to assist the ESED in planning R&D in the LPG Subprogram. This study scope was broad enough, however, to include recommendations that relate to the R&D prerogatives of other agencies and the LPG industry. The following subsections contain specific project recommendations resulting from this assessment.

10.1 RECOMMENDED R&D ON LPG RELEASE PHENOMENA

The DOE Program has focused on the development of models that predict LNG release phenomena and spill tests to provide data that verify the capabilities of these models. As indicated in Section 8, this knowledge of LNG behavior is generally useful in understanding similar phenomena of LPG releases. As demonstrated by the spill tests at China Lake, California (Koopman et al. 1980), each spill series produces a prodigious volume of data on the material spilled. The potential value of these data when extrapolated to the behavior of other gases, such as LPG, represents an enormous resource that has yet to be fully exploited. A similar conclusion applies to the larve volume of existing data on heavy gas behavior, discussed in Section 8.

Another conclusion to be drawn from Section 8 is that the general scenario of an LPG spill, including liquid loss, pool development, boiling, and general aspects of far-field vapor spread and wind-driven diffusion, is reasonably well understood. Near-field flow, dispersion and flow influenced by local topography and obstacles are areas where more work would be worthwhile. Additional effort should address methods of preventing BLEVE. Knowledge gaps relating to LPG fires, deflagration and detonation are also worthy of further investigation.

Based on the above overall perspective, the following eight projects are recommended.

10.1.1 Project Planning Relevant to Information Needs

A principal purpose of the DOE Program is to provide data and information that assists decision-making by other organizations. Planning R&D that may reasonably be expected to provide results of engineering significance requires a comprehensive understanding of the extent and usefulness of current knowledge and the needs of those who will use the resulting information. With the existing abundance of LGF data and new information anticipated from ongoing efforts, it seems both timely and desirable to consider establishing a formal program-planning methodology to aid in defining the scope of new projects on the basis of their relevance to specific information gaps and needs. The following is an approach for establishing such a methodology.

The major thrust of the recommended effort would create a relevance matrix that associates existing data and knowledge gaps with specific aspects of and needs in LPG release prevention and control. This would help justify new project activities by identifying traceable and practical needs for new data. The matrix would be amenable to control and interrogation by computer which would provide a consistent means of:

(1) correlating related information on different LGF systems,
(2) screening out data and plans that have marginal value, and
(3) providing an up-to-date check list of progress toward goals and issues remaining to be addressed.

Such a planning tool would aid the coordination of industry-supported projects with those of other agencies and could provide information helpful in reducing the controversy that is often associated with safety and environmental control issues. It would also reduce the risk of going forward with projects which may become duplicative or redundant before they are completed.

10.1.2 Near-Field Flow and Dispersion

In accidents involving LPG, significant damage and loss of life can occur when negatively buoyant vapor clouds reach ignition sources. In several instances, LPG fires and explosions have occurred when LPG vapor ignited after following low-lying ground from the spill site. In some cases, LPG vapors flowed over roads, causing vehicles to stall, and attempts by drivers to restart their engines ignited the LPG vapor. At least some of these fires took place at night or during early morning, when the LPG vapor was thought only to be fog. The motion of the vapor cloud at these times is largely due to gravity rather than atmospheric forces; hence, Gaussian dispersion models are not directly applicable for predicting vapor cloud movement. Often, the important dispersion region for LPG vapors is near the vapor source, and most quantitative dispersion models are inadequate in this near-field region.

A model is needed for the gravity- and wind-induced motion of LPG vapor clouds. Existing dense gas dispersion models may provide some background for the development of an LPG dispersion model. The model for LPG should specifically incorporate factors related to the difference in density between the atmosphere and the vapor cloud and should also account for weather factors, topography, obstacles and vapor generation. The validity of the model should be evaluated by comparing its results with experimental data for dense vapor clouds as this data becomes available from ongoing and planned future experiments.

An improved description of vapor cloud drift and flow could help in defining safe distances between possible LPG accident sites (e.g., transfer stations, railroad grade crossings) and ignition sources in the neighborhood.

10.1.3 Knowledge Gaps in Far-Field Dispersion Modeling

Models used for the far-field dispersion of LPG have been derived from LNG and other gases. Assumptions in some of these models, such as neutral buoyancy, do not apply to LPG. Vapor concentrations also vary within real gas clouds. Although vapor concentration variations within clouds are reasonably well accounted for as far as spatial differences are concerned, temporal variations are not handled well by current models. Temporal variations in vapor concentrations are strongly dependent on wind, local topography and local obstructions to air flow, and are thus difficult to model in a generic sense.

LPG dispersion characteristics are important considerations in facility siting. However, studies on LNG currently in progress should provide far-field dispersion models that can be modified for use with LPG to give concentrations as accurately as required for plant siting and safety considerations. There is practical value, however, in performing additional correlations between available test data and model predictions for LPG, LNG and other dense gases. Wind tunnel tests provide useful information on the important elements of dispersion that must be modeled. However, these tests must be very closely correlated with large-scale tests to allow the evaluation of the scalability of data from different gases and spill sizes.

Current dispersion models also use gross, empirically determined correlations to simulate turbulent mixing. More detailed models are available (see Section 8) to address the fundamentals of turbulent dispersion, taking into account effects ignored in simpler models. The use of a complex model for LNG dispersion has shown surprising results, i.e., short dispersion distances. This is due to the gravity spreading of the dense gas. These more detailed models developed for LNG thus may provide some insight to the development of LPG dispersion models. This area deserves preliminary evaluation because the improved models are complex and expensive to use.

10.1.4 Phenomena Relating to BLEVE Initiation

The boiling liquid expanding vapor explosion (BLEVE) is perhaps the most serious manifestation of failure in LPG release prevention systems. When LPG is stored at ambient temperatures under pressure, an accidental release will result in significant vapor generation due to the superheated condition of the released liquid. The rapid vaporization (flashing) of LPG involves two-phase fluid mechanics and is a very complex phenomenon.

Critical two-phase flow of a flashing liquid is still poorly understood. Existing specifications governing safety-relief valve flow requirements for tanks containing LPG are based on gas phase flow capacities. The values thus obtained can be grossly different from the actual relieving capacity of the valve under certain operating conditions (e.g., where a change of state from liquid to vapor occurs in the passage of material through the valve) (Bullerdick et al. 1970).

However, the use of insulation on tanks reduces heat input to the lading, which translates to reduced safety valve flow requirements. Thus a smaller valve can maintain safe tank pressure in an insulated tank compared to that required for an uninsulated tank. The insulation also helps reduce heat transfer to the unwetted portions of the tank, further reducing the potential for rupture. (Adams 1974) Studies done on rail tank car accidents and results of actual tests show that tank ruptures are caused primarily by the overheating and weakening of the tank steel, not by high overpressure (Manda 1978). Therefore, a BLEVE can occur even if the relief valve is operating properly.

It appears that any effort to reduce the occurrence of BLEVEs must consider the mitigating effects of changes in tank construction material, insulation materials and safety relief valve capacity, concentrating on the interdependence of these items in tank behavior. Education and training of emergency response personnel may also help reduce the occurrence of BLEVEs.

10.1.5 Fireball Phenomena

The ignition of an LPG vapor/air mixture followed by rapid propagation of the flame may result in the production of a luminous fireball. Different mechanisms are apparent in fireball formation. One is suddenly generated buoyancy propelling the ignited gases upward as the ambient air rushes in from the sides, choking off the remaining heavier fractions, which may continue to burn as a pool fire. Characteristic mushroom or spherical great balls of fire are formed. Fireballs can also be formed during BLEVEs by the flash-evaporating fuel. A release of 33,000 gallons of LPG from a standard tank car may produce a 300-ft diameter fire hemisphere at the ground level, increasing to a 600-ft diameter fireball when fully developed. While the blast damage is usually mild, thermal radiation damage may be considerable.

Both formation mechanisms are fairly well understood, and predictive calculations can be dependable for practical purposes. Although the mass of pressure-expelled LPG in the second case depends on the nature of the tear and on the quantity of the remaining fuel, worst-case scenarios may be constructed using releases from a broad spectrum of containers ranging from a 200-gallon bottle to barges and ships.

Data are available on moderate and large premixed fuel fireballs, and on small and moderate diffusion flame fireballs. The data on change in size and time to maximum size seem to correlate well, but the theories are not detailed and have some questionable points. Radiation data on premixed flames agree with the theory quite well, but the radiation data on much smaller diffusion flames leave some questions as to the validity of their extrapolation to large spills.

Additional work is needed to determine if available knowledge on the radiation characteristics of test fires is practically scalable to fires that actually occur, e.g., truck, railcar and storage tank BLEVEs. This effort can be aided by reviewing information on past accidents involving fireballs. News media films are possible sources of data that have not yet been fully utilized for this purpose. Data on fireball size, radiation and duration, and other information on the circumstances of accidental releases would provide a basis for comparing theoretical knowledge with the recorded characteristics of actual incidents. This comparison would provide a decision basis for possible experimental verification in later phases of this project.

10.1.6 Radiation From Pool Fires

An extensive literature already exists on the subject of radiation from pool fires. Basically, the problem reduces to predicting radiant heat fluxes from the flame surface generated by the fire. These may be anywhere from 25,000 to 50,000 Btu/hr-ft^2, with the higher values associated with spills on water. The flame surface depends on the flame height, which is found to vary directly as a power of the vaporization rate and inversely as a power of the diameter of the liquid surface. This correlation is of wide applicability and applies to pool fires of gasoline, kerosene, diesel oil, LNG, and alcohol. The burning rate (sometimes referred to as the "level regression rate") depends on the heat influx from the ground or water and the heat feedback from the flame.

While current knowledge appears adequate for establishing isolation distances for structures and materials that may be exposed to LPG fires, existing analytic models do not completely describe the complex phenomena of a free-burning LPG fire. There are several additional areas where new knowledge might lead to improvement in fire, injury and damage control strategies.

The following tasks are recommended.

Task 1. Soot Production and Consumption. Small pool fire tests (say 7-ft diameter) should be performed with appropriate instrumentation to document the mechanics of production, growth, and consumption of soot. Soot can alter the radiation characteristics of a fire, and may affect the potential consequences of the fire. The experiments should investigate a range of LPG composition, and should also cover the possible effects of turbulence, mixing and water in the fuel on soot formation.

Task 2. Radiation. A further series of pool fire tests on both land and water appear worthwhile: 1) to increase confidence in current models and 2) to improve the relatively sparse data base and problems of interpreting data from pool fires on water (see Appendix I). Radiation output would be measured over a wide range of locations, distances and directions including feeback to the pool itself. Other measurements would include: 1) fuel consumption rate, 2) vapor and liquid composition, 3) weather conditions--all as a function of time. The experimental matrix should include experiments to determine the effects of wind (no wind, medium, and strong winds) and lapse rate (neutral and severe inversion conditions). Plans should include allowances for adequate replication to permit improvement of predictive models.

Task 3. Model Development/Verification. The above data and other relevant data available in the literature should be combined to verify and enhance currently developed analytical models of LPG pool fires. Available models seem to adequately predict the effect of an LPG fire on the surroundings. These models account for wind, downwind "trailing" of the flame, burning rate, radiation effects, etc. Information from ongoing future tests should be incorporated into these models as it becomes available.

These models should be used to model a large-scale LPG fire. In this effort, extrapolations from small-scale experimental results should be supplemented by large-scale data available on other fuels.

10.1.7 Deflagration Rates and Pressure Effects for Burning LPG Vapor Clouds

Deflagration is the most common and well-known type of combustion phenomenon. The aerothermochemistry and dynamics of LPG combustion under controlled conditions are well understood and documented. In a laminar deflagration, as the fuel/air mixture is consumed while passing through the flame front and as the gases heat up, the density decreases considerably and the flow volume increases correspondingly. When the combustible mixture is turbulent, the flame front is highly distorted, and, in fact, may become diffuse in nature, in contrast with the sharply defined outlines of laminar flames. Other than qualitatively, the entire phenomenon of turbulent deflagration is still not well understood, and is thus a controversial subject in the literature. Also, since a turbulent deflagration propagation at high velocity can produce pressure waves of significant amplitude, this phenomenon is often mistakenly called a detonation.

Before adequate predictions can be made of flame propagation rates and pressures that might result from the ignition of LPG vapor clouds, further experimental data are needed under conditions simulating possible or likely spill conditions. The recommended project consists of three tasks.

Task 1. Balloon Studies. In this task, hemispherical balloon studies of various LPG/air compositions should be used to extend previous studies in three areas. First, the burning velocity in a quiescent mixture would be determined over a range of mixture ratios. Second, the mixtures would be made turbulent by a set of fans or jets. With proper types and velocities of fans or jets and an ability to allow decay of turbulence before ignition, turbulence conditions matched to various atmospheric conditions should be attainable. The flame propagation rate would then be determined as a function of the turbulence conditions (which should be measured) and, indirectly, related to various wind conditions. Third, if need is indicated by the results of the tests in the turbulent environment, larger balloon sizes should be used. Ground level pressure measurements should be included in all tests.

Task 2. Vapor Cloud Ignition Studies. A series of studies should be undertaken in the open by igniting LPG vapor clouds above a reasonably sized pool (say 7-ft diameter) at various positions under various wind conditions. Heating coils in the pool could be used to control the vaporization rate. For various wind velocities, wind classes, sizes of spill, and two limiting fuel compositions, the clouds would be ignited upwind, downwind, and on the side in different experiments. Because of observed transients in flame propagation on passing from a water surface to a land surface, some tests to investigate this phenomenon should be included in the project. Movies could be used to record the flame movement. Pressure gages at ground level would record the pressure effects. Radiation measurements would be made from the side. Samples should be taken throughout the cloud just before ignition, to determine the composition. Wind measurements would also be required. By monitoring the composition at the ignition source as it is moved into the vapor cloud, the average condition for producing a propagating flame would be determined, as well as the main data on the flame propagation rate under a variety of conditions.

Task 3. Flame Speed and Pressure Studies. Accidental spills of LPG will probably not occur in ideal unconfined geometries, but rather in some trenches, water channels, streets, etc., which may more appropriately be classified as semiconfined geometries. In a cloud resulting from a large spill, roughness of the terrain or natural obstacles such as houses or trees may create preferred semiconfined flame paths. These effects and other parameters, such as buoyancy forces and wind, could cause the flame to accelerate and develop into a disastrous fire with damaging heat and overpressures. To determine the effects of turbulence (caused by a variety of sources) on the flame speed, studies of this type performed by Urtiew (1981) should be continued in semiconfined geometries. These tests should be conducted in a channel of considerable length with an open top, with obstacles placed within the channel. LPG vapor from multiple sources would be used to fill the bottom of the channel and sufficient time allowed to elapse before ignition to give the desired fuel/air composition gradient to the top of the channel. Ignition should occur near one end of the channel. The flame speed and pressure rise would be determined as a function of the fuel type, fuel/air gradient, and external wind direction and velocity relative to the channel. Partial confinement at the ignition end, with a covered top, would permit the determination of the effect of a very high initial propagation velocity on the subsequent flame speed in the open.

In addition to the above tasks, tests should be conducted to simulate the effects of concentration gradient and propagation experience with large-scale fires. Much valuable information may be gained from analyses of such dissimilar fire sources as chemical tank farms, forest fires, and other LGF releases.

10.1.8 Detonation Initiation in LPG Vapor Clouds

It is known that the energy requirement to produce detonation in propane/air mixtures is far less than that required to produce detonation in methane/air mixtures. However, there are few data on the energy (size of explosive charge) required for stoichiometric mixtures of propane or butane in air, and data for oxygen/nitrogen mixtures cannot be reliably extrapolated. Also, no direct data are available on the equivalent sizes of 1) explosive charges and 2) detonating LPG/air mixtures exiting from a duct or opening, relative to starting a detonation in an open system. It is debatable whether a detonation, once started in an unconfined environment, will be self-sustaining. Finally, there is considerable evidence that a disproportionate lowering of the critical limit for a pure fuel occurs when a small amount of more easily detonated fuel is added.

Experimental data and analyses are required to investigate these areas of uncertainty to improve the knowledge for release control strategies. The following tasks are recommended.

Task 1. Review Published Damage Reports. A critical evaluation should be made of any published damage reports involving LPG vapor explosions. The purpose of this evaluation should be to estimate, to the extent possible with available data, the probable blast wave velocities and intensities related to the spill sizes and probable vapor cloud geometries. This task would relate experience with actual events to the theoretical limit of energy release associated with detonation mechanisms. This review would guide the design of experiments serving to interpolate between the experimentally available data and the upper limit of potential destructiveness of an unconfined LPG/air vapor cloud.

On completion of Task 1, gaps in the spectrum of possible detonating configurations should be identified and experiments devised to fill them. Experiments should also be performed to resolve the expected uncertainties and contradictions found in the analyzed damage reports.

Task 2. Bag Studies. Studies of the explosive charge strength required to induce detonation in large bags of combustible mixtures, as is now being done for methane and propane, should be extended to include LPG/air mixtures under both turbulent and quiescent conditions. Gradient mixtures in large balloons should also be investigated.

Task 3. Duct Studies. Stoichiometric mixtures of propane and air in ducts of various sizes should be detonated and the detonation wave be permitted to emerge from the open end of the duct into a vapor cloud in the balloon enclosure; the results are to be compared with data obtained with the balloon-enclosed vapor clouds. The data on flame speed when the tube detonation fails to propagate are also of great value relating to efforts described in 10.1.7 above. In all studies, photographic, pressure and radiation records should be analyzed.

10.2 R&D NEEDS IN RELEASE PREVENTION AND CONTROL

This subsection presents recommendations for R&D activities that address gaps in current LPG release prevention and control (RP&C) practices. These recommendations fall in the following categories:

1) Accident Data Collection and Analysis
2) Assessment and Analysis Supporting Regulatory Decision-Making
3) System and Device Improvements
4) Procedure Development and Human Factors

10.2.1 Accident Data Collection and Analysis

As indicated in Section 9, the collection and analysis of accident data are essential first steps in the development of a sound basis for planning RP&C strategies and R&D activities, developing codes and standards, and making regulatory decisions. Experience in other areas has demonstrated that accident records provide a valuable means of identifying hazardous conditions and accident causes. This has been true in aircraft operations, both military and civilian, shipping, automobiles, fires, industrial safety and many other potentially hazardous activities. Detailed accident records can be valuable when:

1. They require an immediate record of all factors pertinent to the accident while events are still vivid and the circumstances at the time can be identified and reported.
2. They generate a timely analysis of cause and effect and serve to identify the succession of events leading to and contributing to the accident.
3. They provide a basis for identifying unsafe practices and for establishing improved procedures in noncontroversial terms.
4. The statistical accumulation of accident records provides a perspective of hazards involved and provides a basis for

estimating the economic value of more complex and expensive accident prevention methods and safety equipment.

5. Such records also reveal problem areas where innovation may be required to provide new types of equipment to solve recurring problems not amenable to existing procedures or devices.

6. Having this information in a public record makes it more accessible to those who may be in a position to offer improvements either from other related fields or through invention or innovation.

There is a general lack of LPG accident information available as a coherent and consistent body of data. With some notable exceptions, much of the information that does exist on LPG accidents is not readily accessible and fails to meet all of the above criteria. This lack precludes the objective assessment of safety-related R&D needs and even the determination of whether such R&D is necessary. This situation arises principally because no single agency has responsibilities for safety and environmental control in all aspects of LPG operations. The need for additional accident data and analysis is described below as it applies to each area of LPG operations.

Production Facilities

There is no national-level reporting requirement for accidents and incidents involving LPG releases at petroleum refineries and natural gas extraction facilities. Because of this lack, information on accidents and releases must be obtained either by direct contact or by reference to public literature. Direct contact with the owners or operators of the facilities is, of course, possible. However, the detail in which records are kept and the time for which records are available vary widely between facilities. Some owners and operators are reluctant to release such information.

Other sources of information are facility insurers and the court records of legal cases concerning LPG accidents. These are subject to the limitations stated above, in addition to their having knowledge only of those accidents or releases which lead to insurance or other legal claims. These sources cannot provide information on such minor releases as leaks at valve stems, flanges, vents, etc., that were corrected by plant personnel without significant loss of product.

Reliance on public literature sources for information on accidents and incidents involving releases has a built-in bias in that only the major accidents or incidents tend to be reported. The major accidents reported may provide information of more significance in determining the need for safety-related research than information on minor releases. However, both major and minor accident and incident data are required to produce a meaningful data base.

Examination of the details that are reported on each incident indicates a variety of causes. The basic problem remains that the data base is too small for meaningful analysis. For example, it would be helpful to know what percentage of the accidents were initiated by human error or the breakdown of specific types of equipment and procedures. Also, it is apparent that a failure in one piece of equipment can start a chain reaction which results in far greater damage.

It is recommended that a study be initiated to determine if it is possible to assemble a sufficiently large data base from available sources so that meaningful analyses of incidents and accidents occurring during LPG production can be made.

Improved DOT/OPSR Report Form for Pipeline Incidents and Data Analysis Project

The DOT Office of Pipeline Safety Regulations (OPSR) has been receiving pipeline leak reports since 1968. The report form does not request information on the size of an operator's LPG system, so that incident data cannot be normalized on a mileage basis. The OPSR has reported, over the years, annual totals of incidents and consequences. These data show that corrosion and third party damage are the two major categories of accidents on all liquid pipelines. Detailed analysis of the accident data has not been performed to evaluate the need for regulatory changes that might help reduce the frequency or consequences of the incidents.

To modify the OPSR form and the subsequent analysis of data would be a rather minimal effort with major potential benefits of obtaining a better understanding of release causes. For example, the effectiveness of "one-call" systems could be evaluated if more data were available.

The following project is recommended as an effort to be conducted by the OPSR because it is directly responsible for issuing the form and has the balance of the data needed in subsequent analysis.

Task 1. Review of Need to Change OPSR Reporting Requirements. The OPSR incident reporting form for liquid pipelines should be reviewed by an expert panel which would recommend whether additional information should be required in future incident reports. Changes in the form would also be considered with the objective of obtaining maximum information for assessing potential accident causes, particularly the seam of ERW pipe, which has been suggested as a potential problem area. Any major change in the form should be accompanied by some information for operators to help them in correctly filling out the new forms.

Task 2. Data Computerization. All of the accident data should be computerized so that statistical summaries and various cross reference options can be studied. If there is a change in the reporting form or format, the new and old forms should be compatible in the computerized system.

Task 3. Analysis of Accident Causes. A detailed study of major accident causes should be undertaken to define any trends in accident data that would suggest regulatory amendments or changes in recommended operating practices.

Truck Accidents

The need exists for the collection of more definitive and comprehensive LPG truck accident and damage data to assist the analysis of accident causes.

Several agencies are currently involved in collecting and analyzing accident data for trucks involved in interstate commerce. The Bureau of Motor Carrier Safety (BMCS) regulations require a report to be filed when an accident involves a motor vehicle engaged in interstate commerce. These accident reports must be filed if any of the following three conditions occurred: a fatality, an injury requiring medical attention, or property damage of $2,000 or more. The Materials Transportation Bureau (MTB) in the DOT collects data on releases of hazardous materials during transportation. Under these regulations (49 CFR 171.16), transportation is defined to include loading, unloading and temporary storage. An incident report is required from interstate hazardous material carriers wherever a release of material occurs from the package or the transport vehicle. The reported release need not be the result of a transportation accident.

A fairly complete data base on LPG accidents and releases from transport vehicles or transported packages thus exists for carriers engaged in interstate commerce. In addition, some states require detailed reporting of cargo tank accidents, and the NTSB reports in detail on major accidents.

However, there are currently no reporting requirements for LPG carriers engaged in intrastate commerce. All carriers should be required to report accident/loss of lading data to the appropriate agency to provide better perspective on LPG truck accident and damage rates.

Import/Export Terminals

There is a lack of detailed information on incidents in LPG import/export terminals and on whether the design of these facilities offers adequate protection to the public in case of an accident. While only one accident report involving a U.S. terminal is on record, the NATALIE WARREN accident at Newark, New Jersey, on July 7, 1951 (Appendix H), illustrates the potential severity of the problem. This accident was a classic example of a BLEVE. There are no data available to determine if conditions exist at import/export terminals that might allow a similar disaster in the case of other accidental releases. A survey should be performed to assess the design features of LPG import/export terminals and to collect information on incidents which have not been reported in the open literature. In most cases, such facilities are under the jurisdiction of port authorities who could be expected to cooperate with Federal authorities seeking this information.

Peakshaving Plants

There is little information available on details of incidents at peakshaving facilities. It is not clear whether accidents or incidents involving LPG/air peakshaving facilities must be reported under the Natural Gas Pipeline Safety Act of 1968. Title 49 of the Code of Federal Regulations covering "Regulations for the Transportation of Natural and Other Gas By Pipeline" under Part 192.11.--Petroleum Gas Systems, states: "(c) For the purpose of this section, petroleum gas means propane, butane, or mixtures of these gases, other than a gas/air mixture that is used to supplement supplies in a natural gas distribution system."

While Paragraph "c" excludes propane/air systems from regulation under Paragraph 192.12, it may not exclude these facilities from the reporting obligations of Part 191, which requires a telephone report of all serious accidents (death, personal injury requiring hospitalization, gas ignition and property damage in excess of $5,000). In addition to incidents reported by telephone, certain other classes of leaks must be reported on a special form within 20 days. Annual summary reports of leak incidents are also required.

The information gathered by the MTB to date does not distinguish between LPG/air and natural gas systems. Thus, regardless of current reporting requirements, accident information for LPG peakshaving operations cannot be extracted from the DOT's leak reports. The only other reports made are usually those required to settle claims due to property damage or personal injury. These are rarely made public. The records of insurance investigators tend to be closely held because of the risk of opportunistic legal actions by parties not otherwise involved. Thus, there is no other readily available source of information on accident experience in this field.

It is recommended that an investigation be made of peakshaving plant activities to determine if the number and frequency of accidents are sufficient to generate a useful body of data. This will require a survey of peakshaving plant operators to obtain estimates of the number of reportable incidents. Fire records, insurance company data and other sources of information, such as court records of legal cases concerning LPG accidents, should also be examined. Many of these sources may be reluctant to provide details of specific incidents; however, in this survey, only data revealing the number, frequency, and general nature of the incidents need be gathered.

The results of this survey should indicate whether or not benefits would result from a national reporting requirement for this information.

Consumer Storage and Appliances

There is no one organization collecting failure or accident data on LPG systems at the consumer level. The Consumer Product Safety Commission and the National Fire Protection Association have some accident data, although these do not describe the broad range of incidents and variables involved in consumer accidents.

More complete data are necessary for analysis to determine the direction of future R&D, code development and possible regulatory actions. A single agency should be assigned the responsibility of operating a collection system for accident data similar to that operated by the MTB for gas distribution and transmission line failures.

Temporary Installations

Temporary propane installations by consumers and industrial users and contractors, such as those in the construction and equipment repair industries, often create potentially dangerous situations. There are no restraints other than the judgment and experience of the individuals involved to control the safe application of LPG in a variety of purposes ranging from warming building enclosures to fueling engines. These practices occur daily, not only in the construction industry but on farms, in industrial shops, garages and private homes.

There is some measure of control exercised over propane installations in mobile homes, recreational vehicles and similar applications that are recognized and categorized for control purposes. However, it is virtually impossible to anticipate or prescribe safety measures for the endless variety of temporary arrangements that are contrived to solve problems occurring in the broad spectrum of industrial activities. Individually these situations are unique; collectively they represent a significant number of potentially hazardous situations.

Clearly, the problem of improving LPG safety in this area is a difficult one. It may only be possible to promote safe practices by wide publication and dissemination of safe operating practices and procedures.

To assess the magnitude of this need, it is recommended that information be gathered from available sources on accident experiences in temporary or emergency situations. These may be obtained from sources such as fire records, insurance claims, court cases, equipment manufacturers, propane suppliers, industrial safety specialists, OSHA and other agencies.

The development of this information may reveal the areas where accidents have been more prone to occur. These are expected to fall in either of two categories: (1) those caused by individuals working on their own behalf and (2) those caused by individuals working as employees of someone else. In the former case, accident mitigation is difficult. Recommendations for protecting such individuals from unsafe practices must rely on warnings and cautionary messages and instructions provided by propane suppliers and equipment sales people. The second case is somewhat easier to control in that employers can be expected to demonstrate some responsibility in selecting personnel for such work and demanding attention to safety.

The data provided by this assessment would be the basis of possible recommendations for additional effort. In formulating such recommendations, the following criteria should be observed.

- The hazard must be real and sufficiently common that safety recommendations can produce a significant benefit.
- The cost of implementing safety recommendations must not be excessive for the value obtained.

10.2.2 Assessment and Analysis Supporting Regulatory Decision-Making

The LPG industry is highly regulated, indeed, to an extent that some experts suggest it is over-regulated. The purpose of existing and new regulations should be to minimize, at reasonable cost, the frequency and consequences of accidental LPG releases. As technology improves and situations develop and change, it would appear prudent for regulatory agencies to review the efficacy of existing regulations before considering additional rules. The following areas of research are recommended to provide information that may assist future decisions to introduce new regulations or to amend those that are no longer relevant.

General Safety Design and Practice at LPG Facilities

Present safety standards and regulations pertaining to LNG are far more restrictive than those pertaining to LPG, and yet there are indications that handling LPG involves greater risks. A large amount of work has been done in recent years in establishing the need for and implementing 49CFR Part 193-Liquefied Natural Gas Facilities: New Federal Safety Standards. No comparable standards have been prepared for LPG.

It is recommended that Part 193, and the literature reporting supportive R&D and discussions leading to its preparation, be reviewed to determine areas of relevance to LPG safety. Not all sections of Part 193 need to be examined. Those pertaining to cryogenic temperature considerations are not pertinent. Those pertaining to impoundment design and capacity relate to LPG liquid storage only. Nevertheless, much of what has been done in examining LNG safety can yield a useful basis for planning the improvement of LPG safety standards as required.

Bulk Storage of LPG at Refineries, Gas Extraction Plants and Terminals

LPG tanks located in large tank farms such as refineries, gas extraction plants and terminals follow oil tank safety standards more closely than LNG tank requirements. These standards do not adequately reflect the nature of the hazards inherent in LPG. In LPG, accidents, and/or explosion in one tank may readily spread to adjacent tanks. While fuel tank installations are usually isolated from public property, large conflagrations may cause damage or expose third parties to unnecessary hazards.

Considerable research and analysis has been done in recent years on safety aspects of LNG storage tanks that may have application to LPG storage tanks. The work done on LNG tank design, diking requirements, spacing and degree of isolation, as well as an assessment of the potential risks with current practice, should be reviewed to assess the need and desirability of upgrading LPG bulk storage practices and requirements.

Peakshaving Plant Storage Facilities

Considerations similar to those above apply to storage at peakshaving plants. Because of their smaller size, peakshaving storage tanks usually are not required to be spaced and diked for damage control to the same degree as larger storage tanks. However, the popular 8-ft diameter pressure vessels used are often nested in groups of a dozen or more, presenting significant hazard in the event of rupture and fire in any one tank. The availability of acceptable peakshaving plant sites is limited, and there is a natural tendency to expand capacity on existing sites as load demand grows. This leads to the crowding of storage tanks to the extent of legal limits.

A technical review of hazards, accident experiences and existing practices is necessary to assess the need for more stringent requirements (number of tanks, spacing, diking, drainage, etc.). Many peakshaving plants are located on flood plains to minimize impacts on adjacent real estate development. Therefore, the use of dikes for flood protection as well as fuel containment should be considered. Sources of information for this review may include equipment suppliers, utilities, local public administrators and insurance underwriters.

Rail Transportation

Until the last few years, the derailment of railroad tanks cars resulted in frequent and sometimes catastrophic spills of hazardous liquids. In derailments where high-energy impacts were involved, the principal reason for failure of tank car bodies has been puncture by the coupler of adjacent cars in the train. Hazardous liquid spills of materials such as liquefied petroleum gas, ammonia and chlorine have caused considerable damage in the vicinity of the derailment (see Appendix H). The increasing frequency and hazardous effects of these accidents has been altered by the development of new safety appliances for tank cars carrying LPG and other hazardous liquids. After extensive research and development sponsored by the Department of Transportation, the Association of American Railroads and the Railroad Progress Institute, three basic safety appliances were mandated as standard equipment for LPG tank cars. These three basic appliances are:

1. "Shelf" couplers, designed to resist vertical disengagement;
2. The tank car head shield, designed to prevent head punctures; and
3. Thermal insulation jacketing or coatings, designed to mitigate the potential for catastrophic rupture as a result of tank overheating.

The application of these appliances has been required by law for the last three years. The entire national fleet of Types 105, 112 and 114 tank cars are fitted with these appliances.

However, there is continued public pressure to improve the safety of all transportation systems. As a result, there is a call for consideration of further improvements in tank car design that may reduce the frequency and severity of accidental releases during derailments. In the environment of increasing safety consciousness, there is a risk that new safety appliances will be developed without the benefit of risk assessment to evaluate their incremental worth. It is thus suggested that risk assessment techniques be applied to evaluate the current worth of safety appliances and potentially new safety device developments that are contemplated. Cost estimates could also be made for implementing new safety requirements. These cost estimates, coupled with the risk information, could be used to help comply with the administration's mandate for cost-benefit analysis of new regulatory requirements.

This study will provide up-to-date risk assessment-based insight on the incremental value of safety appliances on railroad tank cars and the cost of implementing these safety-related changes. This information would provide guidance for establishing the nature and priority of future safety appliance development.

Ship Transportation

Considering the increases in LPG traffic that are projected and the risks to the public and the environment, higher standards should be required for certain conventional ship systems to minimize the possibility of ship collisions and groundings.

The rapid increase in sea transport of liquefied gases in bulk in the late 1960's created a need for international standards to insure their safe carriage with a view to avoiding or minimizing the risk to the ship, its crew, personnel at shore installations and the environment. Recognizing this need, the Inter-Governmental Maritime Consultative Organization (IMCO) adopted on November 12, 1975, the Code for the Construction and Equipment of Ships Carrying Liquefied Gases in Bulk.

The Code deals primarily with ship design and equipment. Other important facets of the safe transport of the cargoes, such as crew training, operations, traffic control and handling in port, remain primarily the responsibility of the individual governments of the countries where the ships trade. The IMCO is currently examining the possibility of establishing uniform minimum-crew and crew-training regulations.

The integrated LPG system is capital-intensive in all its parts (production, shipping, and terminaling), and to be competitive and profitable, it must be safe and reliable. Prudent shipowners and responsible designers recognize this and generally include features and equipment in excess of regulations to enhance reliability as well as safety.

The operating experience of gas ships has led to an apparent consensus that a release of cargo large enough to be hazardous to the environment and the public is significantly more likely to result from external factors, such as collision, stranding (grounding), natural disasters and sabotage, than from failure within the ship. From the beginning of the development of the design standards, it was recognized that a severe collision or stranding could lead to cargo tank failure. Design and construction alone cannot eliminate the adverse results of a severe collision of stranding, although recent groundings of two LNG ships demonstrated that even a fairly severe grounding does not necessarily lead to cargo tank failure. However, some improvements in design, together with traffic control and adequately trained crews and responsible officers, can reduce the possibility in coastal or harbor areas of a collision or stranding of the magnitude that would be required to result in a large release of LPG cargo.

Prevention of serious collisions and strandings may be enhanced by stronger regulations on navigation equipment, collision-avoidance systems, steering apparatus, continuity of electric and propulsive power, and ship control. Many improvements in such gear are incorporated in most recently built LNG and refrigerated LPG ships by action of prudent shipowners and responsible designers, although not required by current regulations.

The following specific improvements to steering gear and control systems should be evaluated on a cost/benefit basis for their potential effect on reducing the risk of collisions and groundings of LPG ships:

1. Position Locating Systems of the Loran C or equivalent type for accurate continuous determination of position in coastal waters, to assure that LPG ships are in the sea lanes designated for approaching and departing U.S. harbors.
2. Long-range (4-in.) and navigational (1.2-in.) radars of demonstrated reliability.

3. Collision Avoidance Systems (CAS) with manual and automatic monitoring of targets, which interface with both the 4-in. and 1.2-in. radars.

4. High-seas position-determining systems of a modern electronic type, such as Omega or a Satellite Navigation System (SNS).

5. Doppler Log Systems for indicating ship speed for use with the CAS and SNS systems.

6. New requirements for the ship's steering apparatus extending the requirements for dual controls and power units to the hydraulic and mechanical equipment, so that dual systems extend from the steering wheel to the rudder stock.

7. A requirement that operation of the steering apparatus not be interrupted during the period between loss of main electric power and its replacement by emergency power. This could be done by requiring a source of temporary power, or by requiring that the emergency generator be operating when in maneuvering waters so it can provide power essentially instantaneously to the steering apparatus on loss of main electric power.

8. The main switchboard should be comprised of three sections interconnected by automatic circuit breakers, so that the ship can be operated and maneuvered at reasonable speed with any one section out of service.

9. Regulations requiring that electric generating sets have sufficient capacity to carry the necessary sea load under normal operation with any one generating set in reserve. This is also a requirement for seaworthiness, so that a ship is not permitted to leave a port without a set in reserve. This requirement should be extended to prohibit a ship from entering port under her own power unless she has in reserve one generating set capable of supplying the necessary electric load for maneuvering the ship at reasonable speed.

10. A bow thruster unit of a specified size to improve control and steering at low speeds, as in channels, where the large sail area of LPG ships makes control difficult.

11. A centralized engine-control and monitoring system with bridge control of speed and direction to minimize response time when maneuvering.

A parallel evaluation should be made to determine what steps would be necessary to implement these improvements should the technical evaluation indicate a positive cost-benefit relationship.

To complete the evaluation, it is necessary to consider that a vessel of more than 5,000 and 10,000 tons displacement must strike a large capacity refrigerated LPG or LNG vessel at an angle of 90° at its most vulnerable point and a speed of 8 to 10 knots in order to penetrate a cargo hold. If the gas carrier is the impacting vessel, a minimum of 12 to 15 knots is required to penetrate the forward cargo hold. From this it would appear that the evaluation should extend to all vessels over 5,000 to 10,000 tons and not be limited to gas carriers.

Risk Assessment of Marine Transportation

Based upon projected growth in the LPG industry, significant increases in maritime traffic can be anticipated. The increased numbers and movements of LPG ships and barges combined with construction of new terminals and expansion of existing facilities suggest that greater emphasis needs to be placed on assessing the associated risk.

The majority of risk analyses have been made in connection with LNG rather than LPG facilities and operations. Many of these studies have been site-specific risk assessments required to obtain approval for the site. In addition, non-site-specific risk studies have been supported by agencies that include the USCG, the U.S. Maritime Administration and the DOE. Many of the LNG safety and risk studies funded by these agencies have been performed to assist them in their decision-making and regulatory roles in general, rather than in the site-approval process.

Previous studies have been quite inconsistent in their methodology. Some considered data for only a specific port and its approaches. Others considered data for "similar" ports or all ports in the United States. Still others were based upon data for ports worldwide. In several of the studies, the sizes of the ships involved in past accidents were taken into account in determining ship-to-ship collision probabilities. Impacts by ships too small to cause damage to gas carriers were not counted. However, if some of these smaller ships carry hazardous cargos which could burn or explode as a result of the collision, this approach underestimates the possible hazards.

The range of times for which accident data were acquired varies widely from report to report. Some investigators considered only the two years preceding the study, others gathered accident data for about 10 years.

Previous risk studies seem to focus on the low-probability, high-consequence events that could affect nearby population centers. High-probability, lower-consequence events are not considered. The assessment of risk should include as complete a spectrum of undesired events as possible, since total risk is the sum of individual risk from each possible undesired event.

Mitigating or risk-reducing factors, such as improved ship traffic control, crew training, cargo handling equipment, etc., can contribute to reducing the probability of an accident in which a cargo release occurs. In some instances credits for such factors were assigned in an arbitrary manner.

Human error is not usually accounted for in previous analyses. When it is accounted for, event sequences initiated by human error were usually treated as independent. Experience shows that human errors are not independent of one another. Further, human errors are often assumed not to occur because they are presumed to be eliminated by safety devices. devices.

It is recommended that a non-site-specific risk assessment for LPG maritime transportation (both barge and ship) based on a critical assessment of previous efforts be conducted. The methodology and input data for site-specific risk assessments should also be developed. Once completed, provisions should be made to periodically update these analyses to accommodate new knowledge or changing conditions. A non-site-specific risk assessment for LPG maritime transport will be a valuable tool for regulatory agencies to assist decision- and rule-making. Site-specific risk analysis would be of assistance both to regulatory agencies and the industry in evaluating sites for LPG terminals. A risk assessment for cargo transfer operations for LPG ships and barges has been performed (Martinsen, Johnson and Welker 1980). This document should be reviewed for applicability to and as background material for this task.

The recommended risk assessment should include the following:

1. Additional accident scenarios for the high-consequence, low-probability events should be evaluated.
2. Risk should be assessed also for the low-consequence, high-probability events.
3. The risks associated with water transportation of LPG and with other hazardous materials should be compared. Comparative estimates are generally more accurate, and more readily understandable, particularly if the basis for comparison is common practice.
4. Better input data should be developed to increase the reliability of risk analyses. A worldwide incident-reporting system, including coverage of minor incidents and near misses, would help to provide relevant data.
5. Data from "man-in-the-loop" trials at a ship-simulator facility should be collected. Such data will increase the reliability of synthesized probabilities for ship collisions at specific sites.
6. Confidence levels and discussions of uncertainties should be provided for the factors considered in the assessment.

BLEVE Potential in Consumer Storage Tanks

The BLEVE phenomenon of LPG pressure vessels exposed to fire has caused casualties comparable to those of vapor cloud explosions. The four principal ways in which LPG vessels can be protected to reduce the probability of a BLEVE are by 1) sloping the ground under the tank, 2) adding insulation, 3) providing water drenching facilities and (4) depressuring the tank in an emergency.

These principles are well understood by refinery, LPG and peakshaving plant operators. However, it is possible that numerous consumer storage tanks existing without these measures could be subject to BLEVEs in the event of a fire. Tanks may typically be installed on level ground or even in a depression so that any released liquid cannot drain away. The installation of water drenching lines is not common practice. The use of fire-resistant insulation and/or specialized depressuring valves may not have been applied on the basis of the costs involved in the absence of specific requirements.

There is little information available on how severe this problem is with consumer-type installations. A cursory review, particularly in rural areas, indicates many installations where a BLEVE could occur.

A project is recommended to determine the possible extent of this problem. The frequency of BLEVE occurrence in consumer storage should be established. It should also be determined if local installation codes and practices fully account for this possibility and if appropriate installation procedures are being followed.

The results of this study should indicate whether changes are appropriate in related codes and regulations or inspection/enforcement procedures.

10.2.3 System and Device Improvements

This subsection contains recommendations for RP&C system and device improvements that are predominantly the domain of industry and supplier R&D initiatives.

Methodology for Assessing Pipeline Integrity

At the present time, the structural integrity of a pipeline cannot be determined with high confidence either for a liquid or a gas pipeline. It is known from the operating history of pipelines that sometimes those pipelines which have been subject to major pressure changes, for example, due to loss of pumping stations, have experienced failures. When a pumping station fails, the pressure gradient in the pipeline to the next pumping station changes, subjecting certain parts of the pipeline to higher pressures than previously encountered; failures have occurred in these areas. Line pipe steel does not experience any general loss in strength with time. Rather, what happens is that defects present in the material are slowly growing under standard operating conditions. These flaws are not critical in normal operations. On occasion, however, when the line is subjected to a higher pressure, even though it is not above the maximum allowable operating pressure (MAOP), the flaws are of a critical size and a failure results.

Currently, there are several methods in use to assess the condition of an operating pipeline. One method is to survey the line using pipe-to-soil potential as a measure of the effectiveness of the cathodic corrosion protection system that is required for LPG pipelines. Highly sensitive instrumental "pigs" can be run through the pipe to check on a number of potential degradation mechanisms that may develop.

Another method used on an infrequent basis is a hydrostatic retesting. It has been suggested that pipelines should be periodically retested hydrostatically to validate them for service. While this concept sounds like a desirable one, it has the undesirable feature of applying large pressure cycles that may cause flaws to enlarge.

It is recommended that a state-of-the-art study be conducted to determine an optimal integrity assessment procedure for LPG pipelines. Appropriate laboratory and field experimentation should be conducted to confirm the results of this assessment.

Storage Tank Foundation Design

Tank and equipment foundations are not always designed with adequate consideration for soil movement. While tanks and connected components may be built on independent but adequate foundations, soil movement between the foundations may strain interconnecting piping and devices.

Industry experience in various climate and soil conditions and with various industrial piping systems should be examined to determine the extent and nature of movement over time that has occurred in existing refineries, gas plants and other similar facilities. Emphasis should be on small groups of tanks and other interconnected units. The nature and extent of movements should be examined, and techniques for preventing strain on piping and components, if necessary, should be reported. This information would be a basis for recommendations to be incorporated in facility design standards and practices as required.

Tank Truck Appurtenance Design

The available statistics for LPG tank accidents indicate a large incidence of tank leakage after overturns. As more LPG is transported by truck, the potential increases for accidents resulting in leakage and more serious consequences.

To reduce the incidence of tank leakage in overturn accidents, a research project should be initiated to improve the design of tank appurtenances. The objectives of this effort should be:

1. Reduce leakage from fittings, vents, and valves, by studying adequacy of location, performance and protection.
2. Develop standard overturn test requirements that would prove out the total system, including vapor recovery components. Methods for supporting and anchoring the shell should be reviewed and improved if necessary.

Fire-Resistant Fuel System in Truck Transportation

There is the potential of fire resulting from failure of the truck fuel system in an accident, even though no damage to the LPG tank or trailer results. Several accident reports by NTSB in 1972 and 1973 indicated that fuel lines ruptured and/or caught fire in relatively minor accidents. The fuel crossover line seemed to be particularly vulnerable. However, valves to eliminate diesel crossover lines have been developed recently. Although not a major cause of fires in the past, this potential is present in any accident and could turn minor incidents into major events.

Fuel system requirements for LPG trucks should be reviewed with the objective of developing improved systems, particularly in more vulnerable areas such as the crossover lines.

Prevention of Sloshing in Truck Trailers

Several accidents have occurred in which a contributing factor was "sloshing" of LPG in partially-filled truck trailers. Sloshing can interfere with driver control. Two NTSB reports of highway accidents involving cargo tanks indicate that liquid cargo sloshing was a critical contributor to the overturn. A study to determine the severity of this problem and mitigation measures should be conducted.

Automatic Leak Detection and Alarm Devices

A number of explosions of LPG resulting in deaths or injuries have been caused by the ignition of LPG leaking from tanks and fuel lines. A study of safety problems associated with LPG tanks carried out by NBS for the Consumer Product Safety Commission indicates this is the primary hazard associated with LPG containers used for consumer storage. Currently an odorant in the gas is used for leak detection. This is ineffective in many cases. A project should be initiated to develop an economical leak detector/alarm for customer use. The detector/alarm should, at a minimum, provide an audible alarm and probably automatic shutdown of lines leading from a tank to any enclosed spaces. Some types of natural gas alarms are available and should be investigated to determine their adaptability for this purpose.

Temperature Settings for Pressure-Relief Valves

The requirements for design, fabrication and operation of LPG tank safety-relief valves appear to be adequate; however, these valves can be set to vent at pressures corresponding to common summer ambient temperatures in many parts of the country. Relief valves on ASME containers may be set to open at the lower limit of 88% of the design pressure of the container, thus allowing the valve to open when tank pressure (which is equal to the vapor pressure of the stored gas) reaches 220 psig. As shown in NFPA 58, the vapor pressure of propane is 220 psig at temperatures between 105°F and 110°F. One of the main causes of incidents at consumer storage tanks is explosions caused by leaks. Such venting, when the tank itself has a safety factor of 4, appears to present unnecessary risk. Uncontrolled and frequent venting of gas from safety-relief valves caused solely by ambient temperature excursions should therefore be minimized. A study is needed to review and recommend possible changes in requirements for relief valve pressure settings on consumer storage tanks. Studies performed in this area for rail tank cars should be reviewed for applicability.

10.2.4 Procedure Development and Human Factors

Recommendations in this subsection address apparent gaps in RP&C procedures in LPG peakshaving, transportation and accident response. In all of these areas, human factors have not yet received sufficient attention. The recommended action includes education, training and communications to address human factor contributions in these problem areas.

Operating Practices in Peakshaving Plants

Peakshaving plants may be left unattended during most of the year except during delivery of LPG. Many are operated only 10-20 days per year during extremely cold weather to provide additional gas to meet peak space heating loads. Deficiencies discovered during plant operation are apt to be tolerated for the short operating period rather than being serviced immediately under possibly severe weather conditions.

Service personnel may be required on short notice to leave other work and assist with repairs. This can result in the use of less experienced personnel perhaps working unsupervised on potentially hazardous assignments. Under these circumstances, simple, well-defined operating procedures and thorough crew training are imperative.

A review of the frequency of these practices may identify the need for improved control and safety equipment, operating procedures and training of personnel. Utilities should be contacted to review existing practices, the application of available technology and procedure enforcement practices. This review should provide a basis for encouraging the whole industry to adopt state-of-the-art practices.

Education and Training of LPG Truck Drivers and Emergency Response Personnel

Driver error, either because of poor judgment or poor operation, is a frequent cause of accidents with the potential of releasing LPG into the public environment. Past studies indicate that human error is the primary cause of truck accidents.

Some educational efforts are underway. The American Trucking Association, Inc., prepared material for safety programs to its members, and the National LP-Gas Association has prepared and issued a number of bulletins setting forth general safety practices for truck transport as well as other LPG activities. These activities do not appear adequate as an organized approach for reaching all persons responsible for RP&C in LPG truck transportation.

A project should be initiated to develop a broad-based safety education and training program for LPG truck drivers, management and emergency response personnel. This should include not only truck operations, but also loading, unloading and maintenance. The program should also provide education for fire and police departments and other emergency personnel in release and damage control procedures.

Delayed Failure of Railroad Tank Cars

A delayed failure of an LPG tank car after mechanical damage creates extremely hazardous situations because such failures cannot be reliably predicted. A delayed failure of a tank car refers to a tank car that has sustained damage in a derailment, did not fail at that time, but subsequently ruptures. A large fraction (some 75%) of tank car ruptures have been shown to be due to mechanical damage, and several known cases of delayed rupture have been reported and studied. In the documented cases, delayed ruptures have occurred many hours (from 40 hours to 8 days) after the derailment. The ruptures have in each case occurred unexpectedly and have caused serious injury and death to personnel in the vicinity.

Based on the information available at the present time, it is believed that it is impossible to accurately estimate the failure pressure of a pressurized tank car or cylinder that has mechanical damage.

Preventing the failure of damaged tank cars may be possible by various measures. In the situation where a tank car has been damaged in a derailment and has been put back into service, the obvious precaution is to inspect the car for remaining mechanical damage. Any dents with gouges, grooves, or scrapes should be cause for rejection. From the few accidents which have been examined, it appears that the circumferential welds are particularly susceptible to damage, because the weld reinforcement protrudes above the surface of the tank, which makes it vulnerable to damage. The new requirement for thermal protective coatings on rail tank cars, however, makes visual inspection very difficult and expensive. An improved means for inspecting these cars should be developed.

Another case is that of a mechanically damaged loaded car. The best solution in this case appears to be to reduce the pressure in the car as rapidly as possible, consistent with other safety considerations and if possible before the tank car is moved. Since the car is filled with a compressed liquefied gas, the pressure in the tank is a function of the temperature of the liquid. To drop the pressure, the temperature of the complete tank and contents must be changed.

In view of the fact that the failure pressures of tanks with mechanical damage defects cannot be accurately estimated and that the measurement of controlling parameters is difficult, if not impossible, it is recommended that research be conducted on methods for handling derailed LPG tank cars and training emergency response personnel. Consideration should be given to developing inspection methods for insulated cars and requiring regular inspections of tanks before loading to discover possible mechanical damage.

Finally, consideration should be given to the possibility that tank cars can be modified to make them more resistant to mechanical damage. This has already been attempted in the area of the heads where head shields are required. From the data available on tank car shell properties, it appears that normalizing the shell steel and improving the fracture properties of the circumferential welds may provide additional protection. Normalizing the tank shell will increase the fracture toughness, and the steel will exhibit more ductility in the presence of a notch.

In an accident at Crete, Nebraska, February 18, 1969, a stationary tank car was struck by part of a derailing train. The tank, which was built to the same specifications as LPG tanks, and contained anhydrous ammonia, fractured and released 30,000 gal of NH_3. Tests showed that brittle fracture had occurred due to the +4°F ambient temperature and a +75 to +85°F nil-ductility transition temperature (NDTT) of the steel tank. It might be useful to perform some tests to determine whether rail tank cars should be required to use the same stronger-type steel as trucks. It appears that improved fracture propagation resistance is not necessary, but rather that fraction initiation resistance should be increased to assure that all reasonable precautions are taken to prevent fracture from initiating.

Handling of Spilled LPG

Removal of spilled LPG is often done by flushing with water as in the case of gasoline spills. While this method is effective, a potential hazard remains until all liquids and vapor have been dissipated, not only from the spill area, but also from the conduit into which the fuel has been flushed. Explosions have occurred in sewer lines filled with combustible vapors from gasoline and LPG spills. These vapors are heavier than air, tend to collect in depressions and may remain in sewer lines from some time following a flushing operation. The potential hazard of ignition of spilled LPG is a serious problem, particularly since it may expose third parties not involved in or even aware of the spill.

Available accident reports should be reviewed to determine the frequency of spills, the frequency of ignitions and the time and distances involved in the migration of spilled vapors to the reported point of ignition. Both gasoline and LPG spills should be reviewed to make comparisons of the relative hazards. Based on this comparison, the need for improved practices should be evaluated. The use, for example, of additional chemicals, water treatment and emergency response training should be examined.

10.3 CONCLUDING REMARKS

The foregoing recommendations represent a measured approach to filling knowledge gaps in LPG release phenomenology and improving release prevention and control practices. A finite number of problems associated with LPG safety and environmental control appear to be worthy areas for further R&D. New projects are recommended in these areas to extend and complement existing knowledge and proven practices.

Contrary to views expressed by some of the news media, problems in LPG safety and environmental control warrant neither a crash program nor a new organization to deal with them. The continuing collaborative and coordinated efforts of the LPG industry and appropriate agencies of government appear to be all that is necessary to address these problems.

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APPENDIX A

U.S. REFINERIES INVOLVED IN PRODUCTION OF LIQUEFIED REFINERY GASES: LOCATION, CAPACITIES, TYPE OF PROCESSING

APPENDIX A

U.S. REFINERIES INVOLVED IN PRODUCTION OF LIQUEFIED REFINERY GASES: LOCATION, CAPACITIES, TYPE OF PROCESSING

Appendix A, a compilation of information on U.S. refineries involved in production of liquefied refinery gases, is derived from a survey published in Oil and Gas Journal (Cantrell 1978). Information given includes crude oil capacities and charge capacities [both expressed in terms of barrels per stream day (b/sd) and barrels per calendar day (b/cd)] for the following LRG producing operations and processes: thermal operations, catalytic cracking, catalytic reforming, catalytic hydrocracking, catalytic hydrorefining, and catalytic hydrotreating. The information is listed for specific facilities which are grouped according to the state in which they are located.

The following listing shows a variety of specific processes which are used in refineries for the production of LPG.

- Thermal Operations
  - -Gas-oil cracking
  - -Visbreaking
  - -Fluid Coking
  - -Delayed Coking
- Catalytic Cracking
  - -Fluid
  - -Thermofor
  - -Houdriflow
- Catalytic Reforming
  - -Semiregenerative
  - -Cyclic
- Catalytic Hydrocracking
  - -Distillate Upgrading
  - -Residual Upgrading
  - -Lube-oil Manufacturing
- Catalytic Hydrorefining
  - -Residual Desulfurizing
  - -Heavy Gas-Oil Desulfurizing
  - -Residual Visbreaking
  - -Cat-cracker and Cycle-stock (Feed Pretreatment)
  - -Middle Distillate
- Catalytic Hydrotreating
  - -Pretreating Catalytic, Reformer Feeds
  - -Naptha Desulfurizing
  - -Naptha Olefin or Aromatics Saturation
  - -Straight-run Distillate
  - -Lube-oil Polishing

APPENDIX A

U.S. REFINERIES INVOLVED IN PRODUCTION OF LIQUEFIED REFINERY GASES: LOCATION, CAPACITIES, TYPES OF PROCESSING

	Crude Capacity			Charge Capacity, b/sd					
				Cat Cracking					
Company & Location	b/cd	b/sd	Thermal Operations	Fresh Feed	Recycle	Cat Reforming	Cat Hydro-cracking	Cat Hydro-refining	Cat Hydro-treating
ALABAMA									
Hunt Oil Co.-Tuscaloosa	28,500	29,900				5,500		9,000	6,000 5,500
Mobile Bay Refining Co.-chickasaw	28,000	21,000				3,000			3,000
Mobil Bay Refining Co.-Chickasaw	28,000	30,000							
Totals	84,500	80,900				8,500		9,000	14,500
ALASKA									
Tesoro Petroleum Corp.-Kenai	38,000	40,000				6,000			6,000
Totals	38,000	40,000				6,000			6,000
ARKANSAS									
Cross Oil & Refining Co. of Arkansas-Smackover	8,600	8,750				1,200			
Tosco Corp. - El Dorado	47,000	48,300		15,500	3,000	5,750			7,500 3,300 1,100
Totals	55,600	57,050		15,500	3,000	5,750			13,100
CALIFORNIA									
Atlantic Richfield Co.-Carson	180,000	186,000	10,000 42,000 32,500	56,000	None	38,000	19,000		35,000 18,000 5,500 16,500
Beacon Oil Co.-Hanford	12,300	12,400	500 2,750			1,650			
Champlin Petroleum Co.-Wilmington	31,200	32,500	11,500						
Chevron U.S.A. Inc.-									
Bakersfield	26,000	NR	9,800			6,000			6,000
El Segundo	405,000	NR	54,000	52,000	8,000	60,000	49,000	24,000 60,000	45,000 11,000
Richmond	365,000	NR		55,000	5,000	84,000	45,000 30,000	14,000 60,000 65,000	18,000 82,000 3,200
Douglas Oil Co.-Paramount	46,500	48,000				11,500			12,000 7,000 10,000
Exxon Co.-Benicia	99,000	105,000	26,000	49,000	11,000	24,000	23,000	23,000	50,000
Fletcher Oil & Refining Co.-Carson	25,000	26,000				4,000			4,000
Golden Eagle Refining Co.-Carson	16,500	17,200							
Gulf Oil Co.-Santa Fe Springs	51,500	53,800	13,800 6,500	13,500	300	19,000 3,000	11,000		12,000 3,000

APPENDIX A (Continued)

Company & Location	Crude Capacity b/cd	Crude Capacity b/sd	Thermal Operations	Charge Capacity, b/sd Cat Cracking Fresh Feed	Cat Cracking Recycle	Cat Reforming	Cat Hydro-cracking	Cat Hydro-refining	Cat Hydro-treating
CALIFORNIA (Continued)									
Kern County Refinery Inc.-Bakersfield	15,900	NR	6,500	3,000					3,000
Mobil Oil Corp.-Torrance	123,500	131,100	16,100	60,000	None	35,500	21,700		21,000
Mohawk Petroleum Corp. Inc.-Bakersfield	22,100	22,800				2,500		2,500	2,500
Pacific Refining Co.-Hercules	85,000	NR				15,000	3,000		14,000
Powerine Oil Co.-Santa Fe Springs	44,120	46,000		11,500	300 6,000	1,500		8,000	10,000
Shell Soil Co.-Martinez	104,000	107,000		46,000	40,000	25,000	20,000	50,000	17,000 10,000 16,000 6,300
Wilmington	108,000	113,000	41,500	35,000	5,000	24,000			11,000 27,000 12,000 33,400
Sunland Refining Corp.-Bakersfield	15,000	15,000				1,000			1,500
Texaco Inc.-Wilmington	75,000	NR	48,000	28,000	NR	35,000	20,000	13,000	20,000
Tosco Corp.-Bakersfield	NR	40,000	7,000	12,000	0	15,500	14,000		7,000
Martinez	137,000	NR	37,000	47,000	0	8,700 21,300	20,000		1,400 34,500
Union Oil Co. of Calif.-Los Angeles	108,000	111,000	20,000	45,000	7,000	49,000	21,000		52,000 33,000
Rodeo	111,000	117,000	42,500			26,000	30,000		21,000 9,000 14,000
USA Petrochem Corp. Ventura	20,000	19,000				6,000			7,500
Totals	2,226,620	1,202,800	479,183	516,111	85,933	527,039	328,922	320,944	789,022
COLORADO									
Asamera Oil U.S. Inc.-Commerce City	NR	18,000		7,000	200	3,000			
Continental Oil Co.-Denver	10,000	11,000				6,500			7,000 8,500
Gary Western Co.-Fruita	13,000	14,000				2,000			2,000
Totals	23,000	43,000		7,000	200	11,500			17,500
DELAWARE									
Getty Refining and Marketing Co.-Delaware City	140,000	150,000	44,000	62,000	15,000	42,000	20,000		45,000 10,000 10,000 45,000
Totals	140,000	150,000	44,000	62,000	15,000	42,000	20,000		110,000

APPENDIX A (Continued)

				Charge Capacity, b/sd					
Company & Location	**Crude Capacity**		**Thermal Operations**	Cat Cracking		Cat Reforming	**Cat Hydro-cracking**	**Cat Hydro-refining**	**Cat Hydro-treating**
	b/cd	b/sd		Fresh Feed	Recycle				
HAWAII									
Chevron U.S.A. Inc.-Barber's Point	44,000	NR		22,000	NR				3,500
Hawaiian Independent Refinery Inc.-Ewa Beach	62,500	65,500				11,000			11,000
Totals	106,500	111,815		22,000	7,333	11,000			14,500
ILLINOIS									
Amoco Oil Co.-Wood River	110,000	115,000		38,000	4,000	12,300			15,600 17,000 3,000
Clark Oil & Refining Corp.-Blue Island	66,500	70,000		26,000	1,000	30,500	11,000		20,500
Hartford	57,000	60,000	13,000	28,000	1,000	9,200			10,000 10,000
Marathon Oil Co.-Robinson	195,000	205,000	2,800 19,000	38,000	400	12,400 35,000	22,000	6,000	22,000
Mobil Oil Corp.-Joliet	180,000	200,000	34,000	92,000	27,600	47,000		75,000	74,000
M. T. Richards Inc.-Crossville	700	727							
Shell Oil Co.-Wood River	283,000	295,000		94,000	0	22,000	33,500	27,000	64,000
Texaco Inc.-Lawrenceville	84,000	NR	9,000	34,000	NR	24,00			24,000 17,000
Lockport	72,000	NR	27,000	30,000	NR	19,000			9,000 10,000 17,000
Union Oil Co. of California-Lemont	151,000	NR	21,000	55,000	8,000	31,000			31,000 13,000 4,000 36,000 2.200
Wirebeck Oil Co. Inc.-Plymouth	1,800	NR	1,800						
Yetter Oil Co.-Colmar	1,000	1,053							
Totals	1,202,000	1,271841	131,600	442,111	63,332	315,178	66,500	108,000	500,456
INDIANA									
Amoco Oil Co.-Whiting	380,000	405,000	25,000	140,000	4,000	76,000			83,000 38,000 5,160 20,000
Energy Cooperative Inc.-East Chicago	126,000	140,000		48,000	2,000	20,000			20,000 25,000
Indiana Farm Bureau Cooperative Assn. Inc.-Mt. Vernon	21,500	NR		7,200	690	3,000			
Princeton Refinery Inc.-Princeton	4,600	NR				1,600			1,500

APPENDIX A (Continued)

	Crude Capacity		Charge Capacity, b/sd						
				Cat Cracking					
Company & Location	b/cd	b/sd	Thermal Operations	Fresh Feed	Recycle	Cat Reforming	Cat Hydrocracking	Cat Hydrorefining	Cat Hydrotreating
INDIANA (Continued)									
Rock Island Refining Corp.-Indianapolis	43,600	44,500		17,000	None	8,700			13,500
Totals	575,700	589,500	25,000	212,200	6,690	109,200			206,160
KANSAS									
CRA Inc.-Coffeyville	49,850	51,250	13,000	16,000	1,500	8,600		3,000	19,800
Phillipsburg	26,600	27,460		8,500	800	5,300			7,500
Derby Refining Co.-Wichita	25,000	27,650	3,800	10,800	1,700	5,000			5,000
Getty Refining and Marketing Co.-El Dorodo	80,577	82,000	11,500	31,000	17,000	21,500		40,000	23,000 4,300
Mobil Oil Corp.-Augusta	50,000	52,000	4,100	21,500	2,000	10,500 11,500			11,500
National Cooperative Refinery Assn.-McPherson	54,150	57,000	17,000	20,000	1,000	7,000			8,000
Pester Refinery Co.-El Dorado	21,800	22,500		11,000	500	4,000			4,000
Phillips Petroleum Co.-Kansas City	78,000	85,000		33,500	16,700	21,000			30,000 27,000 5,000
Total Petroleum-Arkansas City	42,500	47,200		9,600	1,200	16,300	3,000		16,300
Totals	428,477	452,060	49,400	161,900	42,400	110,700	3,000	43,000	161,400
KENTUCKY									
Ashland Petroleum Co.-Catlettsburg	135,800	140,000	4,000	55,000		26,500		40,000	26,500 6,500 4,500 40,000
Louisville	25,200	26,000		10,000		3,000			3,000
Somerset Refinery Inc.-Somerset	5,000	NR				1,000			
Totals	166,000	166,000	4,000	65,000		30,500		40,000	80,500
LOUISIANA									
Atlas Processing Co. Division of Pennzoil-Shreveport	45,000	47,400				10,000			10,000 1,800 3,300 4,800
Canal Refining Co.-Church Point	6,400	6,500				2,100			
Cities Service Co.-Lake Charles	291,000	NR	28,000	125,000	20,000	46,000		6,000] 30,000	46,000 14,000
Clairborne Gasoline Co.-Libon	6,500	6,700					2,200		
Continental Oil Co.-Lake Charles	87,000	90,000	7,000	25,500	5,000	18,500			19,000
Evangeline Refining Co. Inc.-Jennings	NR	5,000				600			

APPENDIX A (Continued)

			Charge Capacity, b/sd						
				Cat Cracking					
Company & Location	**Crude Capacity** b/cd	b/sd	**Thermal** Operations	Fresh Feed	Recycle	Cat Reforming	Cat Hydro-cracking	Cat Hydro-refining	Cat Hydro-treating
LOUISIANA (Continued)									
Exxon Co.-Baton Rouge	500,000	540,000	50,000	154,000	NR	83,000	25,000		2,600
Good Hope Refineries Inc.-Good Hope	86,000	95,000	60,000	165,000	500	14,500			
Gulf Oil Corp.-Belle Chasse	195,900	202,000	16,000	78,000	2,300	37,500		16,000 22,000	42,000
Marathon Oil Co.-Garyville	200,000	205,000				37,500		53,500 33,000	37,500
Murphy Oil Corp.-Meraux	92,500	96,500		10,500	500	23,000			29,000 15,000
Placid Refining Co.-Port Allen	34,200	36,000				5,500			6,000
Shell Oil Co.-Norco	230,000	240,000	18,000 29,000 17,900	10,000	2,000	18,000 28,000	24,000	25,000	29,000
Tenneco Oil Co.-Chalmette	NR	120,000	9,000	22,000	NR	35,000	18,000	18,000	24,000
Texaco Inc.-Convent	140,000	NR	12,000	70,000	NR	30,000			55,000
Totals	1,914,500	1,689,000	196,733	567,777	106,433	402,733	78,500	203,500	457,011
MICHIGAN									
Lakeside Refining Co.-Kalamazoo	5,600	NR				2,000			
Marathon Oil Co.-Detroit	65,000	67,000		25,500	1,300	16,000		12,500	16,500
Osceola Refining Co.-West Branch	12,500	10,000				1,600		1,600 1,150	
Total Petroleum Inc.-Alma	40,000	42,000		16,000	550	10,000		3,500 1,500	10,000
Totals	123,100	119,000		41,500	1,850	29,600		20,250	26,500
MINNESOTA									
Continental Oil Co.-Wrenshall	23,500	24,000		9,500	500	3,600			3,600
Koch Refining Co.-Rosemont	127,300	131,905	23,000	50,000	1,000	15,000		45,000	17,000 28,000 7,000
Northwestern Refining Co., Div. of Ashland Petroleum Co.-St. Paul Park	67,000	69,000		23,000	NR	12,000		20,000	13,000 7,200
Totals	217,800	224,905	23,000	82,500	8,400	30,600		65,000	75,800
MISSISSIPPI									
Amerada-Hess Corp.-Purvis	30,000	NR	7,000	16,200	NR	5,400			5,450
Chevron USA Inc. Pascagoula	280,000	NR		56,000	2,000	90,000	68,000	26,000	48,000
Totals	310,000	0	7,000	72,200	6,860	95,400	68,000	56,000	53,450

APPENDIX A (Continued)

Company & Location	Crude Capacity b/cd	b/sd	Thermal Operations	Charge Capacity, b/sd Cat Cracking Fresh Feed	Recycle	Cat Reforming	Cat Hydro-cracking	Cat Hydro-refining	Cat Hydro-treating
MISSOURI									
Amoco Oil Co.- Sugar Creek	109,000	111,000	13,500	42,000	12,000	16,000			21,000 2,500 38,000
Totals	109,000	111,000	13,500	42,000	12,000	16,000			61,500
MONTANA									
Cenex-Laurel	40,000	42,500		12,000	3,000	12,000		14,000	15,000
Continental Oil Co.- Billings	52,500	56,000		15,000	NR	15,800			16,000 9,500 12,900
Exxon Co.- Billings	45,000	46,000	7,000	19,200	14,500	14,500	4,900		15,500 10,000 10,000
Phillips Petroleum Co.- Great Falls	6,000	6,300		2,100	1,250	650			750
Westco Refining Co.- Cut Bank	5,300	6,000		2,200		2,300			2,300 1,000
Totals	149,200	156,800	9,200	48,300	23,750	45,250	4,900	14,000	93,800
NEBRASKA									
CRA Inc.-Scottsbluff	5,600	6,160		2,400	500	750			
Totals	5,600	6,160		2,400	500	750			
NEW JERSEY									
Chevron U.S.A. Inc.- Perth Amboy	168,000	NR		33,000	8,000	39,000		60,000	39,000 20,000
Exxon Co.-Linden	290,000	307,000		135,000	25,000	19,000		50,000	45,000 14,000 99,000
Mobil Oil Corp.- Paulsboro	98,000	100,500	23,700	25,000	None	23,500			23,500 33,000 6,100
Texaco Inc.- Westville	88,000	NR	13,000	40,000	NR	13,000			17,000 23,000
Totals	644,000	676,973	38,144	237,444	46,333	121,944		110,000	324,043
NEW MEXICO									
Navajo Refining Co.- North Artesia	NR	6,250	1,500			1,700 6,000			10,000 3,200 1,200
South Artesia	NR	23,750		5,600	400				
Plateau Inc.- Bloomfield	12,900	14,000		5,000	NR	2,250			2,250
Shell Oil Co.- Ciniza	18,000	19,000		7,200	3,600	6,800			6,800
Southern Union Refining Co.-Monument	5,000	5,200				650			
Totals	35,900	68,200	1,500	17,800	5,500	17,400			23,450
NEW YORK									
Ashland Petroleum Corp.-North Tonawanda	64,000	66,000		23,000	NR	11,500		20,000	27,000
Cibro Petroleum Products Inc.-Albany	28,000	30,000							

APPENDIX A (Continued)

Company & Location	Crude Capacity b/cd	Crude Capacity b/sd	Thermal Operations	Charge Capacity, b/sd Cat Cracking Fresh Feed	Charge Capacity, b/sd Cat Cracking Recycle	Cat Reforming	Cat Hydro-cracking	Cat Hydro-refining	Cat Hydro-treating
NEW YORK (Continued)									
Mobil Oil Cor.-Buffalo	43,000	44,000		19,000	6,000	11,500			12,500
Totals	135,000	140,000		42,000	12,900	23,000		20,000	39,500
NORTH DAKOTA									
Amoco Oil Co.-Mandan	52,000	53,000		24,000	7,200	8,200			10,000
Westland Oil Co.-Williston	4,658	5,000	1,100			2,000			1,600
Totals	56,658	58,000	1,100	24,000	7,200	10,200			11,600
OHIO									
Ashland Petroleum Co.-Canton	64,000	66,000		25,000	NR	11,000		22,5000 12,000	12,000
Gulf Oil Co.-Cleves	42,700	44,000		18,000	2,000	10,000		5,000	11,000
Toledo	50,300	51,000		19,800	2,000	11,000		5,500	11,000
Standard Oil Co. of Ohio - Lima	168,000	177,000	16,200	37,700	7,800	47,000	20,000		59,000
Toledo	120,000	126,000	11,200	55,000	16,500	40,700	35,000		37,000
Sun Co. Inc.-Toledo	125,000	130,000		50,000	7,500	25,000 16,000	26,000		27,500
Totals	570,000	594,000	27,400	205,500	43,300	160,700	81,000	45,000	157,500
OKLAHOMA									
Champlin Petroleum Co.-Enid	53,800	56,000	5,000	19,000	300	15,000			20,400
Continental Oil Co.-Ponca City	132,000	136,000	17,000	45,000	NR	31,000			31,000 7,000
Hudson Refining Co. Inc.-Cushing	19,000	19,814	4,000	7,500	3,000	4,500			4,500 2,000 1,000 1,000
Kerr-McGee Refining Corp.-Wynnewood	50,000	51,000		11,500	2,000	7,500	4,500		7,500 4,000
OKC Refining Inc.-Okmulgee	25,000	24,000		8,000	2,000				
Oklahoma Refining Co.-Cyril	14,000	14,700		6,700	1,675	1,125			1,125
Sun Co. Inc.-Duncan	48,500	50,000	12,000	25,000	10,500	8,000			8,000
Tulsa	88,500	90,000	8,200	30,000	1,400	23,000			24,000 11,000
Texaco Inc.-Tulsa	50,000	NR	6,000	18,000	NR	20,000			8,000 17,000
Vickers Petroleum Corp.-Ardmore	NR	64,500		21,500	1,000	12,000		20,000	12,000
Totals	480,800	506,014	52,866	194,200	42,875	124,347	4,500	20,000	162,301
PENNSYLVANIA									
Atlantic Richfield Co.-Philadelphia	185,000	195,000				56,000	30,000	24,000 50,000	54,000
BP Oil Corp.-Marcus Hook	164,000	177,000		48,000	1,600	50,000	25,000	48,000	64,000 20,000 21,000

APPENDIX A (Continued)

				Charge Capacity, b/sd					
				Cat Cracking					
Company & Location	Crude Capacity b/cd	Crude Capacity b/sd	Thermal Operations	Fresh Feed	Recycle	Cat Reforming	Cat Hydrocracking	Cat Hydrorefining	Cat Hydrotreating
PENNSYLVANIA (Continued)									
Gulf Oil Co.-Philadelphia	208,000	214,000		84,600	6,500	26,000 26,000		62,000	52,000
Kendall-Amalie Div., Witco Chemical Co.-Bradford	9,000	9,500							
						2,000			2,500
Pennzon Co. Wolf's Head Div.-Rouseville	12,000	12,500				3,600			4,500
Quaker State Oil Refining Corp.-Emlenton	3,320	3,500				1,250			1,450
Farmers Valley	6,500	6,800				1,860			2,300
Sun Co. Inc.-Marcus Hook	165,000	180,000		75,000	15,000	13,100 31,200			55,000 10,000 13,000
United Refining Co.-Warren	42,000	42,800		11,500	200	10,000			15,000 1,500
Totals	794,820	841,100		219,100	23,300	221,010	55,000	184,000	318,950
TENNESSEE									
Delta Refining Co.-Memphis	42,500	43,820		12,500	None	9,300			9,300 4,200
Totals	42,500	43,820		12,500		9,300			13,500
TEXAS									
American Petrofina Inc.-Big Spring	60,000	65,000	10,000	24,000	1,000	20,000		8,000	36,000
Port Arthur	90,000	110,000	10,000	34,000	2,000	22,000		30,000	22,000
Amoco Oil Co.-Texas City	415,000	432,000	33,500	184,000	33,000	134,000	42,000		139,000 50,000
Atlantic Richfield Co.-Houston	363,000	381,000	30,000	76,000	5,000	95,000		44,000 88,000 37,000	95,000 8,000 6,600
Champlin Petroleum Co.-Corpus Christi	155,000	159,000		65,000	NR	6,300 25,000		50,000	27,000 6,300
Charter International Oil Co.-Houston	65,000	70,000	10,000	40,000	NR	13,500			15,000 6,000 1,800 16,000 7,500
Chevron U.S.A. Inc.-El Paso	76,000	NR		22,000	NR	25,000		14,000 4,000	25,000
Coastal States Petrochemical Co.-Corpus Christi	185,000	NR	12,000	19,000	600	15,000 20,000		25,000	30,000 10,000 10,000
Crown Central Petroleum Corp.-Houston	100,000	103,000	9,500	50,000	0	8,000 14,000			22,000
Diamond Shamrock Corp.-Sunray	51,500	53,500	2,500	11,500 11,500	2,000 2,000	14,000			14,000

Company & Location	Crude Capacity		Thermal Operations	Charge Capacity, b/sd Cat Cracking		Cat Reforming	Cat Hydrocracking	Cat Hydrorefining	Cat Hydrotreating
	b/cd	b/sd		Fresh Feed	Recycle				
TEXAS (Continued)									
Dorchester Refining Co. -Mt. Pleasant	26,000	28,500		10,000	500	4,000			4,000 6,000
White Deer	NR	1,000				1,000			
Exxon Co. U.S.A.-Baytown	640,000	668,000		145,000	15,000	88,000 60,000	21,000	75,000 78,000 45,000	175,000 192,000 41,000 8,500
Gulf Oil Co.-Port Arthur	334,500	342,000	30,000	120,000	6,000	65,000	15,000	65,000	65,000
Howell Corp.-Corpus Christi	15,000	15,790				9,500	5,000		10,000
San Antonio	3,000	4,000				1,300			
Independent Refining Corp.-Winnie	16,000	15,360				5,000 2,700	3,000		8,000 300
LaGloria Oil & Gas Co.-Tyler	29,300	29,700	3,000 12,000	10,000	5,000	9,500			7,000
Longview Refining Co., Division of Crystal Oil Co.-Longview	8,827	9,000				5,500		3,000	4,000
Marathon Oil Co.-Texas City	66,000	68,000		38,000	1,000	8,000			
Mobil Oil Corp.-Beaumont	325,000	335,000	27,000	90,000 24,000	NR NR	102,000	29,000		85,000 5,000 116,000
Phillips Petroleum Co.-Borger	97,000	100,000		52,000	10,400	7,500 21,000			26,500 27,400 12,600
Sweeny	97,000	100,000	8,000	35,500	5,200	36,000			54,500
Quintana Refinery Co.-Corpus Christi	15,000	15,790				9,500	5,000		10,000
Shell Oil Co.-Deer Park	285,000	310,000	65,000 20,000	70,000	NR	28,000 40,000		50,000	71,000 35,000 7,000 85,000
Odessa	32,000	35,000		10,500	5,500	11,000			11,000
Sigmor Refining Co.-Three Rivers	22,800	24,000				8,500			8,500 2,000
South Hampton Refining Co.-Silsbee	20,500	22,500				4,000			
Southwestern Refining Co. Inc.-Corpus Christi	120,000	122,450		12,000	700	30,000		18,000	35,000 24,000
Sun Co. Inc.-Corpus Christi	57,000	60,000	7,700	25,000	6,500	13,000 11,000		12,500	3,200
Tesoro Petroleum Corp. -Carrizo Springs	26,100	27,500				3,000			3,000
Texaco-Amarillo	20,000	NR	4,000	8,000	NR	5,000			5,000
El Paso	17,000	NR	4,000	7,000	NR	3,500			3,500
Port Arthur	406,000	NR	18,000	135,000	NR	60,000	15,000		60,000 62,000 18,000

APPENDIX A (Continued)

				Charge Capacity, b/sd					
				Cat Cracking					
Company & Location	Crude Capacity b/cd	b/sd	Thermal Operations	Fresh Feed	Recycle	Cat Reforming	Cat Hydrocracking	Cat Hydrorefining	Cat Hydrotreating
TEXAS (Continued)									
Texas City Refining Inc.-Texas City	119,600	130,000	9,000	35,000	0	11,000			11,000
Union Oil Co. of Calif.-Beaumont	120,000	NR		38,000	4,000	36,000			36,000 6,000 7,000
Winston Refining Co. -Fort Worth	20,000	20,500		3,400	2,600	1,700			
Totals	4,499,127	3,857,590	328,088	1,422,066	249,133	1,120,612	136,667	646,500	1,906,103
UTAH									
Amoco Oil Co.-Salt Lake City	39,000	41,500		18,000	4,000	6,000			6,000
Caribou Four Corners Inc.-Woods Cross	7,050	7,400		2,000	NR		1,100		3,500
Chevron U.S.A.-Salt Lake City	45,000	NR	8,500	11,000 7,000	None 1,000	5,500		5,500	5,500
Husky Oil Co.-North Salt Lake City	25,000	26,000		4,400	2,500	5,000			6,000
Phillips Petroleum Co.-Woods Cross	24,000	25,000		8,400	2,600	4,700			11,000 1,600
Plateau Inc.-Roosevelt	8,000	8,500		5,200	None				
Wesreco Inc.-Woods Cross	NR	12,500				2,300			
Totals	148,050	120,900	8,500	56,000	10,700	23,500	1,100	5,500	33,600
VIRGINIA									
Amoco Oil Co.-orktown	53,000	55,000	15,000	28,000	5,000	9,500			9,500
Totals	53,000	55,000	15,000	28,000	5,000	9,500			26,500
WASHINGTON									
Atlantic Richfield Co.-Cherry Point, Ferndale	106,000	110,000	30,000			39,000	39,000	12,000	27,000
Chevron U.S.A. Inc.-Seattle	4,500	NR	5,000						
Mobil Oil Corp.-Ferndale	71,500	75,000	7,000	25,500	2,000	11,000 12,000			13,000 21,000
Shell Oil Co.-Anacortes	91,000	94,000		36,000	17,000	20,000		8,500	20,000 7,000 21,000
Texaco Inc. Anacortes	78,000	NR		30,000	NR	20,000			25,000 17,000
U. S. Oil & Refining Co.-Tacoma	21,400	NR				3,000			3,000
Totals	372,400	279,000	37,000	94,833	38,000	107,222	39,000	20,500	158,667
WEST VIRGINIA									
Pennzoil Co., Elk Refining Div. Falling Rock	4,900	5,200				2,000			2,500
Quaker State Oil Refining Corp.-Newell	9,700	10,000				2,860		4,440	3,060
St. Marys	4,850	5,000				1,300			1,450
Totals	19,450	20,200				6,160		4,440	7,510

APPENDIX A (Continued)

			Charge Capacity, b/sd						
				Cat Cracking					
Company & Location	Crude Capacity b/cd	b/sd	Thermal Operations	Fresh Feed	Recycle	Cat Reforming	Cat Hydrocracking	Cat Hydrorefining	Cat Hydrotreating
WISCONSIN									
Murphy Oil Corp.-Superior	40,000	46,800		9,700	1,000	10,000		5,800	10,000
Totals	40,000	46,800		9,700	1,000	10,000		5,800	10,000
WYOMING									
Amoco Oil Co.-Casper	47,000	48,000		13,000	1,500	5,800			7,100
Husky Oil Co.-Cheyenne	24,200	25,200		10,000	2,500	1,000 5,200			6,200 4,900
Cody	10,800	11,300		3,300	1,000	1,500			1,500 1,800
Little America Refining Co.-Casper	24,500	NR		6,500	4,000	3,750			5,000 3,750
Sage Creek Refining Co. Inc.-Cowley	1,000	1,200				500			
Sinclair Oil Corp.-Sinclair	49,000	50,000		17,700	1,200	9,700		12,200	13,000 12,000
Texaco Inc.-Casper	21,000	NR	4,000	7,000	NR	4,000		4,000	4,000
Wyoming Refining Co.-Newcastle	10,500	11,000		4,000	3,000				
Totals	188,000	146,700	4,444	62,278	15,533	31,894		16,644	59,694
GRAND TOTAL	15,955,302	13,926,128							

APPENDIX B

FACILITIES INVOLVED IN EXTRACTION OF NATURAL GAS LIQUIDS FROM NATURAL GAS

APPENDIX B

FACILITIES INVOLVED IN EXTRACTION OF NATURAL GAS LIQUIDS FROM NATURAL GAS

Appendix B is a listing of facilities engaged in extraction of NGL from natural gas. The information was derived from a survey published in Oil and Gas Journal (Cantrell 1979).

This listing includes the gas capacities and gas throughputs for each plant [measured as million cubic feet per day (MMcfd)], the processing methods, and average daily amounts of NGL produced based on the previous 12 months production (measured as 1000 gal per day).

The process methods in the third column are identified by the following numbers: 1--absorption, 2--refrigerated absorption, 3--refrigeration, 4--compression, 5--adsorption, 6--cryogenic/Joule-Thompson, 7--cryogenic/expander. Figures in parenthesis do not represent primary production and are not added into State totals.

APPENDIX B

FACILITIES INVOLVED IN EXTRACTION OF NATURAL GAS LIQUIDS FROM NATURAL GAS

Company, Plant, County, Location	Gas Capacity	Gas Throughput	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
ALABAMA							
Cities Service Co.-Citronelle, Mobile Co., 35-2n-3w	2.5	0.6	3			5.2	
Getty Oil Co.-Hatter's Pond, Mobil Co., 10-2s-1w	50.0	24.0	3	32.0		27.0	
Mallard Exploration Inc.-Big Escambia Creek	20.0	17.0	7	22.0		24.7	
Placid Oil Co.-Womack, Choctaw Co., 0-10n-2w	4.0	3.8	2	10.2		10.0	
TOTALS	76.5	45.4		64.2		65.9	
ALASKA							
Chevron U.S.A. Inc. -Swanson River LTS, Kenai Borough 4-7n-9w	20.0	18.0	2-3	12.1			
Marathon Oil Co.-Trading Bay, Kenai Borough 5-18n-15wsm	40.0	28.1	7				86.1
TOTALS	60.0	46.1		12.1			86.1
ARKANSAS							
Arkansas Louisiana Gas Co.-Hamilton, Columbia Co. 9-18s-21w	75.0	13.8	3	15.0	5.0	6.4	
CALIFORNIA							
Aminoil USA Inc.-Huntington Beach Orange Co.	15.0	5.4	3				1.6
Inglewood, Los Angeles Co.	25.0	4.4	2	7.5			
Arco Oil & Gas Co.-No. Coles Levee No. 8, Kern Co., 32-30S-25e MDB&M	190.0	35.0	2	74.8		42.0	
Belridge Oil Co. -Belridge, Kern Co., 27s-20e-26	80.0	41.0	1	3.6	0.3		
Chevron U.S.A. Inc.-Carpinteria, Santa Barbara Co., 33-4n-25w	23.0	10.4	3	1.4	3.2		
One-C, Kern Co., 1-32s-23e	45.0	19.0	1	22.3			
Seventeen-Z, Kern Co., 17-30s-22e	90.0	24.5	2	27.6			
Three-P, Kings Co., 3-22s-17e	100.0	14.8	2	16.5		7.3	
Getty Oil Co.-Buena Vista Hills, Kern Co., 8-32-24e	20.0	9.3	1	6.7			
Cymric, Kern Co., 26-29-22e	30.0	4.5	1	4.8			
Ventura Co., 27-3n-23w	20.0	11.5	2	19.1	13.7	5.9	
Signal Hill, Los Angeles Co.	12.0	7.0	3	4.0			25.0
Marathon Oil Co.-South Coles Levee, Kern Co., 3-31s-25e	80.0	30.0	1	33.8	4.9	8.5	
Santa Fe Energy Co.-Rincon, Ventura Co.	16.0	4.7	1				9.0
Shell Oil Co.-Molino Santa Barbara Co., 35-5n-31w	45.0	1.0	2	1.2			
Ventura, Ventura Co., 38-3n-23w	120.0	7.0	3	4.8	1.1	2.1	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
CALIFORNIA (Continued)							
Sun Gas Co.-Newhall, Los Angeles Co., 27-4n-17w	70.0	28.0	2	9.3			
Shiells Canyon Ventura Co., 4-4n-19w	10.0	NR	1	5.6		3.7	
Union Oil Co. of Calif. Bell, Los Angeles Co. 6-35-11w							
Coalinga Nose. Fresno Co., 7-20s-16e	46.0	52.8	1	17.2	16.6	7.4	
Dominguez, Los Angeles Co., 33-3s-9w	20.0	3.7	1	3.0			
Santa Clara Valley Ventura Co., 4n-18w	20.0	15.2	3	11.9			
Santa Maria, Santa Barbara 24-10n-34w	35.0	11.5	1	17.7		11.7	
Stearns, Orange Co., 7-3s-9w	20.0	7.8	1	15.3			
Warren Petroleum Co.-Yowlumne, Kern Co., 2-11n-22w	13.0	9.0	3				
TOTALS	1,148.5	407.5		308.1	39.8	91.6	35.6
COLORADO							
Amoco Product Co.-Peoria, Arapahoe Co., 4s-60w-33	10.0	8.0	7	12.9			18.7
Continental Oil Co. -Fruita, Meso Co. 34-9s-10w	20.0	18.0	5	9.2		7.1	
Crystal Oil Co.-Roggen Weld Co., 24-2n-63w	21.0	10.3	2	2.3			
Vallery Morgan Co. 15-3n-59w	3.0	3.0	2	4.3			
Excelsior Oil Corp.-Yenter, Logan Co., 2-8n-54w	10.0	3.4	2	6.2			
Matrix Land Co.-Piceance Creek, Rio Blanco Co., 15-25-96w	40.0	26.0	2	7.0			
Northwest Pipeline Corp.-Ignacio, La Plata Co., sw1/4-36-34n-9w	300.0	197.6	1	45.4		63.0	
Planet Engineers Inc. -McClave, Kiowa Co. sw4-32-20s-48w	7.5	3.3	2	3.6			
Sun Gas Co.-Denver Central, Arapahoe Co. 5-5s-62w	20.0	7.2	2	21.4			
Dragon Trail Rio Blanco Co., 35-2s-102w	22.5	11.2	2	10.3		6.1	
Texaco Inc.-Wilson Creek, Rio Blanco Co., 27-3n-94w	10.5	NR	3	22.3			
Trend Exploration Ltd. -Buck Peak, Moffat Co., se-sw9-6n-90w	2.0	1.0	3	0.8	1.6		
Union Oil Co. of Calif.-Adena, Morgan Co. 12-1n-58w	28.0	2.2	2	4.9		0.8	
TOTALS	494.5	291.2		149.6	1.6	88.4	18.7
FLORIDA							
Exxon Co. U.S.A.-Jay, Santa Rosa Co. 9643-5n-29w	112.0	136.0	7	359.2		188.8	
Florida Hydrocarbons Co.-Brooker, Bradford Co.	NR	547.0	NR	38.6		24.7	
TOTALS	112.0	683.0		397.8		213.5	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
ILLINOIS							
U.S. Industrial Chemicals Co., Div. of National Distillers & Chemicals Corp.-Tuscola, Douglas Co., Ficklyn Township	550.0	396.0	2	380.4	68.7		
KANSAS							
Amoco Product Co.-Ulysses, Grant Co. 5-29s-38w	325.0	354.0	1	112.2	27.3	95.5	
Andarko Production Co.-Cimarron, Seward Co. 26-33s-32w	15.0	17.0	2	10.0			
Interstate, Morton Co., 29-34s-43w	16.0	4.0	2	5.0			
Woods, Seward, 22-33s-34w	10.0	9.0	2	4.0			
Hutchinson, Reno Co., 22-23s-6w				(844.4)	(123.3)	(336.5)	
Spivey, Harper Co. 5-31s-82	70.0	30.6	1	25.6		9.3	
Wichita, Sedgwick Co., 17-28s-1e	130.0	76.0	1	37.7	11.9	28.0	
Getty Oil Co.-Medicine Lodge 13-32s-12w	30.0	6.0	1	5.2			
Minneola, Ford Co. 13-29s-25w	25.0	13.9	2	6.9			
Mobil Oil Corp. -Hickok, Grant Co., 31-28s-35w	210.0	134.4	1	16.0		26,7	
National Helium Corp. -Seward Co. 23-33-32w	1,000.0	632.2	3	182.0		105.0	
Northern Gas Products-Buston, Ellsworth Co. 31-17s-9w	950.0	NR	2&6-7	660.0	71.0	175.0	
TOTALS	2,781.0	1,273.1		1,067.3	110.2	439.5	
LOUISIANA							
Anchor Gasoline Corp. -Krotz Springs, Pointe Coupee Par., 40&41-6s-7e	50.0	NR	1				11.9
Arco Oil & Gas Co. -Bayou Sale St. Mary Par., 17-11s-9e	97.0	77.0	2	- 21.9		13.6	
Arkansas Louisiana Gas Co.-Bistineau Webster Par., 6-17n-10w·31-17n-9w	60.0	28.8	1				5.2
Beacon Gasoiline Co.-Webster Par. 26-21n-9w	72.0	20.0	2	15.0	6.0	7.0	
Lake Charles, Calcasieu Par., 19-10s-9w				(343.3)	(105.7)	(85.2)	
St. Amelia, St. James Par., 12-12s-16e	38.0	24.5	2	3.8		4.8	
Claiborne Gasoline Co.-Claiborne Par., 20-21n-4w	NR	44.6	2	28.5	16.0	13.0	
Continental Oil Co. -Acadia Par., 31-95-1w	290.0	49.1	2&7	160.0		97.0	108.0
Gillis, Calcasieu Par., 14-9s-8w	265.0	173.0	2&5	69.5	22.5	20.8	
Grand Chenier, Cameron Par., 2-39-40-15s-6w	750.0	570.0	2&7	107.5	38.2	26.2	
Kings Bayou, Cameron Par., 34-14s-7w	80.0	28.7	2	5.7	2.1	1.5	
Opelousas, St. Landry Par., 32-6s-4e	110.0	84.0	2	24.2		0.8	
Thibodaux, Lafourche Par. 35&36-15s-16e	60.0	9.0	2	5.0			
Venice, Plaquemines Par., 25-21s-30e	1,000.00	609.0	2	144.0	43.0	37.0	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
LOUISIANA (Continued)							
Dubach-Calhoun complex, Lincoln Par. 526&34-20n	175.0	NR	2	66.0	25.0	33.2	
Marathon Oil Co.-Cotton Valley, Webster Par., 26-21n-10w	220.0	37.4	2	35.6	13.9	12.4	
Mobil Oil Exploration & Producing Southeast Inc.-Cameron Par. 23-15s-13w	470.0	429.0	267				90.4
Riverside, Ascension Par., 49-95-se				152.8	65.1	40.0	
Norco fractioner St. Charles Par. 6-12s-8e				(371.1)	(119.5)	(133.0)	
Tebone fractionator, Ascension Par., 8&46-10s-2e				(275.2)	(83.6)	(71.4)	
Weeks Island, Iberia Par. 13-14s-6e	129.0	59.0	2	12.0			
Cocodrie, Evangeline Par., 35-2s-2e	50.0	NR	2	22.8		16.8	
St. Landry, Evangeline Par., 35-2s-2e	60.0	NR	2	27.6		20.6	
Delhi, Richland Par.,15-17-9e	22.0	22.0	3	24.2	12.5	6.6	
South Sarepta, Bossier Par., 21-22n-11w	25.0	13.5	2	4.6		5.8	
Lowry, Cameron Par.,16-12s-4w	300.0	282.0	267	82.6	32.1	21.0	
Tenneco Oil Co.-Stephens, Claiborne Par.,6&7-13s-12e	35.0	18.9	2	2.2		2.4	
Fordoche, Pointe Coupee Par., 28-6s-8e	30.0	NR	2	22.0		16.0	
Henry, Vermilion Par.,21-13s-4e	825.0	NR	2	283.5	93.0		
Paradis, St. Charles Par., 29-14s-20e	800.0	NR	2	429.0		237.0	
Sea Robin, Vermilion Par., 21-13s-4e	900.0	NR	7	198.0	71.0	69.0	
Texas Gas Exploration Corp.-Eunice, Acadia Par.	1,100.0	593.4	2	225.6	69.6	62.0	
Union Texas Petroleum-Rayne Acadia Par.	750.0	627.4	2	85.2	22.7	28.0	
Sligo, Bossier Par.,	290.0	33.9	2	12.5	4.3	5.2	
Toco, St. Bernard Par.	190.0	76.2	2	44.3	15.0	15.3	
Krotz Springs, St. Landry Par., 22-6s-7e	91.0	45.0	2	53.0			
TOTALS	9,334.0	3,925.4					
MICHIGAN							
Marathon Oil Co.-Scipio, Hillsdale 2-5s-3w	38.0	18.4	2	32.5			27.1
Michigan Wisconsin Pipeline Co.-Loreed Osceola Co. 30-18n-10w	54.0	34.0	2	4.8	0.7	2.5	
Mobil Oil Corp.-Aurelius, Ingham Co., 36-2n-2w	23.0	15.8	2	15.6			
TOTALS	115.0	68.2		52.9	0.7	2.5	27.1

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
MISSISSIPPI							
Getty Oil Co.-Bay Springs, Jasper Co., 27-2n-10e	10.0	2.7	3				6.6
Shell Oil Co.-Goodwater, Clarke Co., 5-10-8w	15.0	4.9	3	2.7		1.9	
Tallahala Creek Smith Co. 5-1n-9e	10.0	3.5	4	1.0		2.0	
Texas Oil & Gas Corp.-Harmony, Clarke Co., 26-2n-14e	30.0	12.0	3	12.0		13.0	
TOTALS	65.0	23.1		15.7		16.9	6.6
MONTANA							
Tule Creek, Roosevelt Co., 14-30n-48e	2.5	0.3	3	0.3			
McCulloch Gas Processing Corp.-Fairview, Richland Co., 10-24n-57e	6.0	3.5	2	7.0		2.0	
PGP Gas Products Inc. -Mon-Dak, Richland Co.	4.0	2.0	3	4.0		6.0	
Thunderbird Resources Inc.-Westco Refining Glacier Co., 22-33n-5w	30.0	18.5	1	10.4		10.9	
True Oil Co.-Bob Rhodes, Richland Co. 4-25n-58e	3.0	0.8	3				5.0
TOTAL	45.5	25.1		21.7		12.9	11.0
NEBRASKA							
Cities Service Co. Kimball, Kimball Co., 10-12n-55w	10.5	1.1	1	2.6			
Marathon Oil Co. -West Sidney, Cheyenne Co. 4-12n-50w	12.5	3.9	2	4.2		2.4	
TOTALS	23.0	5.0		6.8		2.4	
NEW MEXICO							
Amoco Production Co.-Empire Abo, Eddy Co., 3-18s-17e	42.5	37.3	6-7	58.5		39.4	
Cities Service Co.-Bluitt, Roosevelt Co., 15-8s-36	37.0	30.8	2	44.3	5.5	12.1	
Continental Oil Co.-Maljamar, Lea Co. 21-17s-32e	28.0	11.4	2	28.5		14.5	
JAL No. 4B, Lea Co. se4-se4-31-23s-37e and s2-sw4-32-23s-37e				(82.3)		(102.0)	
San Juan River, San Juan Co., 1-29n-15w	71.0	41.8	1	10.5		14.3	
Wingate, McKinley Co. 16 & 17-15n-17w				(318.0)	(107.0)	(210.5)	
Gas Co. of New Mexico, Division of Southern Union Co.-Avalon, Eddy Co., 9-21s-27e	30.0	14.0	2				11.9
Getty Oil Co.-Eunice No. 1, Lea Co., 27-22s-37e	100.0	76.6	7	63.7	11.0	31.8	
Northern Natural Gas Co.-Hobbs, Lea Co., 6-19-37e	220.0	186.8	1				174.9
Lybrook, Rio Arriba Co. 14-23n-7w	70.0	61.6	NR	1.9			
Monument, Lea Co., 36-19s-36e	77.0	63.0	7	24.0			
TOTALS	675.5	534.3		231.4	16.5	112.1	186.8

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
NORTH DAKOTA							
Aminoil USA Inc.-Tioga, Williams Co., 26-157n-95w	105.0	72.7	2	76.2	50.2		
nergy Operating Corp. -Lignite, Burke Co., nw4-7-162n-91w	20.0	2.9	1	6.6		4.6	
Kerr-McGee Corp.-Boxcar Butte, McKinzie Co. 21s-148n-102w	6.0	NR	3	3.0		2.4	
True Oil Co.-Red Wing Creek, McKensie Co.27-148n-101w	10.0	3.0	2	7.5		5.0	
Williston, Williams Co., sw4-se26-154n-102w	3.0	0.6	3				2.8
Warren Petroleum Co.-Little Knife, Billings Co.16-141n-99w	20.0	13.0	3	2.0		1.0	
TOTALS	164.0	97.0		95.3	50.2	13.0	2.8
OKLAHOMA							
Aminoil USA Inc.-Aline, Alfalfa Co., 7-23n-11w	45.0	23.2	2	15.2			
Fox, Carter Co., 28-2s-3w	70.0	41.6	7		7.6	23.4	
Anadarko Production Co.-North Richland Center, Texas Co. 33-6n-12ecm	25.0	22.0	2	20.0			
Drumright, Creek Co. nw1/4-sw1/4-28-18n-7e	15.0	5.0	1	13.8			
Seminole, Seminole Co. sw1/4-ne1/4-10-8n-6e	20.0	9.1	1	19.0			
Champlin Petroleum Co.-Enid, Garfield Co.	62.0	29.0	2				73.8
Witcher, Oklahoma Co.	15.0	7.3	3	8.4		3.6	
Chevron U.S.A. Inc.-Marietta, Love Co., 31-7s-3e	50.0	6.0	2	4.1		3.6	
Cities Service Co. -Ambrose, Kay Co., 9-27n-1w	375.0	314.6	2	100.6	28.9	68.3	
Rodman, Garfield Co. 17-10n-8W	90.0	37.2	7	48.8	6.7	16.5	
Continental Oil Co.-Hennessey, Kingfisher Co., 36-19n-7w	30.0	25.0	7	31.8	3.9	10.0	
Medford Grant Co. 32-27n-5w	30.0	18.0	2	6.6	2.0	4.3	
CRA Inc.-Lamont, Grant., 20-26n-3w	12.5	4.4	2	4.1		2.8	
Dorchester Gas Producing Co.-Hooker, Texas Co., 8-4n-17ecm	75.0	70.0	7	60.6	8.6	20.3	
Exxon Co. U.S.A.-Camargo, Dewey Co. 10-18n-19w	15.0	11.0	2	5.4	4.4		
Grimes Gasoline Co. -Okemah, Okfuskee Co. 23-11n-9e	1.0	0.6	1				0.5
ITT Eason Oil Co.-Crescent, Logan Co., sw4-34-17n-4w	35.0	30.0	2	30.0	6.0	18.5	
Kerr-McGee Corp.-Milfay, Creek Co. 21-15n-7e	12.0	NR	2	4.4		1.0	
Koch Oil Co.-Fitts, Pontotoc Co. 30-2n-7e	3.5	2.0	3	1.6			
NE Trail, Dewey Co., 1-17n-18w	25.0	15.0	2	23.7			

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
OKLAHOMA (Continued)							
Postle Hough, Texas Co., 13-5n-13ecm	18.5	6.5	2	15.1			
Putnam-Oswego Dewey Co., 35-16n-16w	50.0	42.5	6	21.0			
Seiling, Woodward Co. 32-20n-17w	20.0	9.0	3	(All products fractionated at N.E. Trail)			
Sholem Alechem, Stephens Co., 2-1s-4w	70.0	42.7	6	7.4			
West Putnam, Dewey Co. 9-17n-17w	10.0	2.0	3	(All products fractionated at N.E. Trail)			
Mustang Gas Products Co.-Calumet, Canadian Co., ne4-28-14n-9w	250.0	175.0	2	82.7			
Pioneer Gas Products Co.-Binger, Caddo Co. 26-10n-11w	15.0	4.7	7				17.0
Madill, Marshall Co., 32-7e-5s	27.0	19.8	2	17.0	1.6		
Ringwood, Major Co. 11-22n-10w	80.0	60.0	2&7				125.6
Shell Oil Co.-Seiling Dewey Co., 4-19n-17w	75.0	32.0	2	28.0			
Sohio Petroleum Co.-Elmore, Garvin Co., 17-2n-1e	70.0	64.0	2	8.0		20.0	
Laverne, Harper Co. 20-26n-25w	225.0	164.0	7	81.3	17.7	42.2	
Texaco Inc.-Apache Caddo Co. 2-5n-12w	7.5	NR	3	3.0			7.0
Enville, Love Co., 7-7s-3e	23.0	NR	2	10.3		6.5	
Union Oil Co. of Calif.-Caddo, Carter Co.,23-3s-1e	10.0	6.2	3	6.8			
Warren Petroleum Co.-Knox, Grady Co.,33-3n-5w	50.0	21.0	2	7.0			
Maysville, Garvin Co., 18-4n-2w	60.0	55.0	7	4.0	8.0	27.0	
Mocane, Beaver Co. 18-5n-25e-ecm	175.0	101.0	2		6.0	27.0	
TOTALS	2,213.5	1,476.4		687.0	101.4	293.6	222.5
TEXAS							
Aminoil USA Inc.-Birthright, Hopkins Co.	30.0	11.2	3	8.0			
Amoco Gas Co.-Texas City Extraction, Galveston Co. John Grant A-72	140.0	93.2	3	15.5			
Amoco Production Co. -Burnell-North Pettus, Bee Co. A-591	180.0	58.9	1	16.4		8.3	
Edgewood, Van Zandt Co.,7 Roberts A-702	60.0	55.7	2	17.5		22.0	
Hastings, Brazoria, 1-ACH-D A-416.	70	87.1	6&7	20.2		15.7	
Levelland, Hockley Co., Labor 7, League 72, Val Verde School Land	40.0	17.8	1	54.1		27.0	
Luby, Nueces Co., Canutillo Colony Dutch Co.	90.0	21.7	2	5.6		6.4	
Midland Farms. Andrews Co., 8-42 T-T-N G&MMB&A	45.0	12.1	6&7	28.8	5.3	12.7	
Monahans, Winkler Co.,24-10-PSL	5.0	5.7	3	2.5	0.4	1.5	
North Cowden, Ector Co.,34-35-43-1n T&P Ry	45.0	27.9	6&7	61.6		40.4	
Old Ocean, Brazoria Co., Charles Breen League A-46	570.0	238.5	2	79.9	23.5	22.2	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
TEXAS (Continued)							
Prentice, Yoakum Co., 20K PSL	6.0	5.3	1	2.1			3.6
Slaughter, Hockley Co., 14-15-59 Edwards & Scurry CSL	80.0	41.6	1	16.1			0.9
South Fullerton, Andrews Co., 8-A 48-PSL	10.0	9.9	667	37.6	5.1	15.1	
Anchor Gasoline Corp. -Tabasco, Hidalgo Co. NW corner of Tract 322, Las Ejidas de Reynosa Vieja Grant	67.0	NR	1				6.1
Arco Oil & Gas Co. -Block 31, Crane Co. 33-31 Univ. Lands	180.0	180.6	2	125.6		73.6	
El Dorado, Schleicher Co. 81-TT-TCRR	56.0	22.1	1	34.0		13.6	
Fashing, Atascosa-Karnes Cos., 131 Wm. Smith	12.0	5.2	1				0.8
Hull, Liberty Co., Wm. Smith A-342	18.0	3.0	2	2.4			
Longview, Gregg Co., J. Moseley	35.0	17.0	266	90.9			
Nueces River, Live Oak Co., Cameron CSL 32-A	90.0	29.0	2	15.0		9.0	
Price, Rusk Co. J. B. Cadena	15.0	1.0	1	8.5			
Silsbee, Hardin Co., George W. Brooks A-4	50.0	13.0	2	8.8	0.2	3.5	
Taft, San Patricio Co., 48&48A Coleman Fulton Pasture Lands	40.0	26.0	2	6.7		5.0	
Waskom, Harrison Co., J. Blair	205.0	46.4	2	17.0	7.0	8.5	
Blackhawk Gasoline Corp.-Jack Co., 9 mi. E of Graham	1.5	0.5	3				0.6
Breckenridge Gasoline Co.-Eliasville, Stephens Co. 1174 TE&L A-503	5.0	2.1	2	1.9		1.0	
East Texas, Panola Co.	220.0	202.4	7	87.9		56.7	
Gulf Plains, Nueces Co.	135.0	130.0	7	46.4		29.8	
Kermit, Winkler Co.	50.0	17.0	2	5.0		8.9	
North Snyder, Scurry Co.	44.0	43.0	3	163.3	18.6	62.2	
Sherman, Grayson Co.	40.0	22.0	2	15.6			17.7
Sivells Bend, Cooke Co.	5.0	1.0	1	0.7			0.9
Cities Service Co.-Chico, Wise Co., CH&HRR A-384	65.0	57.0	2	115.0	5.0	43.5	
Corpus Bay, San Patricio Co., Lot A Gregory subdiv. Geronimo Valdez A-296	75.0	46.4	2	19.9	10.3	5.5	
East Texas, Gregg Co., Wm. Castleberry A-38	27.0	18.1	2	74.8		53.7	
Mont Belview fractionation, Chambers Co., Henry Griffin League A-12				(858.7)	(118.8)	(268.3)	
Myrtle Springs, Van Zandt Co., J. Salnga A-765	30.0	11.3	2	5.2		6.2	
Pampa, Gray Co., 133 & 136 IGNRR	50.0	6.8	2	14.5	0.9	5.8	
Panola Co., Matthew Parker A-527	100.0	18.2	2	5.4	2.7	1.8	
Robstown, Nueces, Simmons & Perry's subdiv. of Ellifer Tract	65.0	20.7	1	4.6	2.0	2.3	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
TEXAS (Continued)							
San Antonio Bay, Calhoun Co. Lot 2 Miguel Castillio A-7	12.4	2.5	2	1.6	0.7	0.7	
Stonewall Co., E. Borden, A-831	20.0	3.8	2	17.7		10.0	
West World, Crockett Co., 19 O-GCSFRR	15.0	2.8	2	2.7		2.4	
Frear, Webb Co.	190.0	81.0	2	12.5		19.1	
Hidalgo Co.	80.0	10.0	2	1.6		2.0	
Mission, Hidalgo Co.	30.0	23.0	2				13.2
Coates, George H. Estate of-Jay Simmons, Starr Co., San Jose Grant	5.0	2.0	2				3.0
Continental Oil Co.-Chittim, Dimmit Co., 120 miles sw of San Antonio, 24 miles w of Carrizo Springs	5.0	2.2	3	4.6		4.7	
Hamlin, Fisher Co., 50 miles nw of Abilene, T&TC-1	20.0	10.5	3	24.6	3.7	10.5	
Port, Port Arthur	150.0	5.7	2	1.1	0.4		
Ramsey, Reeves Co., 36-38-1, 8 miles nw of Orla. 100 miles w of Odessa	10.0	2.8	3	2.8		3.5	
Rincon, Starr Co., 485-CCSD R6N6RR, 40 miles nw of McAllen	26.0	13.1	2&5	4.4		4.0	
Quitman, Wood Co., S.G. Purse A-456	30.0	4.6	1	2.5		11.7	
Davis, J.L.-Bowie, Montagu Co.	NR	3.5	1	7.0			
Delta Drilling Co.-Ozona. Crockett Co., ne1/4 L MN A-2310 GC&SFRR	50.0	45.0	2	45.0		35.0	
Diamond Shamrock Corp.-McKee, Moore Co., 399-44 H&TC	375.0	327.0	2&7	287.0	55.9	119.1	
Dorchester Gas Producing Co.-Cargray, Carson Co.,46-4-I&GN White Dear	100.0	30.0	1	15.5	6.1	14.5	
Woodlawn, Harrison Co., L. Watkins	100.0	10.0	1	3.6		3.0	
El Paso Natural Gas Co.-Midkiff, Regan Co., nw46n2-sw4 22-T&PRR 37-5-5	168.0	75.8	1	71.5		73.2	
Etexas Producers Gas Co.-Chapel Hill, Smith Co. Stilwell Box A-169	6.0	3.0	1			3.3	
Exxon Co. U.S.A.-Anahuac. Chambers Co., H&TC RR 51-A-112	290.0	206.7	2	37.9	15.5	12.3	
Clear Lake, Harris Co. James Lindsey	220.0	224.7	2	64.5	20.8	17.6	
Conroe, Montgomery Co., Random House	120.0	85.4	7	53.2	12.6	21.6	
East Texas, Rusk Co., T. J. Martin	25.0	13.5	3	58.3	13.0	29.7	
Hawkins, Wood Co., H. Watson	150.0	119.9	7	82.8	38.7	49.0	
Jourdanton, Atascosa Co. Edward Estes	26.0	19.0	1	3.4			4.0
Katy, Waller Co., T. S. Reese 110 A-322	1,260.0	696.4	2	167.1	42.1	47.4	
Kellers Bay, Calhoun Co., N. Cavasaos A-2	47.0	7.0	2	3.0	1.2	0.1	
Kelsey, Brooks Co., LaBlanca Grant A-459	250.0	79.0	2	33.1	10.3	10.8	
King Ranch, Kleberg Co., R. King 172	2,650.0	1,5[illegible].0	2	230.0	93.9	80.2	
Neches, Cherokee Co. J. H. Shaw	40.0	[illegible].1	2	31.1	6.3	17.2	
NE Loma Novia, Duval Co., J. Poitevent [illegible] A-923	46.0	9.0	2	5.2	1.5	1.8	

Company, Plant, County, Location	Gas Capacity	Gas Through-out	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
TEXAS (Continued)							
Pledger, Brazoria Co. W. C. Carson	210.0	231.8	7	34.7	10.6		
Santa Fe, Brooks Co., San Salvador del Tule A-290	47.0	8.0	2	2.0	0.9	0.6	
Sarita, Kenedy Co., J.A.Balli A-2	255.0	64.0	2	15.9	5.6	4.9	
Thompson, Fort Bend Co., John Rabb	40.0	30.8	7	7.4	1.2	1.9	
Tomball, Harris Co.,C.Goodrich	80.0	69.5	2	17.0	5.0	6.1	
Tom O'Connor, Refugio Co., Maria Ximines A-324	150.0	133.0	1	15.6		24.2	
Getty Oil Co.-East Vealmor, Howard Co., 20-27-H&TC	50.0	46.0	2	110.6	15.0	50.6	
New Hope, Franklin Co., Isaac Barre A-20	50.0	32.0	1	12.3		14.2	
Normanna, Bee Co., Thomas Duty A-21	32.0	11.0	2			4.6	
Shafer, Carson Co. 88-4 I&GN RR	30.0	17.3	1	27.3	10.9	29.0	
Spearman, Ochiltree Co., 23-RBB	50.0	52.1	2	28.2			
West Bernard, Wharton Co., J.M. Rose Heirs A-322	30.0	10.0	1	1.1		2.3	
Rio Grande, Starr Co.	32.0	30.0	2	7.5		6.0	
Runge, Karnes Co.	52.0	25.0	2	16.0	2.5	4.0	
HNG Petrochemicals Inc.-Bammel, Harris Co., HT&BRA A-420	100.0	27.3	2	6.7		2.8	
Gregory, San Patricio Co., Geranimo Valdez A-269	70.0	27.5	2	19.3		13.6	
Robstown, Nueces Co., Mathis Garcia A-116	75.0	25.4	2	10.5		8.8	
Tuleta, Bee Co., Brooks & Burleson	45.0	16.8	1				11.0
Victoria Co., James Reed A-236	94.0	19.1	2	4.7		4.5	
LoVaca Gathering Co.-Corpus Christi, Nueces Co.	200.0	162.0	2	50.0	15.0	15.0	31.0
Gholke, Dewitt Co.	125.0	134.0	2				46.0
Marathon Oil Co.-Markham, Matagorda Co., 4-9-9	165.0	94.0	2				62.3
Susan Peak, Tom Green Co., 193 SP RR Dist. 11	3.0	3.0	3	2.7			3.2
Welder, San Patricio Co., 49 Robert Montgomery 199 and Ewan Cameron A-97	55.0	16.2	1				4.2
Yates, Pecos Co., I.G.Yates 194 SF 1234-1	25.0	19.0	2	6.5			29.9
Desdemona, Eastland Co., J.W. Carruth Farm, W.H.Fundenburg A-139	1.3	0.9	3				2.4
Electra, Wilbarger Co.,17-13 H&TC RR	1.4	0.9	3				7.9
LaGloria, Jim Wells Co., 9-3 Ragland & Herring subdiv. of LaGloria Townsite subdiv.	318.0	185.8	7	86.0	26.3	21.0	
Seeligson, Jim Wells Co.,Los Jaboncillos Grant, Antonio Ramirez	318.0	171.0	7	45.0	13.5	11.4	
Wilcox, Lavaca Co., J.R.Ragsdale A-377	255.0	49.3	2	29.9		14.2	
Monsanto Co.-Diamond "M", Scurry Co., se1/4-182-97 H&TC	41.0	22.0	3	105.0		50.0	
Martin County No. 1, 31-37-2n T&PR	10.0	3.8	3				20.0

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
TEXAS (Continued)							
Odessa Natural Corp.-Foster, Ector Co., 18-42-2s T&PRR	24.0	NR	7	53.3		26.3	
Parade Co.-Giles Rusk Co.	7.5	3.5	2&4	20.0			
Permian Corp.-Alba, Wood Co., Lat.32-46-52 Long. 95-37-08	2.0	1.0	4				8.9
Possum Kingdom, Stephens Co., E. Romerhausen A-149	5.0	1.5	3				4.6
Todd, Crockett Co., 28-WX GC&SF	5.0	2.0	3				13.7
Petroleum Corp. of Texas-Ibex, Shackelford Co. nw28-3AL	10.0	6.0	1	11.4			
South Bend, Young Co., J. Garrett	8.0	6.0	1	14.4			
PGP Gas Products Inc. -Imperial, Crane Co. 21-11H&GN RR	18.0	13.0	7				42.0
Pioneer Gas Products Co.-Arrington, Hemphill Co. A-2-62	40.0	28.7	2				40.2
East Goldsmith, Ector Co.,34-34	50.0	11.2	2&7				29.1
Fain, Potter Co., G&M 10-1818	130.0	70.3	2	37.6		46.4	
Pamp, Gray Co., H&GN 96 B-2	60.0	9.7	2	5.5			
Turkey Creek, Potter Co., G&M 36-M2	100.0	47.9	2	33.1		35.4	
Shell Oil Co.-Bryand Mill, Cass Co., B. F. Lynn A-651	70.0	56.9	2	42.1			
Conley, Hardeman Co. 80-H W&NW RR	6.0	1.0	3	2.3		0.8	
Houston Central, Colorado Co., F. Mayher A-400 K. Winn A-589	425.0	157.9	2	102.9	27.4	32.9	
NW Ozona, Crockett Co., 46-OP-GC&SF RR	10.0	8.0	3				16.0
Person, Karnes Co., Jesus Hernandez, A-140	54.0	25.1	2	25.3		10.0	
Tippett, Crockett Co., 28-31-H&TC RR	75.0	44.0	2&7	44.0			
TXL, Ector Co., 17-45-1 ST&P RR	65.0	34.0	1	56.0		27.0	
Wasson, Yoakum Co., 827-D J.H.Gibson	175.0	136.0	1&2	410.0			
Suburban Propane Gas Corp.-Martha F. Berry, Frio Co., M.C.Patton, 1173 A-542	22.0	5.0	2	4.1		2.0	
Sun Gas Co.-Big Wells, Dimmit Co., I&GN RR A-223 82-1	20.0	32.6	2	22.2		21.0	
Concho Co., 153-72-T&N	10.0	4.0	2	3.0			
Jameson, Coke Co. 315-1s H&TCRR	60.0	39.1	2	73.2	9.7	25.4	
Luby, Nueces Co., 9-G Part Petronilla Ranch	10.0	6.6	2				5.3
Snyder, Scurry Co., 16-1 J.P.Smith	150.0	90.3	3	169.0	29.5	83.0	
Sun, Starr Co. 239-AB-225 CCSdrRqn Grr	92.0	75.0	2	36.9		26.4	
Tijerina-Canales, Jim Wells Co., 343-CCSD&RGNG	75.0	32.7	2	5.7		10.1	
Superior Oil Co. -Portilla, San Patricio Co., J. Francisco-E. Portilla A-53	15.0	12.0	2				8.5
Tenneco Oil Co.-Chesterville, Colorado Co., Thomson A-708	55.0	22.4	2	14.5	4.0	4.8	
Ward, Hidalgo Co., Porcion 68, Gregario Comacho A-28	140.0	34.2	2	8.6		7.1	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
TEXAS (Continued)							
Texaco Inc.-Blessing, Matagorda Co., 59-C J.E. Pierce sub H&GN	65.0	NR	2	20.0		15.0	
Fuller, Scurry Co., 642-97-H&TCRR	58.0	NR	2	58.0		46.0	
Lamesa, Dawson Co., 36-34-5n T&PRR	6.0	NR	3	10.0			
South Kermit, Winkler Co., 22-22 B-3 PSL	35.0	NR	2	7.3			
East Texas, Marion Co., John H. Kennedy A-235	75.0	65.0	7	25.0		14.0	
Fort Trinidad, Houston Co., RCS-A23	40.0	14.7	2	13.6	7.2		
Van, Van Zandt Co., JWS a-891	15.0	11.5	2	16.1		24.0	
Union Texas Petroleum -Benedum, Upton Co.				30.1		13.5	
United Texas Transmission Co.-Galveston Chambers Co., J. Armstrong A-2	40.0	17.0	7				20.3
Upham Oil & Gas Co.-Upham, Wise Co., Wm. Hersee A-367	10.0	2.0	2	5.6		3.7	
Bridgeport, Wise Co., J.J. McBride Tract out of P. Nicholas A-654	160.0	116.0	2	106.0	19.0	34.0	
Como. Hopkins Co., Nacagdoches Univ. A-703	30.0	5.0	2	8.0			
Fashing, Atascosa-Karnes Cos., Emma Tar H. Tract in the James Wilkerson 149 A-882	100.0	40.0	2	14.0			
Waddell, Crane & Ector Cos. 25-B-25	105.0	69.0	2	13.0			
Worsham, Ward Co. 56-34-H&TC RR	15.0	12.0	7	19.0			
TOTALS	15,314.1	8,384.8		4,693.0	606.6	1,928.5	460.6
UTAH							
El Paso Natural Gas Co.-Aneth, San Juan Co., nw4-6-41S, 24e	100.0	18.2	1	16.2			
Gary Operating Co. -Altonah, Duchesne Co., sw-nw5-2s-3w	12.5	4.4	3	4.4		3.9	
Bluebell, Duchesne Co., sw-nw23-1s-2w	23.0	15.8	2	17.6			
Koch Industries Inc. -Cedar Rim, Duchesne Co., 21-3S-6w	10.0	8.0	3	5.6		4.5	
Quasar Energy Inc.-Pineview, Summit Co. nw4-sw-4 3-2n-7e	10.0	14.5	3	40.2			
Shell Oil Co.-Altamont, Duchesne Co., 34-1s-4w	40.0	17.0	3	11.3		12.0	
Union Oil Co. of California-Lisbon, San Juan Co. 22-30s-24e	80.0	53.1	3	36.0		21.8	
TOTALS	275.5	131.0		131.3		42.2	
WEST VIRGINIA							
Consolidated Gas Supply Corp.-Hastings Wetzel Co.	150.0	98.0	3	95.0	18.0	33.0	
WYOMING							
Amoco Production Co.-Bairoil, Sweetwater Co. 7-26n-90w	5.0	3.3	3	3.9			
Beaver Creek, Fremont Co., 10-33n-96w	65.0	37.4	2	12.8		11.4	
Elk Basin, Park Co. 29-58n-99w	17.0	11.7	1	21.0		14.4	

APPENDIX B (Continued)

Company, Plant, County, Location	Gas Capacity	Gas Through-put	Process Method	Prop.	Isobut.	Normal Or Unsplit Butane	LP-gas Mix
WYOMING (Continued)							
Arco Oil & Gas Co.-Gilette, Campbell Co., 18-50n-73w	31.0	10.9	3	28.0		14.5	
Patrick Draw, Sweetwater Co.	30.0	18.3	2	13.8	2.1	7.4	
Colorado Interstate Gas Co.-Rawlins, Carbon Co., sw4-sw4-25-21n-86w	220.0	202.0	2	54.2		34.7	
Continental Oil Co. -Sussex, Johnson Co. 243-41n-78w	15.0	1.5	3	1.7			
CRA Inc.-Joe Creek, Campbell Co., 34-57n-75w	2.0	0.[illegible]	3				0.8
Lazy B. Campbell Co., 1-+49n-74w	3.0	2.5	3	4.5		4.2	
Gas Producing Enterprises Inc.-Patrick Draw, Sweetwater Co., 12-19n-99w	12.0	7.5	2	6.2		3.5	
Ginther Gas Processing Plants-Rozet Campbell Co., 18-50n-69w	4.0	0.2	2	0.9			1.5
Springen Ranch Campbell Co., 28-51n-71w	8.0	0.7	2	3.0			5.0
Husky Oil Co.-Ralston, Park Co., 4-56n-100-w	7.0	3.0	4	0.2		0.8	
Kansas-Nebraska Natural Gas Co.,Inc.-Casper, Natrona Co. 10-33n-78w	80.0	35.9	2	16.3		9.3	
Flat Top, Converse Co., 20-33n-68w	6.0	3.0	2	3.7			
McCullouch Gas Processing Corp.-Hilight, Campbell Co. 26-45n-71w	60.0	12.0	3	16.0			
Oedekoven, Campbell Co.,30-55n-73w	12.5	3.0	3	9.0			
Wall Draw, Converse Co., 26-35n-69w	3.0	2.0	3	3.0			
Natomas North America Inc.-Recluse, Campbell Co., 15-56n-74w	9.0	3.9	3	8.4			
Northwest Pipeline Corp.-Opal, Lincoln Co., 27-21n-114w	250.0	155.1	2	86.3		31.0	
Ralston Processing Associates Inc. Oregon Basin, Park Co., nw4-B-52n-100w	6.0	4.1	2	2.1		2.5	4.3
Union Oil Co. of California-Worland Washakie Co., 17-48n-92w	50.0	20.2	1	17.7		6.4	
TOTALS	897.5	538.3		302.7	2.1	135.9	15.8
GRAND TOTALS	34,570.1	18,512.7					

APPENDIX C

FOREIGN FLAG, SELF-PROPELLED, LIQUID GAS CARRIERS

APPENDIX C

FOREIGN FLAG, SELF-PROPELLED, LIQUID GAS CARRIERS

LEGEND

Name of vessel and all previous names (the most recent shown first).

Flag:

Ag.	Algerian
Am.	American
Ar.	Argentine
Au.	Australian
Bd.	Bermudan
Be.	Belgian
Br.	British
Bz.	Brazilian
Ca.	Canadian
Ch.	Chilean
Cu.	Cuban
Cy.	Cyprian
Da.	Danish
Du.	Dubai
Fi.	Finish
Fj.	Fijian
Fr.	French
Ge.	German
Gr.	Greek
Ia.	Indonesian
Ir.	Iranian
It.	Italian
Ja.	Japanese
Ko.	Korean (South)
Ku.	Kuwaiti
Li.	Liberian
Ma.	Maltese
Mg.	Malagasy
My.	Malaysian
Me.	Mexican
Mr.	Morocco
Ne.	Netherlands
N.A.	Netherlands Antilles
Nu.	Norwegian
Pa.	Panamanian
Ph.	Philippine
Po.	Portuguese
Pp.	Papua New Guinea
Sa.	Saudi Arabia
Se.	Senegalese
Sg.	Singapore
Sp.	Spanish
Sw.	Swedish
Th.	Thailand
To.	Tongan
Tr.	Trinidad
Tu.	Turkish
USSR	Union of Soviet Socialist Republics
Ve.	Venezuelan

Number and Type of Tanks:

ch	cylindrical horizontal
co	conical
cv	cylindrical vertical
dc	double-cylinder
pr	Prismatic
r	rectangular
sp	spherical
sq	square
I	Insulated
P	Pressurized
R	Refrigerated
S	Semi-Refrigerated*

Cargo Containment Systems:

(BE)	Bridgestone
(CH1)	Conch 1
(CH2)	Conch 2
(EO)	Esso
(GT)	Gaz Transport
(KM)	Kvaerner-Moss
(LGA)	Liquid Gas Analgen
(ST)	Sener Techia
(TZM)	Technigaz (Conch Ocean) - Membrane
(TZS)	Technigaz - Spherical
(WGF)	Worms/Gaz de France

*Category "S" vessels have facilities for a degree of refrigeration and pressure.

APPENDIX C

FOREIGN FLAG, SELF-PROPELLED, LIQUID GAS CARRIERS*

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
AL GHASSANI	MR	Marphocean	4,100	S	1977	
ALPHAGAS	Ge	Sloman Neptun	2,768	2 dc S 1 ch S steel 7.5 -34°C	1970	Ice Strengthened
ANNA SCHULTE	Ge.	Bernhard Schulte	2,420	2 S (LGA) aluminum -162°C	1973	Methane/Ethylene
ANTILLA BAY	N.A.	Schpy, Mtj. "Volharding" N.V. Holland Bulk Transport	53,424	4 r R steel 2.8 -48°C	1973	Ice class 3 7/22/79
ANTILLA CAPE	N.A.	Schpy, Mtj "Volharding" N.V. Holland Bulk Transport	29,540	4 r R steel 1.3 -51°C	1968	Ice class C 7/22/79
BARFONN	No	Skibs A/S Dalfonn	8,500	6 S 5 -48°C	1969	Ice class C 6/15/79
BAVARIA MULTINA	Ge.	Liquid Gastanker Multina Schiff	6,000	S	1977	Ethylene/L.P.G.
BEN FRANKLIN	Fr.	Gazocean Gazocean Armement	120,000	6 pr I (TZM) (membrane) -165°C	1975	
BERGA	Ag. Ex-Pascal	Algerienne, Cie Nationale	6,310	6 ch S	1967	Can transfer to semi-refrigerated storage
BETAGAS	Ge.	Sloman Neptun	2,768	2 dc S 1 ch S steel 7.5 -34°C	1971	Ice strengthened 10/6/80
HULL 577	USSR	Latvian Shipping Company	12,000	2 ch R 1 co R -48°C	1976	
BOW ELM	No.	A/S Rederiet Odfjell	7,418	6 ch S 3.4 -104°C	1971	Ethylene
CAPO OVEST	Fr.	Compagnie des Messageries Maritimes	14,800	3 R -45°C	1967	
CARIBGAS X	Pa.	Caribgas Inc.	1,120	2 ch S steel 6.3 -11°C	1965	L.P.G./Solvents

*Vessels which have contacted, applied for, and/or received United States Coast Guard Letter of Compliance.

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
CATY MULTINA	Li	Cabo Tres Montes Inc.	3,735	6 ch S 7.2 -48°C	1969	Ice class C
CAVENDISH	Br.	Ocean Gas Transport Ltd. Houlder Bros.	40,213	3 pr R steel -45°C	1971	11/17/80
CENTUM HULL 204	No.	Moss-Rosenberg	52,000	4 pr R -48°C	1978	
CHANTIERS DE L ATLANTIQUE G-26		Algerienne, Cie Nationale	125,000	pr I (GT) (membrane)	1979	Methane/L.P.G.
CHANTIERS DE L ATLANTIQUE HULL I.26		Arab Maritime Pet. Trans. Co.	75,000	R	1978	
CHANTIERS DE L ATLANTIQUE J.26		Arab Maritime Pet. Trans. Co.	75,000	R	1979	
CHANTIERS DE L ATLANTIQUE L-26		Algerienne Cie Nationale	125,000	pr I	1980	Methane/L.P.G.
CLERK MAXWELL	Br.	Ocean Gas Transport Ltd. Houlder Bros.	11,753	3 r R steel ----- -50°C	1966	7/3/79
CNIM HULL 1416	Sw.	Multinational Gas/Malmros R/S	53,400	4 pr R steel -48°C	1978	Can carry 2 types of cargo simultaneously
CNIM HULL 1417	Li.	Multinational Gas	53,400	4 pr R steel-48°C	1978	Can carry 2 types of cargo simultaneously
CORAL ISIS (GC)	N.A.	Koraal Scheepr. Mij. N.V.	5,500	3 dc S steel 7.5 -48°C	1976	1/11/79
CORAL MAEANDRA	N.A.	Koraal Scheepr. Mij. N.V.	4,576	4 r P steel* 1.3 -48°C	1968	Ice class 2 *Nos. 2 and 3 stainless steel and can carry phosp. acid and nitric acid
CORAL OBELIA	N.A.	Koraal Scheeper. Mij.	3,038	6 ch S 7.4 -5°C	1966	Finnish ice class 1C
CHEMTRANS CAPELLA KROGERWERFT HULL 1394	Ge.	German (Unknown)	3,000	S	1978	
DANIAN GAS	Li.	Reliance Gas Transport Corp.	26,504	3 pr R -46°C	1969	5/7/79
DESCARTES	Fr.	Gazocean-Armement	50,000	6 r I (TZM) (membrane) -160°C	1971	3/25/79
DEVONSHIRE	Br.	Bibby Line Ltd. Bibby Bros. & Co.	52,650	4 r R steel -48°C	1974	

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
DRYBURGH	Br.	Anchor Gas Tankers Ltd. George Gibson & Co. Ltd.	1,530	2 ch S steel 8.8 -5°C	1952	
DUBULTY HULL 579	USSR	Latvian Shipping Co.	12,000	2 ch R 1 co R -48°C	1976	6/25/79
DZINTARI HULL 578	USSR	Latvian Shipping Company	12,000	2 ch R 1 co R -48°C	1976	12/29/80
ERIK RAUDE	No.	A/S Gasskib	6,170	2 ch S steel 4.1 -40°C	1967	Finnish ice class 1C
ELISABETH	Ge.	Bernhard	2,420	2 S (LGA) aluminum -162°C	1974	Methane/Ethylene
EMILIANO ZAPATA	Me.	Petroleos Mexicanos	3,344	2 r R steel -103.7°C	1970	Ethylene
ENRICO FERMI	It.	Carbocoke	7,500	3 dc S 4.5 -48°C	1977	12/2/79
EPSILONGAS	Ge.	Sloman Neptun	5,600	3 dc S steel 7.5 -48°C	1977	4/19/79
ESSO BREGA	It.	Prora Trasporti S.p.A. Esso Transport Co. Inc.	40,000	4 r I (EO) aluminum 1.3 -162°C	1969	Methane* for propane as alternative cargo (33,330 cu. met. on same draught)
ESSO PUERTO RICO	Pa.		12,788			L.P.G./Oil.
ESSO WESTERNPORT	Li.	Esso Tankers Inc. Exon International Co.	101,000	8 pr R	1977	
EUCLIDE EX-EUCLIDES	Li.	Antartic Gas Inc. The Counties Ship Management Co. Ltd.	4,000	4 sp S TZS 9% Ni Steel -160°C	1971	Methane/Ethylene 9/30/79
EVA THOLSTRUP	Da.	Kosan Tankers A/S	889	5 sp P steel 18	1958	Finnish ice class 1B
EVANGELISTA TORICELLI	It.	"Carbocoke" Societa di Navigazione S.p.A.	1,421	6 ch S steel 9.1 -10°C	1964	Finnish ice class Ic Can load propane at -42°C
FARADAY	Br.	Ocean Gas Transport Ltd. Houlder Bros.	31,215	3 pr R Steel -45°C	1971	10/6/80

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
FERNBROOK	No.	A/S Hanico	12,060	4 ch S steel 5 -48°C	1976	
MOSS HULL 184	No.	A/S Kim	12,000	4 ch S -49°C	1976	
FERNVALLEY	No.	Fearnley & Eger	22,240	3 r R steel -48°C	1969	Ice class C 12/18/79
FERNWAVE	No.	Fearnley & Eger	12,000	4 ch S 4.1 -48°C	1972	3/23/79
FERNWIND	No.	Fearnley & Eger	22,246	3 R -48°C	1968	1/13/80
FERNWOOD	No.	Fearnley & Eger	21,795	3 R -48°C	1969	8/10/80
FRITZ HABER	Ge.	Fritz Haber K.G. Bernard Schulte	5,174	3 dc S steel 7.5 -48°C	1971	Ice class E2 12/14/80
FROSTFONN	No.	I/S Frostfonn	4,163	6 ch S	1965	
GALPARA THYSSEN NORDSEE-WERKE HULL 460	Br.	P. & O. Steam Nav. Co.	53,000	4 R -50°C	1978	
GAMBADA	Br.	P.&.O. Steam Nav. Co.	29,791	4 pr R steel -51°C	1973	Can carry two grades of cargo simultaneously Ice class 3 7/17/80
GAMBHIRA	Br.	P.& O. Steam Nav. Co.	14,103	4 r R steel -48°C	1969	Ice class 3 7/4/80
GAMMAGAS	Ge.	Sloman Neptune	5,202	3 dc S	1972	Ice class E2
GANDARA	Br.	P. & O. Steam Nav. Co.	22,500	4 pr R 4 -50°C	1976	Ice class 3 Can carry two grades of cargo simultaneously 7/11/80
GARBETA	Br.	P. & O. Steam Nav. Co.	22,000	4 pr R -48°C	1975	Ice class 3 Can carry two grades of cargo simultaneously 3/1/80
GARINDA TNSW HULL 459	Br.	P. & O. Steam Nav. Co.	53,000	4 pr R Steel -50°C	1977	5/25/79
GARMULA	Br.	P. & O. Steam Nav. Co.	52,649	4 pr R -48°C	1972	Ice class C 1/6/79
GAS AL KUWAIT		Kuwait Oil Tanker Company	72,000	4 pr R	1978	
GAS LION	No.	Kristian Gerhard Jebsen	11,842	3 r R steel 1.3 -50°C	1968	Will carry and load/discharge simultaneously two grades of liquid gas
GAY LUSSAC	Pa.	Trans-oceangas Shipping S.A.	40,232	3 pr R -48°C	1969	
GAZANA	Br.	P. & O. Steam Nav. Co.	29,791	4 pr R	1972	Ice class 3 Can carry two grades of cargo simultaneously

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
GDYNIA HULL B550/3	No.	Leif Hoegh & Co. A/S	75,000	R -48°C	1978	
GDYNIA HULL B550/4	No.	Lief Hoegh & Co. A/S	75,000	R -48°C	1978	
GEROLAMO CARDANO	It.	"Carbocoke" Societa di Navigazione S.p.A.	4,100	6 ch S	1965	
GIMI	Li.	Cotaas-Larsen	125,000	6 sp I (KM) Aluminum -163°C	1976	Methane/L.P.G.
HAMPSHIRE	Br.	Bibby Line Ltd. Bibby Bros. & Co.	52,650	4 r R steel -48°C	1974	Ice class 3
HARDANGER	No.	Westfal-Larsen & Co. A/S	7,418	6 ch S 3.4 -104°C	1972	4/8/80
HAVFROST	No.	A/S International Gas Carriers	11,400	3 r R steel -51°C	1966	Ice class C
HAVIS	No.	A/S International Gas Carriers P. Meyer	15,285	3 r R -48°C	1970	Can carry vinyl chloride Ice class C
HEBE	No.	Helge R. Myhre	5,100	3 ch S 1 dc S 4 -48°C	1975	
HELEN	No.	Helge R. Myhre	2,200	4 ch R	1971	
HELIOS	No.	Fearnley & Eger	12,000	4 ch S 5 -49°C	1976	5/14/80
HERA	No.	Helge R. Meyhre	12,000	3 ch S 1 dcs steel 5 -104°C	1977	Ethylene/L.P.G. 11/2/79
HERIOT	Br.	Gibson Gas Tankers Ltd. George Gibson & Co., Ltd.	2,469	2 S	1972	Ice class 3 Ethylene/Methane
HEROS	No.	Helge R. Myhre	12,000	4 ch S -48°C	1977	Ethylene/L.P.G.
HESTIA (GC)	No.	Helge R. Myhre	2,450	4 ch S	1978	
HILLI	Li.	Gotaas-Larsen	125,000	6 sp I	1975	Methane/L.P.G.
HOEGH MULTINA	No.	Leif Hoegh & Co. A/S	52,000	4 pr R -48°C	1971	Ice class C 1/19/79
HOEGH SHIELD	No.	Leif Hoegh & Co. A/S	8,610	4 ch S	1969	
HOEGH SWALLOW	No.	Leif Hoegh & Co. A/S	75,500	4 pr R	1977	

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
HOEGH SWIFT	No.	Leif Hoegh & Co. A/S	75,500	4 pr R	1977	
HUMBOLDT	Br.	Ocean Gas Transport Ltd.	6,327	6 ch S steel 6.4 -45°C	1968	Ice strengthened
INGA THOLSTRUP	Da.	Kosan Tankers A/S	2,109	3 ch S steel 8.5 -10°C 1 cv P steel 17.6	1965	Finnish ice class 1B
INGE MAERSK	Da.	A. P. Moller	12,060	4 ch S steel 4 -48°C	1972	
ISABELLA	Li.	Multimare Shipping Co.	35,000	5 pr I	1975	
ISFONN	No.	Skibs A/S Dalfonn Sigval Bergesen	18,790	3 R	1967	
JAMES COOK	To.	Gaspac Shipping Co. Ltd. Bruce Reid & Sons Pty. Ltd.	1,580	4 ch S	1971	Ice class B
JOHANN KEPLER	Ge.	Partenreederei "Kap Roland" Bernard Schulte	2,150	1 sp S	1968	
JOULE	Br.	Ocean Gas Transport	11,200	3 r R steel -51°C	1965	8/31/80
KAWASAKI HULL 1225	Li.	World Wide (Shipping) Ltd.	79,600	R	1977	
LA CIOTAT HULL 315	Ku	Kuwait Govt.	70,000	R	1978	
LA CIOTAT HULL 316	Ku	Kuwait Govt.	70,000	R	1979	
LA CIOTAT HULL 317	Ku	Kuwait Govt.	70,000	R	1979	
LANRICK	Br.	Anchor Gas Tankers Ltd. George Gibson & Co. Ltd.	1,589	2 ch S steel 10.3 -5°C	1957	
LAVOISIER	Ar.		5,230			
LEIV ERIKSSON	No.	Einar Bakkevig	2,700	4 ch S 5% nickel steel 6 -104°C	1972	Ethylene
LIELUPE MEYER HULL 581	USSR	Latvian Shipping Co.	12,000	2 ch R 1 co R -48°C	1977	

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
LINCOLN ELLSWORTH	No.	Einar Bakkevig	830	1 ch S Nickel steel 6.3 -104°C	1966	Ice class C Ethylene
LINCOLNSHIRE	Br.	Bibby Line Ltd.	31,290	3 r R steel 1.3 -50°C	1972	Ice class 3/14/80
LNG POLLANGER	Br.	LNG Carriers Ltd. P. & O. Bulk Shipping Div.	87,600	5 sp I (KM) 9% Nickel steel -163°C	1974	Methane/L.P.G.
LORD KELVIN	Br.	Ocean Gas Transport Ltd.	31,000	R -48°C	1977	
LPG KATRISA	Li.	Tanker Trading Inc.	18,422	4 r R steel 1.3 -51°C	1968	Finnish Ice class 1C Transverse bow propeller 3/7/80
LUCIAN	No.	Hilmar Reksten	29,000	4 sp R (KM) aluminum 3 -163°C	1974	Ice strengthened Methane/Ethylene/L.P.G. 2/6/79
LUIGI CASALE	It.	"Carbocoke" Societa di Navigazione S.p.A.	14,268	3 r R steel -48°C	1967	3/7/79
LUIGI GALVANI	It.	"Carbocoke" Societa di Navigazione S.p.A.	4,600	6 S -48°C	1971	
LUIGI LAGRANGE	It.	"Carbocoke" Societa di Navigazione S.p.A.	31,000	4 pr R -48°C	1976	2/21/80
MAGELLAN	To.	Gaspac Shipping Co. Ltd. Bruce Reid & Sons Pty Ltd.	900	4 P 17.6 -25°C	1967	Ice strengthened One grade only
MALMROS MULTINA	Sw.	Malmros R/A	53,400	4 pr R steel 1.3 -48°C	1974	
MARCO POLO	No.	A. I. Langerfeldts Rederi	3,000	6 ch S steel 5.5 -40°C	1967	Ice class C
MARIAN P. BILLUPS	Li.	Tropigas Tankers Inc.	2,033	19 v P 17.6	1956	1/13/79
MARIANNE THOLSTRUP	Da.	Kosan Tankers A/S	2,500	3 ch S 8.5 -33°C	1968	Ice class 2
MAYORI	USSR	Latvian Shipping Co.	12,000	2 ch R 1 co R -48°C	1977	10/22/79

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
MELROSE	Br.	Gibson Gas Tankers Ltd. George Gibson & Co. Ltd.	2,725	2 R (LGA) aluminum -162°V	1971	Ice class III Methane, Can carry Ethylene
MERCURY GAS	Ja.	Far East Shipping K.K. Koji Fukushima	1,600	2 ch P steel 18 -5°C	1973	
MITSUBISHI HULL 959	Li	Edgington Prince Shipping Co.	76,900	R	1977	
MONGE HULL 312	Fr.	Gazocean Armement	70,000	4 r R -45°C	1977	10/26/79
MONOMER VENTURE	Pa.	Mundogas Inc. (Storage)	5,748	4 R -45°C	1945	
MONTANA HULL 1402		C.N.I.M.	35,500	5 pr I	1975	Methane/Ethylene/L.P.G. Vessel completed but delivery postponed
MOSTEFA BEN BOULAID HULL 302	Ag.	Algerienne, Cie Nationale	125,000	6 I (TZM) (membrane) -160°C	1976	12/20/80
MUNDOGAS ATLANTIC	No.	A/S Gasskib Oivind Lorentzen	8,565	4 ch S steel 4.6 -48°C	1969	12/14/79
MUNDOGAS BRASILIA	No.		7,739	5 sp S Steel -5°C		L.P.G./Oil. 12/20/80
MUNDOGAS RIO	No.	A/S Gasskib	19,462	4 r R steel 1.3 -42°C	1967	
NESTAGAS	Fi.	Neste Oy (Finnish Govt.)	4,100	6 ch R steel 5.5 -104°C	1974	
NESTEFOX	Fi.	Neste Oy (Finnish Govt.)	6,000	6 ch S Steel 5 -48°C	1977	4/11/79
NICOLE	Sw.	A/B Transmarin	3,000	6 ch S steel 9 -48°C	1967	Can receive cargo at -48°C
NIELS HENRIK ABEL	No.	Einar Bakkevig	2,500	1 ch S 2 do S steel 7.2 -48°C	1973	Ice class B
NORDFONN	No.		4,050			L.P.G./Chemical
NORMAN LADY	Br.	Buries Marks Ltd./Leif Hoegh	87,600	5 sp I (KM) 9% nickel steel 1.3 -163°C	1973	Methane/L.P.G.

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
NYHAMMER	No.	Chr. Haaland	66,341	4 pr R steel -48°C	1975	
NYHAVN	No.	Chr. Haaland	19,450	8 pr R steel 1.3 -47°C	1968	Ice class C
OLAV TRYGVASON	No.	Einar Bakkevig	4,100	4 ch S 1 dc S 5% nickel steel 5-105°C	1975	Ethylene/L.P.G.
PASCAL	Fr.	Cie Generale Maritime	15,022	6 ch S steel 5 -48°C	1976	3/3/79
PENTLAND BRAE	Br.	Liquid Gas Equipment Ltd. George Gibson & Co.Ltd.	3,850	2 dc S Steel 7 -48°C	1976	9/19/79
PENTLAND GLEN	Br.	Liquid Gas Equipment Ltd. George Gibson & Co. Ltd.	2,503	2 ch S steel 8.5 -34°C	1972	Ice class 2
PERMIAN GAS	Li.	Global Gas Transport Inc.	9,000	6 ch S steel 6.3	1968	
PETROGAS I	Li.	Norfolk Multina Shipping Co.	25,102	5+ pr R steel -51°C	1964	Also 4 pressure tanks on deck (4 pumps, 14 t.p.h. each), 1,216 cu. m.
PROVIDENCE MULTINA	Fr.	S.O.F.R. A.N.G.A.Z. Multinational Gas & Pet. Ser.	53,400	4 pr R 1.3 -48°C		
PYTHAGORE	Pa.	Antartic Gas Inc.	14,258	3 pr R steel -48°C	1967	2/21/79
RAZI HULL 313	Pa.	Iranocean	70,000	4 pr R Steel -46°C	1977	1/2/81
ROALD AMUNDSEN	No.	Einar Bakkevig	4,100	4 ch S 5.5 -104°C	1971	Ethylene Ice class C 2/9/80
SIGURD JORSALFAR	No.	Einar Bakkevig	2,500	1 ch S 2 dc S steel 7.2 -48°C	1973	
SINE MAERSK	Da.	A. P. Moller	12,000	4 ch S 4 -48°C	1976	2/1/79
SOFIE MAERSK	Da.	A. P. Moller	12,000	4 ch S 4 -48°C	1977	10/18/79
SOPHIE SCHULTE	Ge.	Bernhard Schulte	2,420	2 S (LGA) aluminum -162°C	1973	Methane/Ethylene

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
SUNNY BABY	No.	Sameiet Sungas Olaf Pedersens Rederi A/S	1,616	2 ch S steel 8 -48°C	1965	Can load vinyl chloride Ice class C
SUNNY BOY	No.	Sameiet Sungas Olaf Pedersens Rederi A/S	1,616	2 ch S steel 8 -48°C	1967	Can load vinyl chloride Ice class C
SUNNY DUKE HULL 131	No.	Olaf Pedersen	3,850	2 dc S steel 6.5 -48°C	1977	6/13/79
SUNNY FELLOW	No.	Olaf Pedersen	1,526	4 ch S steel 7.5 -10°C	1968	Ice class C
SUNNY GIRL	No.	Sameiet Sungas	900	4 ch S 17.6 -10°C	1967	Finnish ice class 1B
SUNNY QUEEN	No.	Olaf Pedersen	3,850	2 dc S steel 6.5 -48°C	1976	9/3/80
SYDFONN	No.	Sigval Bergesen	22,240	3 pr R steel -48°C	1978	8/2/80
TATSUNO MARU	Ja.	Nippon Yusen	50,670	4 r R steel -45°C	1967	
THYSSEN HULL 461	Br.	P. & O. Steam Nav. Co.	53,000	4 R -50°C	1978	
THYSSEN HULL 462	Br.	P. & O. Steam Nav. Co.	53,000	4 R -50°C	1979	
TINE THOLSTRUP	Da.	Kosan Tankers A/S	1,622	2 ch S steel 8.5 -42°C	1968	Ice class 2
TOKUHO MARU	Ja.	Iino Kaiun K.K.	61,200	4 R steel	1974	
TOMAS RUIZ DE VELASCO	Da.	B.P., Denmark	920	P	1976	
TORDENSKIOLD	No.	A. I. Langfeldt & Co.	4,100	4 ch S 6 -48°C	1971	
TROIKA	No.	Chr. Haaland	3,700	6 ch S	1968	Ice class C
TROPIGAS FAR EAST	Pa.	Viking Asia Inc. Far East Shipping Co. Ltd.	5,000	5 sp S 12 -45°C	1976	12/12/80
TROPIGAS ISLANDER	Li.	Tropigas Tankers Inc.	227	2 ch P 17.6	1967	10/23/80

APPENDIX C (Continued)

Name	Flag	Owners Managers	Cargo Capacity Cubic Metres	Tanks (No.& Type) Max. Pressure (Kgf/cm^2) Min. Temp.	Year Built	Remarks
VASCO DE GAMA	No.	A. I. Langfeldts Rederi	6,100	4 ch S steel 5 -104°C	1976	Ice strengthened Ethylene/L.P.G.
VENATOR	No.	Peder Smedvig	29,000	4 sp R (KM) aluminum -163°C	1973	Methane-can carry Ethylene or L.P.G.
VENUS GAS	Ja.	Far East Shipping K.K. Koji Fukushima	2,500	2 ch P steel 18 -5°C	1973	
VESTRI GAS	No.	Peder Smedvig	12,058	4 ch S nickel steel 4 -104°C	1972	Ethylene 10/10/79
WARTSILA HULL 1229	No.	Fearnley & Eger/Nissho Iwai Co. (Jointly)	75,000	4 sq R 1 -48°C	1978	
WARTSILA HULL 1230	No.	Fearnley & Eger	75,000	4 sq R 1 -48°C	1978	
WARTSILA HULL 1231	No.	Fearnley & Eger/Nissho Iwai Co. (Jointly)	75,000	4 sq R 1 -48°C	1979	
WARTSILA HULL 1232	No.	Sig Bergesen d.y. & Co.	75,000	4 sq R 1 -48°C	1979	

APPENDIX D

LISTING OF LIQUEFIED FLAMMABLE GAS BARGES

APPENDIX D

LISTING OF LIQUEFIED FLAMMABLE GAS BARGES

LEGEND

G.T. - GROSS TONNAGE

Refers to the tank barge's gross tonnage

DBL - S/B - DOUBLE SIDE, DOUBLE BOTTOM

Side Codes
* - Single side/skin
0 - Non water-tight (independent tanks)
W - Double sides

Bottom Codes
* - Single bottom/skin
P - Partial double bottom
F - Full double bottom

BUILT

Refers to the year in which the tank barge was initially constructed regardless of subsequent rebuilding, if any.

MAT

Refers to the tank barge's hull construction whereby S = Steel

RTE

Route - Refers to the route for which the tank barge is certified in accordance with the following codes:

Ocean	OO
Coastwise	CC
Great Lakes	GG
Lakes, Bays, Sounds	LL
Rivers	RR
LBS & Coastwise (lim.)	LC
LBS & Great Lakes (lim.)	LG
Rivers & Great Lakes	RG
Coast & Great Lakes	CG

SBCH

Refers to the subchapter indicated on the "Certificate of Inspection" whereby:

1. Subchapter D
2. Subchapter O/D
3. Subchapter O/I

HULL TYPE

Refers to the tank barge hull type indicated on "Cartificate of Inspection" whereby 1 = Type I hull; 2 = Type II hull; and 3 = Type III hull. Where the Hull Type is not indicated on the "Certificate of Inspection," it is assumed that the tank barge is a Type III.

CAPACITY

The total capacity is indicated by amount having B for number of barrels, G for number of gallons, and T for number of short tons (i.e., 200 lbs/ton).

APPENDIX D

LISTING OF LIQUEFIED FLAMMABLE GAS BARGES

VESSEL NAME VESSEL NUMBER	GT DBL-S/B	BUILT MAT	LENGTH BREADTH	OPERATOR OWNER	RTE. SBCH	HULL TYPE	GRD CARGO CAPACITY
Alamo 3000 DN 293548	2866 **	1963 S	380.0 53.2	Alamo Chem. Trans. Co. Alamo Chem. Trans. Co.	LL 2	3	LF 31420 B
Alamo 3001 DN 293549	2816 **	1963 S	370.0 53.2	Alamo Chem. Trans. Co. Alamo Chem. Trans. Co.	LL 2	3	LF 31403 B
Amoco A-51 DN 294858	2267 WF	1964 S	290.0 50.0	Southern Towing Co. Amoco Oil Company	LL 3	3	LF 804663 G
Amoco A-52 DN 294859	2254 WF	1964 S	290.0 50.0	Southern Towing Co. Amoco Oil Company	LL 3	3	LF 885110 G
BEG 3002 DN 295147	2435 0*	1964 S	320.0 53.2	BF Goodrich Chem. Co. B F Goodrich Chem. Co.	LL 2	3	LF 26807 B
BFG 3003 DN 500748	2866 N*	1965 S	3800.0 53.2	B. F. Goodrich Chem. B. F. Goodrich Chem.	LL 2	2	LF 31422 B
BFG 3004 DN 517396	2812 **	1968 S	355.0 53.2	B. F. Goodrich Chem. Alamo Chem. Trans. Co.	LL 2	2	LF 31425 B
BWG 102 DN 285777	1394 W*	1961 S	195.0 50.0	B F Goodrich Chem. Co. G W Gladders Towing Co.	RR 2	3	LF 13264 B
CBC 261 DN 510425	2502 W*	1967 S	295.0 50.0	Canal Barge Co., Inc. Canal Barge Co., Inc.	LL 2	2	LF 23322 B
CBC 263 DN 510738	2502 W*	1967 S	295.0 50.0	Canal Barge Co., Inc. Canal Barge Co., Inc.	LL 2	2	LF 23322 B
CC-107 CG 007195	648 0*	1955 S	195.0 35.0	Union Carbide Corp. Union Carbide Corp.	LL 2	3	LF 6416 B
CC-108 CG 006007	648 0*	1955 S	195.0 35.0	Union Carbide Corp. Union Carbide Corp.	LL 2	3	LF 6416 B
CC-114 CG 006304	1572 0*	1962 S	195.0 52.5	Union Carbide Corp. Union Carbide Corp.	LL 2	3	LF 18938 B
CC-116 CG 006403	1370 **	1967 S	214.0 50.0	Union Carbide Corp. Union Carbide Corp.	LL 2	2	LF 1490 T
CC-117 CG 006452	1370 W*	1967 S	214.0 50.0	Union Carbide Corp. Union Carbide Corp.	LL 2	2	LF 1490 T
CC-118 DN 509367	1716 **	1967 S	214.0 50.0	Union Carbide Corp. Union Carbide Corp.	00 2	2	LF 745 T
CC 134 CG 019653	2008 W*	1975 S	184.0	Union Carbide Corp. Union Carbide Corp.	LL 2	2	LF 2000 T
CC-139 CG 019703	2008 W*	1975 S	184.0	Union Carbide Corp. Union Carbide Corp.	LL 2	2	LF 2000 B
CC-500 DN 501179	1851 WF	1965 S	250.0 52.5	Union Carbide Corp. Union Carbide Corp.	00 2	2	LF 16700 B
CE-765 CG 007377	22 **	1951 S	45.0 18.0	US Army Corps Eng. US Army Corps Eng.	LL 1	3	LF 1400 S
CE 851 CG 007328	18 **	1967 S	35.0 14.0	US Army Corps Eng. US Army Corps Eng.	RR 1	3	LF 333 B
Chemical 701 CG 003953	1600 **	1965 S	225.0 52.0	E I DuPont Cenemours Connecticut General Life	LL 2	3	LF 17218 3
Chemical 702 CG 003193	2188 WF	1966 S	250.0 53.0	E I DuPont Denemours & Co. Connecticut General Life	LL 2	1	LF 832196 G
Chemical 703 CG 003152	2188 WF	1967 S	250.0 53.0	E I DuPont Cenemours Connecticut General Life	LL 2	1	LF 832196 G
Chemical 704 CG 003202	3180 WF	1967 S	346.0 53.0	E I DuPont Genemours Connecticut General Life	LL 2	1	LF 30730 B

APPENDIX D (Continued)

VESSEL NAME VESSEL NUMBER	GT DBL-S/B	BUILT MAT	LENGTH BREADTH	OPERATOR OWNER	RTE. SBCH	HULL TYPE	GRD CARGO CAPACITY
Cherokee DN 535335	823 **	1971 S	175.0 42.0	Warren Petroleum Corp. Inc. Warren Petroleum Corp. Inc.	LC 2	2	LF 1862 T
Choctaw DN 535943	823 **	1971 S	175.0 42.0	Warren Petroleum Corp. Inc. Warren Petroleum Corp. Inc.	LC 2	2	LF 1862 T
Chotin 2090X DN 532073	1591 WF	1969 S	185.0 54.0	Chotin Transportation Inc. Midland Enterprises Inc.	LL 2	2	LF 2200 T
City of Lake Charles DN 262194	862 0*	1951 S	195.0 44.0	Warren Petroleum Corp. Inc. Warren Petroleum Corp. Inc.	LC 2	3	LF 8571 B
City of Pensacola DN 270463	1920 0*	1955 S	247.0 48.0	Warren Petroleum Corp. Inc. Warren Petroleum Corp. Inc.	00 2	2	LF 26020T
City of Tampa DN 26674	1913 W*	1953 S	245.0 48.0	Warren Petroleum Corp. Inc. Warren Petroleum Co.	00 2	2	LF 17381 B
EIDC-51 CG 009373	1500 **	1964 S	255.0 50.0	E I DuPont Cenemours & Co. E I DuPont Cenemours & Co.	LL 1	2	LF 17331 B
EIDC 52 CG 012070	1500 W*	1964 S	255.0 50.0	Houston Barge Line Inc. Connecticut General Life	LL 3	2	LF 1701 T
Gasco Huki Kai I DN 276238	1256 0*	1958 S	229.8 44.1	Honolulu Gas Company First Hawaiian Bank	00 2	2	LF 818 T
Herpro I DN 507418	2798 W*	1967 S	295.0	Allied Chemical Corp. International Barge Inc.	LL 3	2	LF 10980 B
Herpro II DN 507489	2773 W*	1967 S	282.0	Allied Chemical Corp. International Barge Inc.	LL 3	2	LF 10980 B
Hollywood Chem 102 DN 285645	960 *	1961 S	195.0	Hollywood Terminals Inc. Hollywood LPG Carriers	LL 2	3	LF 10219 B
Hollywood Chem 103 DN 599031	1193	1978 S	228.0	Hollywood Marine Inc Hollywood LPG No 2	LL 2	2	
Hollywood Chem 151 DN 283631	1536 0*	1960 S	273.0 41.0	Union Texas Petroleum Co. Hollywood LPG No 2 Ltd	LL 2	3	LF 15515 B
Hollywood Chem 152 DN 283632	1536 0*	1960 S	273.0 41.0	Union Texas Petroleum Co. Hollywood LPG No 2 LTD	LL 2	3	LF 15734 B
Hollywood 6 78 DN 276047	1254 0*	1958 S	220.0	Petro Tex Chem Corp Hollywood LPG Ltd	LL 1	2	LF 11204 B
James A. Lyles DN 298556	2482 W*	1965 S	293.0	Southern Towing Co. Mutual Life Ins Company	LL 2	3	LF 20000 B
Jo Anne DN 541373	9841 WF	1972 S	500.1	Union Carbide Corporation F B Smith & Ben Musharet	00 2	1	LF 95025 B
N.M.S. No 2501 DN 29444	2482 WF	1964 S	293.0 50.0	Natl Marine Service Inc Natl Marine Service Inc	LL 2	2	LF 21720 B
N.M.S. No 2502 DN 294678	2431 WF	1964 S	280.0 50.0	Natl Marine Service Inc. Natl Marine Service Inc.	LL 2	2	LF 21720 B
N.M.S. No 2503 DN 294394	2700 WF	1964 S	281.0 50.0	Natl Marine Service Inc. Natl Marine Service Inc.	LL 3	3	LF 21720 B
P 394-A DN 276376	1037 0*	1958 S	210.0 44.1	Southern Terminal & Trans Southern Terminal & Trans	LL 2	3	LF 9000 B
P-394-E DN 276377	1037 0*	1948 S	210.0 44.1.	Southern Terminal & Trans McKenzie Service Company	LL 2	3	LF 9000 B
PANAMA CITY DN 26253	940 0*	1951 S	210.0 44.0	Warren Petroleum Corp Inc Warren Petroleum Corp Inc	LL 2	3	LF 9306 B
Propane Butae Barge C.E.No 762 CG 016303	22 **	1949 S	45.0 18.0	US Army Corps Eng US Army Corps Eng	LL 2	3	LF 3400 G
Puerto La Cruz DN 275344	2171 **	1957 S	295.0 48.0	Warren Petroleum Corp Inc Warren Petroleum Corp Inc	00 2	1	LF 21260 B

APPENDIX D (Continued)

VESSEL NAME	GT	BUILT	LENGTH	OPERATOR	RTE.	HULL	GRD CARGO
VESSEL NUMBER	DBL-S/B	MAT	BREADTH	OWNER	SBCU	TYPE	CAPACITY
SEMINOLE	1068	1972	175.0	Warren Petroleum Corp Inc	LL	2	LF
DN 536946	0*	S	42.0	Warren Petroleum Corp Inc	2		1862 T
Southern Big-N	932	1960	195.0	Hercules Inc	LL	3	LF
DN 283989	OF	S	44.0	Hercules Inc	2		10004 B
TMST 100	1441	19[illegible]	203.0	Houston Barge Line Inc	LL	3	LF
DN 509855	WF	[illegible]	52.0	Houston Barge Line Inc	2		1200 T
USL-122	1545	1970	180.0	Union Carbide Corporation	LL	2	LF
CG 004044	WF	S	52.5	F B Smith & Ben Maushardt	2		1566 T
USL-123	1545	1970	180.0	Union Carbide Corporation	LL	2	LF
CG 004093	WF	S	52.5	F B Smith & Ben Maushardt	2		1566 T
USL 137	2008	1976	184.0	Union Carbide Corp	LL	2	LF
DN 571659	WF			Trust Company	2		2000 T
Yazoo 1	712	1957	175.0	Texas Olefins Company	LL	3	LF
DN 276236		S		Texas Olefines Company	2		6439 B

APPENDIX E

UNITED STATES UNDERGROUND STORAGE SITES

APPENDIX E

UNITED STATES UNDERGROUND STORAGE SITES

(Capacity = Barrels x 1000)

Company County-Type	Propane	E-P Mix	nButane	iButane	Others	Remarks
ALABAMA						
Amoco Oil Co.						
Marengo-Mined	179	—	—	—	—	
ARIZONA						
California Liquid Gas Corp.						
Maricopa-Salt	—	—	—	—	1,800	Butane-Propane (Expanding to 2,000 m bbls)
Williams Energy Co.						
Apache-Salt Dome	—	—	—	—	1,583	11 wells suitable for propane or any B-P mix
Totals					3,383	
GEORGIA						
Dixie Pipeline Co.						
Lamar-Granite	220	—	—	—	—	
ILLINOIS						
Amoco Oil Co						
Madison-Mined	150	—	—	—	—	Olefin is propylene
Continental Oil Co.						
Madison-Mined	77	—	—	—	—	
Hydrocarbon Transportation, Inc.						
Will-Mined	100	—	100	50	—	
Mid America Pipeline						
Fulton-Mined	410	—	—	—	—	
Phillips Petroleum Co.						
Kankakee-Mined	150	—	—	—	—	
Shell Oil Co.						
Madison-Mined	230	—	500	—	—	
Texas Eastern Products Pipeline Co.						
Will-Mined	—	—	100	200	—	
U. S. Industrial Chemicals Co.						
Douglas-Mined	681	—	152	—	—	
Warren Petroleum Co.						
White-Mined Shale	55	—	—	—	—	
Totals	1,853	—	852	250	—	
INDIANA						
Continental Oil Co.						
Huntington-Mined	77	—	—	—	—	
Lake-Mined	250	—	—	—	—	
Silgas, Inc.						
Jadeson-Mined	500	—	—	—	—	
Texas Eastern Products Pipeline Co.						
Gibson-Mined	155	—	—	—	—	
Jackson-Mined	—	—	421	207	—	
Totals	982	—	421	207	—	
IOWA						
Hydrocarbon Transportation Inc.						
Polk-Mined	150	—	50	—	—	
Mid America Pipeline						
Johnson-Mined	385	200	—	—	—	
Totals	535	200	50	—	—	
KANSAS						
Amoco Oil Co						
Reno-Salt	1,539	—	—	—	—	
Grant-Salt	151		84	—	—	

APPENDIX E (Continued)

Company County-Type	Propane	E-P Mix	nButane	iButane	Others	Remarks
KANSAS (Continued)						
Atlantic Richfield Co.						
Reno-Salt	1,160	—	—	—	—	
Cities Service Co.						
Reno-Salt Dome	3,772	—	1,689	540	574	
Consolidated Storage Inc.						
Reno-Salt Strata	—	—	—	—	6,225	E-P Mix, Propane & Butanes
Dalco Marketing & Storage, Inc.						
Reno-Salt	2,000	—	—	—	—	
El Paso Products Co.						
Reno-Salt	2,000	—	—	—	—	Operated by Dalco
Empire Underground Storage, Inc.						
Reno-Salt	2,000	—	—	—	—	
Getty Oil Co.						
McPherson-Salt Dome	3,100	—	—	—	—	
Home Petroleum Corp.						
McPherson-Salt Layer	9,000	—	1,500	—	—	
Mid America Pipeline						
McPherson-Salt	5,386	—	715	216	499	Other is demethanized mix
McPherson-Salt	—	—	—	—	192	Other is field grade butane
Rice-Salt	—	1,293	—	—	—	
Mid-West Underground Storage						
McPherson-Salt	5,000	—	—	—	—	Also butanes & gasoline
National Cooperative Refinery Ass'n.						
McPherson-Salt	2,645	—	232	149	—	
Phillips Petroleum Co.						
Kingman-Salt	940	—	—	—	—	
Sentry Underground Storage Co.						
Rice-Salt	2,000	—	—	—	—	
UPG, Inc.						
Ellsworth-Salt Strata	5,000	1,400	1,700	510	220	Other is mixed butane
Ellsworth-Salt Strata	—	—	—	—	—	
Totals	45,693	2,693	5,920	1,415	7,518	
KENTUCKY						
Cincinnati Gas & Electric						
Kenton-Mined	167	—	—	—	—	Peakshaving
Columbia Hydrocarbon						
Greenup-Mined	200	—	80	—	—	
B. F. Goodrich Co.						
Marshall-Mined	220	—	—	—	—	Olefin is propylene
Totals	587	—	80	—	—	
LOUISIANA						
Amoco Oil Co.						
Bienville-Salt	574	—	—	—	—	
Cities Service Co.						
Cameron-Salt Dome	1,797	—	3,173	1,106	—	
Cameron-Salt Dome	—	—	—	—	—	
Cameron-Salt Dome	—	—	—	—	—	
Continental Oil Co.						
Evangaline-Salt Dome	714	—	—	—	—	
Dow Chemical Co.						
Assumption-Salt	—	—	—	—	14,832	Other is LPG
Dow Chemical-Promix						
Assumption-Salt Dome	1,100	—	992	765	—	
St. Martin-Salt Dome	—	—	735	—	—	
Enterprise Products Co.						
Bienville-Salt Dome	1,900	—	867	—	—	
Exxon Co., U.S.A.						
Ascension-Salt	1,030	—	2,550	566	—	
Ascension-Salt	—	—	—	—	—	

Company County-Type	Propane	E-P Mix	nButane	iButane	Others	Remarks
LOUISIANA (Continued)						
Hercules Petroleum Co., Inc.						
Bienville-Salt	1,250	—	—	—	—	
Mobil Oil Corp.						
Assumption-Salt	750	—	—	—	—	
Shell Oil Co.						
Ascension-Salt Dome	675	—	1,040	—	—	
Texaco Inc.						
St. Martin-Salt Dome	760	—	—	—	—	
Ascension-Salt Dome	1,200	—	628	439	—	
UCAR Pipeline, Inc.						
Assumption-Salt	—	—	—	—	—	
Union Texas Petroleum						
St. Martin-Salt Dome	2,850	—	340	520	—	
Wanda Petroleum Co.						
St. Martin-Salt Dome	—	—	—	—	5,200	LP-gas
Warren Petroleum Co.						
Bienville-Salt Dome	883	—	—	—	—	
Plaquemine-Salt Dome	1,677	—	1,208	2,211	1,327	
Totals	17,160	—	11,533	5,604	20,032	
MARYLAND						
Baltimore Gas & Electric Co.						
Baltimore-Mined	143	—	—	—	—	
Washington Gas Light Co.						
Montgomery						
Total	143					
MICHIGAN						
Amoco Oil Co.						
St. Clair-Salt	292	—	—	—	—	
Cities Service Co.						
Kent-Salt Dome	1,006	—	—	—	—	
Cook Investments						
Kent-Salt Dome	1,000	—	—	—	—	
Consumers Power Co.						
St. Clair-Salt	—	—	—	—	739	Butane-propane mix (4 caverns)
Marathon Oil Co.						
Wayne-Salt	883	—	625	186	—	
Mobil Oil Corp.						
Wayne-Salt Dome	347	—	—	—	—	
Alto Underground Storage (Ohio Northwest Development, Inc.)						
Kent-Salt Dome	1,025	—	—	—	—	
Phillips Petroleum Co.						
Wayne-Salt	200	—	—	—	—	
Sun Petroleum Co.						
Wayne-Salt Dome	417	—	372	—	—	
Totals	5,170	—	1,007	186	739	
MINNESOTA						
Solar Gas, Inc. (Aminoil, USA)						
Polk-Mined	333	—	—	—	—	
MISSISSIPPI						
Delta Underground Storage Corp.						
Forrest-Salt	800	—	—	—	—	
Enterprise Products Co.						
Forest-Salt	3,400	—	450	—	—	
Hattiesburgh Storage Facility						
Forrest-Salt Dome	3,000	—	—	—	—	
Mobil Oil Corp.						
Forrest-Salt Dome	2,601	—	—	—	—	
Union Texas Petroleum						
Forrest-Salt Dome	1,980	—	—	—	—	
Warren Petroleum Co.						
Forrest-Salt Dome	4,640	—	—	—	—	
Totals	16,421	—	450	—	—	
MISSOURI						
Continental Oil Co.						
Lawrence-Mined	77	—	—	—	—	

APPENDIX E (Continued)

Company County-Type	Propane	E-P Mix	nButane	iButane	Others	Remarks
MISSOURI (Continued)						
LaClede Gas Co.						
St. Louis-Mined	785	—	—	—	—	
Williams Pipe Line Co.						
Jasper-Mined	195	—	—	—	—	
Total	1,057	—	—	—	—	
NEBRASKA						
Metropolitan Utilities District of Omaha						
Douglas-Mined Shale	145	—	—	—	—	
Douglas-Mined Limestone	240	—	—	—	—	
Mid America Pipeline						
Lancaster-Mined	370	—	—	—	—	
Total	755	—	—	—	—	
NEW JERSEY						
Exxon Co., U.S.A.						
Union-Mined	250	—	255	—	—	
NEW MEXICO						
Arrow Gas Co.						
Eddy-Salt	170	—	—	—	—	
Cities Service Co.						
Roosevelt-Salt Dome	83	—	—	—	—	
El Paso Natural Gas Co.						
Lea-Salt	172	—	63	—	—	
Getty Oil Co.						
Lea-Salt Dome	166	60	—	—	—	
Warren Petroleum Co.						
Lea-Salt Layer	250	—	—	—	—	
Totals	841	60	63	—	—	
NEW YORK						
Atlantic Richfield Co.						
Courtland-Salt	400	—	—	—	—	
Mobil Oil Corp.						
Steuben-Salt Dome	778	—	246	—	—	
Texas Eastern Products Pipeline Co.						
Schuyler-Salt	3,810	—	—	—	—	
Totals	4,988	—	246	—	—	
OHIO						
Ashland Oil Co.						
Canton-Salt	—	—	100	—	—	
Cincinnati Gas & Electric Co.						
Butler-Mined	167	—	—	—	—	Peakshaving
Hamilton-Mined	191	—	—	—	—	Peakshaving
Columbia Hydrocarbon Corp.						
Lake-Salt	600	—	—	—	—	
Dayton Power & Light						
Butler-Mined	191	—	—	—	—	
Lake Underground Storage						
Lake-Salt	3,500	—	—	—	—	
Standard Oil Co. (Ohio)						
Allen-Mined	325	—	60	—	—	
Lucas-Mined	115	—	—	—	—	
Texas Eastern Products Pipeline Co.						
Butler-Mined	1,102	—	531	212	—	
Totals	6,191	—	691	212	—	
OKLAHOMA						
Atlantic Richfield Co.						
Seminole-Mined Shale	112	—	—	—	—	
Continental Oil Co.						
Grant-Salt Strata	473	—	183	39	42	Other is field butane
Grant-Salt Strata	—	—	—	—	56	PBC mix
Kay-Mined	—	—	—	—	256	LP-gas/LRG mix
Koch Oil Co.						
Grant-Salt	116	175	40	—	—	
Mid America Pipeline						
Beaver-Salt	—	51	—	—	—	

APPENDIX E (Continued)

Company County-Type	Propane	E-P Mix	nButane	iButane	Others	Remarks
OKLAHOMA (Continued)						
Shell Oil Co.						
Beckham-Salt	18	—	—	—	—	Inactive
Sun Petroleum Co.						
Tulsa-Mined Shale	230	—	—	—	—	
Warren Petroleum Co.						
Beaver-Salt Layer	40	33	40	37	—	
Totals	989	259	263	76	354	
PENNSYLVANIA						
Sun Gas Co.						
Delaware-Mined Granite	—	—	240	240	—	
Sun Gas Terminals & Storage, Inc.						
Delaware-Mined Granite	1,170	—	—	—	—	
Texas Eastern Products Pipeline						
Westmoreland-Mined	—	—	185	100	—	
Totals	1,170	—	425	342	—	
SOUTH CAROLINA						
Carolina Pipeline Co.						
York-Mined Granite	400	—	—	—	—	
York-Mined Granite	1,200	—	—	—	—	
Total	1,600	—	—	—	—	
TEXAS						
Amoco Oil Co.						
Brazoria-Salt	—	2,430	—	—	—	
Hockley-Salt	178	—	462	—	—	
Ector-Salt	250	—	—	—	—	
Andrews-Salt	—	—	—	—	97	
ARCO Chemical Co.						
Chambers-Salt Dome	575	—	607	—	—	
Atlantic Richfield Co.						
Ector-Salt	160	—	220	—	—	
Butane Supplies, Inc. (Pargas)						
Wood-Salt Dome	1,315	—	—	—	—	
Cities Service Co.						
Gray-Salt Dome	406	—	100	—	—	
Coastal States Crude Gathering Co.						
Harris-Salt Dome	3,183	—	2,376	4,472	1,681	LP-gas mix
Cosden Oil & Chemical						
Howard-Salt	100	80	80	80	—	
Dalco Petroleum Co.						
Yoakum-Salt	400	—	—	—	—	
Diamond Shamrock Oil & Gas Co.						
Chambers-Salt	6,000	4,000	4,000	—	—	
Dorchester Gas Producing Co.						
Carson-Salt Dome	247	—	—	—	—	
Dow Chemical Co.						
Brazoria-Salt	—	7,000	—	—	—	
El Paso Natural Gas Company						
Reagan-Salt	640	493	337	—	—	
Ector-Salt	99	50	58	—	—	4 wells
El Paso Products						
Ector-Salt Dome	107	187	615	130	—	Ethylene
Exxon Co., U.S.A.						
Chambers-Salt	—	—	—	1,143	—	
Getty Oil Co.						
Howard-Salt Dome	36	43	26	36	—	
Carson-Salt Dome	414	43	115	141	—	
Mid America Pipe Line						
Gaines-Salt	77	511	73	98	—	
Mobil Oil Corp.						
Martin-Salt Dome	116	—	—	—	—	
Liberty-Salt Dome	1,780	—	1,240	340	—	
Liberty-Salt Dome	—	—	—	—	1,299	Leaching for future butane or propane
Monsanto Co.						
Scurry-Salt	525	—	—	—	—	Not in use; to be abandoned

APPENDIX E (Continued)

Company County-Type	Propane	E-P Mix	nButane	iButane	Others	Remarks
TEXAS (Continued)						
Phillips Petroleum Co.						
Brazoria-Salt	1,333	3,139	1,321	333	—	
Hutchinson-Salt	2,141	105	807	383	—	
Sid Richardson Carbon & Gaso. Co.						
Winkler-Salt	118	100	—	—	—	
Sante Fe Pipeline Co.						
Andrews-Salt	—	150	—	—	—	
Shell Oil Co.						
Ector-Salt Bed	310	—	380	—	—	
Yoakum-Salt Bed	350	—	—	—	—	
Sun Gas Co.						
Scurry-Mined Shale	114	—	17	—	—	
Tenneco Oil Co.						
Chambers-Salt Dome	3,560	—	1,770	1,780	—	
Texaco Inc.						
Hardin-Salt Dome	—	—	830	309	—	
Potter-Salt Dome	77	—	162	—	—	
Andrews-Salt Dome	—	—	—	—	—	
Scurry-Salt Dome	150	—	—	—	—	
Texas Brine Corp.						
Harris-Salt Dome	6,300	—	650	900	1,600	Other is LP-gas pipeline mix
Matagord-Salt Dome	—	1,800	—	—	—	Under development
Texas Eastman Co.						
Smith-Salt Dome	2,800	—	—	—	—	
Texas Eastern Products Pipeline Co.						
Chambers-Salt Dome	14,322	1,660	7,204	1,245	1,430	Other is propylene & propane mix
Three Bar Underground Storage Corp.						
Ector-Salt	250	—	—	—	—	
Union Oil Co. of California						
Andrews-Salt	350	—	—	—	—	
Jefferson-Salt	—	—	—	—	640	
Union Texas Petroleum						
Upton-Salt Bed	1,300	—	—	—	—	
UPG, Inc.						
Freestone-Salt Dome	667	—	—	—	—	
Chambers-SaltDome	—	680	—	—	—	
Wanda Petroleum Co.						
Harris-Salt Dome	—	—	—	—	4,300	LP-gas
Warren Petroleum Co.						
Jefferson-Salt Dome	—	—	1,200	—	550	Other is propane-propylene
Chambers-Salt Dome	6,407	8,920	4,720	2,501	—	
Chambers-Salt Dome	—	—	—	—	1,216	
Xral Storage & Terminating Co.						
Chambers-Salt Dome	1,200	700	1,050	—	—	
Totals	58,512	32,152	30,532	13,990	12,763	
UTAH						
Phillips Petroleum Co.						
Davis-Frozen Earth	164	—	—	—	—	
Williams Energy Co.						
Grand-Salt Dome	—	—	—	—	250	Not currently in use
Totals	164	—	—	—	250	
VIRGINIA						
Washington Gas Light Co.						
Fairfax-Mined	—	—	—	—	302	

APPENDIX F

DETAILED INFORMATION ON U.S. IMPORT/EXPORT TERMINALS

APPENDIX F

DETAILED INFORMATION ON U.S. IMPORT/EXPORT TERMINALS

Atlantic Energy Corporation
2901 South Military Highway
P.O. Box 1156
Chesapeake, Virginia 23320
Phone (804) 485-1018

Sea Port:
Norfolk, Virginia

Berth:
Eastern Branch Elizabeth River

Ship Restrictions:
Length: 720'
Draught: 35' (Brackish)
Breadth: 110'

Products & Off-Loading Rate:
Propane
i Butane — 8,000-12,000 BPH
n Butane — (refrigerated)
Mixed Butanes

Unloading Facilities:
Unloading Connection: 12", 150# RF Flange
Dock Hose: 4"
Product Heating: Not required
Storage Capacity: Propane 220 m bbl
Butane 200 m bbl

Fractionation & Delivery:
No fractionation facilities; storage available for propane HD5; butane delivered only to SNG plant by pipeline; delivery to owner's tank truck or tank car.

Port Facility:
Port Authority: U.S. Coast Guard Marine Safety Office
Bunkers & Stores: Not available at dock

Comments:
Restricted lay time allows only limited supplies to be taken from barge or dock after completion of unloading operations.

California Liquid Gas Corporation
P.O. Box 28397
Sacramento, California 95828
Phone (916) 428-7010

Sea Port:
Ferndale, Washington (Cherry Point)

Berth:
Intalco Aluminum Co.

Ship Restrictions:
Length: 1,000'
Draught: 37' (Salt)
Breadth: No limit

Products & Off-Loading Rate:
Propane
n Butane
i Butane — 10,000 BPH (any single product)
Mixed Butanes
B-P

Unloading Facilities:
Unloading Connection: 12" ASA 150#
Dock Hose: 12" pipe
Product Heating: Not required
Storage Capacity: 350,000 bbl tank

Fractionation & Delivery:
No fractionation facilities; storage available for mixed product; delivery to importer by tank truck or tank car.

Port Facility:
Port Authority: Bellingham, Washington
Bunkers & Stores: No bunkers; stores can be delivered to dock.

Cities Service Company
P.O. Box 300
Tulsa, Oklahoma 74102
Phone (918) 586-2391

Sea Port:
Lake Charles, Louisiana—Calcasieu Ship Channel

Berth:
Hackberry

Ship Restrictions:
Length: 750'
Draught: 38' (Fresh)
Breadth: 125'

Products & Off-Loading Rate:
Propane
i Butane
n Butane — 10,000 BPH
Mixed Butanes

Unloading Facilities:
Unloading Connection: 10" 150#
Dock Hose: Rigid Arm
Product Heating: Required—heater provided at dock
Storage Capacity: 2,000,000 bbl

Fractionation & Delivery:
Fractionation facilities available. Propane to Dixie Pipeline; other products by pipeline to Houston area. Tank truck and tank car loading at Lake Charles. Exchange for all products at other domestic locations.

Port Facility:
Port Authority: Lake Charles Harbor Terminal
Bunkers & Stores: No bunkers; stores could be provided by barges

Comments:
The facility is estimated to be in operation during last quarter of 1979.

Coastal States Crude Gathering Company
Nine Greenway Plaza East
Houston, Texas 77046
Phone (713) 877-6504

Sea Port:
Corpus Christi, Texas

Berth:
Oil Dock 11

Ship Restrictions:
Length: None (dependent on adjacent ships)
Draught: 40' (Brackish)
Breadth:

Products & Off-Loading Rate:
Propane
i Butane
n Butane — 4,500 BPH
Mixed Butanes
BP Mix

Unloading Facilities:
Unloading Connection: 8" ASA 300#
Dock Hose: Provided at dock
Product Heating: Required; no heater at dock 35°F acceptable for unloading
Storage Capacity: LPG Mix—1,500 m bbl
Mixed Butanes—1,060 m bbl
Propane—6,600 m bbl

Fractionation & Delivery:
Fractionation facilities available; propane, n butane, and butane delivered back to importer; storage available for mixed product, purity product, n butane, propane, and i butane; delivery to importer by pipeline, tank truck, or tank car.

Port Facility:
Port Authority: Corpus Christi, Navigation District
Bunkers & Stores: Available at dock.

APPENDIX F (Continued)

DETAILED INFORMATION ON U.S. IMPORT/EXPORT TERMINALS

Dorchester Sea-3 Products, Inc.
2003 Capital National Bank Bldg.
Houston, Texas 77002
Phone (713) 658-0646

Sea Port:
Portsmouth, New Hampshire (Piscataqua River)

Berth:
Storage Tank Development Corp.

Ship Restrictions:
Length: 640'
Draught: 32' (Brackish)
Breadth: 100'

Products & Off-Loading Rate:
Propane—(Commercial grade or better)—12,500 BPH

Unloading Facility:
Unloading Connection: 16"—150# RF Flange Liquid
8"—150# RF Vapor
Dock Hose: Provided at dock
Product Heating: Not required
Storage Capacity: 400,000 bbl

Fractionation & Delivery:
Fractionation facilities not available; storage is provided for HD5 and commercial propane; delivery by truck loading racks and by rail loading racks.

Port Facility:
Port Authority: New Hampshire State Port Authority
Bunkers & Stores: Not available

Comments:
Marine traffic of LPG is subject to Coast Guard restrictions.

Exxon Company, U.S.A.
P.O. Box 2180
Houston, Texas 77001
Phone (713) 656-4054

Sea Port:
Everett, Massachusetts

Berth:

Ship Restrictions:
Length: 800-1,000'
Draught: 38' (Salt) Maximum size—360,000 bbl
Breadth: 120'

Products & Off-Loading Rate:
Propane: 10,000 BPH

Unloading Facilities:
Unloading Connection: 16" liquid line
10" vapor line, 150# ASA pressure rating
Dock Hose: Provided at dock
Product Heating: Not required
Storage Capacity: 400,000 bbl

Fractionation & Delivery:
Fractionation not available, storage provided for propane; delivery to importer at Everett, Massachusetts by tank truck.

Port Facility:
Port Authority: Coast Guard
Bunkers & Stores: Available at dock.

Gulf Oil Corporation
P.O. Box 1589
Tulsa, Oklahoma 74102
Phone (918) 584-7121

Sea Port:
Alliance Refinery

Berth:
New Orleans, Louisiana

Ship Restrictions:
Length: 650'
Draught: 40' (38' fresh at river bar entrance)
Breadth: None

Products & Off-Loading Rate:
i Butane
n Butane 4,000 BPH
Mixed Butanes

Unloading Facilities:
Unloading Connection: 8" 150# ASA
8" 300# ASA liquid
8" vapor
Dock Hose: Available at dock
Product Heating: Not required
Storage Capacity: Propane 72 m bbl
Butane 170 m bbl

Fractionation & Delivery:
Fractionation facilities available for butanes, 97.5% propane delivered back to importer; storage and delivery facilities not available to importers.

Port Facility:
Port Authority: New Orleans Port Commission
Bunkers & Stores: Not available

Comments:
Refinery requires butane supplies by marine transportation; on occasion can deliver propane to marine equipment.

Gulf Oil Corporation
P.O. Box 1589
Tulsa, Oklahoma 74102
Phone (918) 584-7121

Sea Port:
Philadelphia, Pennsylvania Gulf Refinery

Berth:
LPG-Girard Point Dock

Ship Restrictions:
Length: 735'
Draught: 32' (Brackish)
Breadth: 110'

Products & Off-Loading Rate:
n Butane
Mixed Butanes 1,500 BPH

Unloading Facilities:
Unloading Connection: 10" liquid
Dock Hose: Available at dock
Product Heating: Not required
Minimum Temperature: 28°F
Storage Capacity: Butane 200,000 bbl

Fractionation & Delivery:
No fractionation facilities available; storage provided for n butane; delivery to importer by tank truck and tank car.

Port Facility:
Port Authority: Philadelphia Gulf
Bunkers & Stores: Bunkers available.

APPENDIX F (Continued)

DETAILED INFORMATION ON U.S. IMPORT/EXPORT TERMINALS

Petrolane Inc.
P.O. Box 1410
Long Beach, California 90801
Phone (213) 427-5471

Sea Port:
Port of Los Angeles, California

Berth:
Berth 120

Ship Restrictions:
Length: None
Draught: 34' (Brackish)
Breadth: None
Water Depth: 35' mean low water

Products & Off-Loading Rate:
Propane
B-P 10,000 BPH
Mixed Butanes

Unloading Facility:
Unloading Connection: 12" 150# ANSI
Dock Hose: Provided
Product Heating: Not required
Storage Capacity: 600,000 bbl propane *or* 300,000 bbl propane and 300,000 bbl butane

Fractionation & Delivery:
No fractionation available; storage available for propane and butanes; delivery by tank truck and tank car at terminal.

Port Facility:
Port Authority: Port of Los Angeles
Bunkers & Stores: Available.

Petrolane Inc.
P.O. Box 1410
Long Beach, California 90801
Phone (213) 427-5471

Sea Port:
Providence, Rhode Island

Berth:
Municipal Dock, Berth #5

Ship Restrictions:
Length: None
Draught: 34' (Brackish)
Breadth: None

Products & Off-Loading Rate:
Propane 7,500 BPH

Unloading Facility:
Unloading Connection: 10" 150# ANSI
Dock Hose: Provided
Product Heating: Not required
Storage Capacity: 400,000 bbl

Fractionation & Delivery:
No fractionation available; storage available for propane; delivery by tank truck at terminal.

Port Facility:
Port Authority: Port of Providence
Bunkers & Stores: Bunkers not available; stores can be arranged.

Petro-Tex Chemical Corporation
P.O. Box 2584
Houston, Texas 77001
Phone (713) 477-9211

Sea Port:
Houston, Texas

Berth:
Sims Bayou Tanker Berth (Mile 395 on Houston Ship Channel)

Ship Restrictions:
Length: 674'
Draught: 36' (Fresh)
Breadth: 74'

Products & Off-Loading Rate:
Propane
i Butane
n Butane
Mixed Butanes 2,000 BPH
BP-Mix
Butadiene

Unloading Facility:
Unloading Connection. 6"—300#
Dock Hose: Provided at dock
Product Heating. Required; no heater at dock; —10°C acceptable for unloading
Storage Capacity: If pumping to Salt Dome Storage at Pierce Junction, no limitation.

Fractionation & Delivery:
Fractionation facilities not available; storage available for all products; delivery by pipeline, tank truck, tank car, or barge.

Port Facility:
Port Authority: Port of Houston
Bunkers & Stores: Available.

Phillips Petroleum Company
Bartlesville, Oklahoma 74004
Phone (918) 661-5023

Sea Port:
Houston, Texas

Berth:
Adams Terminal

Ship Restrictions:
Length: 750'
Draught: 36' (Fresh)
Breadth: 110'

Products & Off-Loading Rate:
Propane—5,000 BPH up to 200,000 bbl; then 1,800 BPH
i Butane
n Butane
Mixed Butanes 1,700 BPH
B-P Mix

Unloading Facility:
Unloading Connection: 10" and 8"—150#
Dock Hose: Provided at dock
Product Heating: Required; provided at dock
Storage Capacity: No limitation, pipeline delivery to Mont Belvieu storage.

Fractionation & Delivery:
Fractionation facilities available in 1980. Delivery by pipeline at Mont Belvieu; by tank car and tank truck (by prior arrangement) at Sweeny; by exchange to other domestic locations.

Port Facility:
Port Authority: Houston Port Authority
Bunkers & Stores: Not available.

Promix
P.O. Box 3387
Houston, Texas 77001
Phone (713) 623-3104

Sea Port:
Plaquemine, Louisiana

Berth:
(Privately owned)

Ship Restrictions:
Unknown

Products:
Propane
i Butane
n Butane
Mixed Butanes
B-P Mix

Unloading Facility:
Incomplete

Fractionation & Delivery:
Fractionation available for private use when operable; storage available for mixed product, purity product, propane, i butane, n butane; delivery available by pipeline, tank truck, and tank car to Breaux Bridge, Louisiana, and Napoleonville, Louisiana.

Port Facility:
Bunkers & Stores: Not available

Comments:
This is a privately owned facility—not a merchant terminal—Dock incomplete—due to unavailability of pipelines to storage and fractionation facilities.

Sun Gas Company
Sun Company, Inc.
100 Matsonford Road
Radnor, Pennsylvania 19087
Phone (215) 293-6429

Sea Port:
Marcus Hook, Pennsylvania

Berth:
Pier 3-C

Ship Restrictions:
Length: 1,000'
Draught: 38' (Fresh)
Breadth: None

Products & Off-Loading Rate:
Propane—18,000 BPH
i Butane
n Butane 9,000 BPH
Mixed Butanes

Unloading Facility:
Unloading Connection: Two 16" arms with 14" couplers and adapters for 12" 14", and 16" (150#)
Heater: Provided
Storage Capacity: Propane—1,170,000 bbl
Butane—400,000 bbl

Fractionation & Delivery:
No fractionation facilities; delivery by pipeline, tank car and tank truck.

Port Facility:
Port Authority: U.S. Coast Guard
Bunkers & Stores: Available.

Tropigas Inc. of Florida
1701 Ponce de Leon Boulevard
P.O. Box 341218
Coral Gables, Florida 33134
Phone (305) 446-5515

Sea Port:
Port Everglades, Florida

Berth:
Berths 5 & 6

Ship Restrictions:
Length: 378'
Draught: 25' (Salt)
Breadth: 65'

Products & Off-Loading Rate:
Propane—830 BPH

Unloading Facility:
Unloading Connection: 4"—300#
Dock Hose: Available at dock
Product Heating: Required, heater provided, minimum acceptable temperature for unloading —40°F
Storage Capacity: 10,500 bbl

Fractionation & Delivery:
No fractionation facilities; storage provided for propane; delivery by pipeline at Port Everglades.

Port Facility:
Port Authority: Port Everglades Port Authority
Bunkers & Stores: Available.

Warren Petroleum Company
P.O. Box 1589
Tulsa, Oklahoma 74102
Phone (918) 584-7121

Sea Port:
Port Everglades, Florida

Berth:
LPG Berth

Ship Restrictions:
Length: 700'
Draught: 39' (Salt)
Breadth: None

Products & Off-Loading Rate:
Propane—1,580 BPH

Unloading Facility:
Unloading Connection: 6"—300# liquid
4"—300# vapor
Dock Hose: Provided for barges only; 6"—300# fitting – 4"—300# vapor
Product Heating: Required; heater not provided, 10°F acceptable temperature of product
Storage Capacity: 38,000 bbl

Fractionation & Delivery:
No fractionation facilities available; propane HD5 delivery by tank truck at 5,000 bbl/day.

Port Facility:
Port Authority: Port Everglades Port Authority
Bunkers & Stores: Not available.

Warren Petroleum Company
P.O. Box 1589
Tulsa, Oklahoma 74102
Phone (918) 584-7121

Sea Port:
Houston, Texas

Berth:
Warren Gas Terminal

Ship Restrictions:
Length: 750'
Draught: 39' Fresh (channel restriction)
Breadth: 116' (channel restriction)

Products & Off-Loading Rate:
Propane—15,000 BPH (refrigerated); 4,000 BPH (ambient)
i Butane
n Butane
Mixed Butanes — 15,000 BPH (refrigerated)
B-P Mix
Naphtha/Natural Gasoline—5,000 BPH

Unloading Facility:
Unloading Connection: 16", 12", 8"—275# (liquid)
6"—275# vapor
Dock Hose: Loading arm provided at dock
Product Heating: Required; heater provided; 35°F minimum acceptable temperature
Storage Capacity: No limit

Fractionation & Delivery:
Fractionation available; any product delivered back to importer by pipeline, tank car, tank truck and barge.

Port Facility:
Port Authority: Port of Houston Authority
Bunkers & Stores: Bunkers available at dock and by barge; stores are not available at dock.

APPENDIX G

LPG/AIR PEAKSHAVING FACILITIES

APPENDIX G

LPG-AIR PEAKSHAVING FACILITIES

Company and location	Max. Dly. sendout (Mcf)	Year Inst-alled	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
ALABAMA						
Alabama Gas Corp.						
Birmingham, Ala.	24,000	1958	Man	1350	1.34	1,500,000
Montgomery, Ala.	10,000	1957	Man	1350	1.34	740,000
Gadsden, Ala.	12,000	1970	Rem Auto	1350	1.34	720,000
CITY OF ATHENS, ALA.						
Athens, Ala.	3,600	1962	Auto	1400	1.3	90,000
City of Florence						
Florence, Ala.	3,600	1966	Auto	1430	1.296	240,000
City of Fultondale, Ala.	3,600	1978	Auto	1400	1.287	90,000 approx
City of Graysville	3,600	1977	Auto	1400	1.287	9,000 approx
CITY OF HARTSELLE, ALA.						
Hartselle, Ala.	2,400	1963	Auto	1400	1.287	30,000
CITY OF HUNTSVILLE, ALA.						
Eighth Street	5,400	1960	Auto	1300	1.3	360,000
Oakwood Avenue	3,000	1963	Auto	1300	1.3	240,000
Jasper Utilities Bd.						
Jasper, Ala.	3,600	1972	Auto	1400	1.287	60,000
CITY OF LANETT						
Lanett, Ala.	16,200	1973	Auto	1400	1.287	60,000
NORTHWEST ALABAMA GAS DISTRICT						
Hamilton, Ala.	2,640	1965	Auto	1400	1.287	50,000
PHENIX NATURAL GAS SYSTEM						
Phenix City, Ala.	1,500	1960	Man	1300	1.2	90,000
CITY OF PLEASANT GROVE, ALA.						
Pleasant Grove, Ala.	3,456	1963	Auto	1400	1.287	120,000
Pleasant Grove, Ala.	3,600	1973	Auto	1400	1.287	
CITY OF RUSSELLVILLE, ALA.						
City of Russellville, Ala.	720	1964	Auto	1400	1.3	30,000
CITY of TRUSSVILLE						
Trussville, Ala	3,000	1971	Auto	1400	1.287	
SOUTHEAST ALABAMA GAS DISTRICT						
Dothan, Ala.	7,200	1977	Auto	1400	1.287	unknown
CALIFORNIA						
PACIFIC GAS & ELECTRIC CO.						
Willows, Calif.	1,440	1954	Auto	1325	1.3	30,000
Eureka, Calif.	8,400	1955	Auto	1325	1.3	420,000
Eureka Calif.	12,000	1976	Auto	1400	1.287	420,000
Arcata, Calif.	1,800	1954	Auto	1375	1.3	90,000
SAN DIEGO GAS & ELECTRIC CO.						
San Diego, Calif.	30,000	1949	Auto	1500	1.28	1,110,000
COLORADO						
GUNNISON UTILITIES						
Gunnison, Colo.	600	1968	Auto	1000	1.21	30,000
CONNECTICUT						
CONNECTICUT LIGHT & POWER CO.						
Danbury, Conn.	4,099	1970-1961	Auto	1400	1.29	324,000
Kensington, Conn.	7,380	1967	Auto	1600	1.33	480,000
Norwalk, Conn.	2,599	1960	Auto	1400	1.29	150,000
Shelton, Conn.	4,480	1966	Auto	1400	1.29	360,000
Vernon, Conn.	7,074	1964	Auto	1400	1.29	420,000
Waterbury, Conn.	12,810	1962	Auto	1600	1.33	968,000
CONNECTICUT NATURAL GAS CO.						
Hartford, Conn.	2,500	1950	Man	1400	1.29	180,000
New Britain, Conn.	4,800	1961	Man	1050-1400	1.21 1.29	300,000
Rocky Hill, Conn.	3,000	1964	Man	2560	1.52	480,000
HARTFORD ELECTRIC LIGHT CO.						
New London, Conn.	1,622	1965	Auto	1250	1.3	170,000
Stamford, Conn.	3,189	1948	Man	Varies	Varies	242,000
Torrington, Conn.	1,341	1964	Man & Auto	1250	1.25	108,800
SOUTHERN CONNECTICUT NATURAL GAS CO.						
Trumbull, Conn.	12,000	1950	Man	1300	1.21	600,000
Chapel St. Plant	7,000	1946	Auto	850	1.155	180,000
Conn. Coke Co.	8,000	1963	Man	2500	1.53	450,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
DELAWARE						
CHESAPEAKE UTILITIES CORP.						
Dover, Del.	1,200	1948-1970	Auto	1300	1.3	48,000
Seaford, Del.	400	1947	Auto	1300	1.3	30,000
DELMARVA POWER & LIGHT CO.						
Hockessin, Del.	24,000	1970	Auto	1400	1.287	587,400
Wilmington, Del.	33,600	1963	Auto	1400	1.287	840,000
FLORIDA						
CITY GAS CO. OF FLORIDA						
Hialeah, Fla.	4,800	1962	Auto	1400	1.287	
CLEARWATER MUNICIPAL GAS CO.						
Clearwater, Fla.	4,000	1952	Auto	1120	0.95	
CITY OF FT. MYERS						
Ft. Meyers	1,150	1973	Auto	1400	1.287	
PEOPLES GAS SYSTEM, INC.						
Ft. Lauderdale, Fla.	6,000	N.A.	Man	1450	1.35	18,000
N. Miami Beach, Fla.	4,000	N.A.	Man	1450	1.35	240,000
Tampa, Fla.	2,500	N.A.	Man	1450	1.35	60,000
CITY OF TALLAHASSEE, FLA.						
Tallahassee, Fla.	8,000	1966	Auto	1420	1.292	120,000
W. FLORIDA NG						
Panama City, FL	2,300	1973	Auto	1400	1.287	
GEORGIA						
AMERICUS UTILITY DIST.						
Americus Ga.	1,900	1971	Auto	1400	1.287	60,000
ATLANTA GAS LIGHT CO.						
Atlanta, Ga.	77,200	1947	Man	1450	1.300	3,840,000
Augusta, Ga.	11,600	1956	Man	1450	1.300	660,000
Macon, Ga.	19,200	1952	Man	1450	1.300	900,000
Riverdale, Ga.	67,600	1956	Man	1450	1.30	6,455,000
Rome, Ga.	7,800	1956	Man	1450	1.300	360,000
Savannah, Ga.	6,000	1958	Man	1350	1.274	240,000
Valdosta, Ga.	3,200	1952	Man	1450	1.300	300,000
Waycross, Ga.	1,940	1952	Man	1350	1.274	180,000
CITY OF CARTERSVILLE, GAS DEPT.						
Cartersville, Ga.	3,000	1968	Auto	1400	1.287	120,000
CITY OF COVINGTON, GAS DEPT.						
Covington, Ga.	3,600	1970	Auto	1400	1.287	60,000
CITY OF DOUGLAS						
Douglas, Ga.	1,900	1972	Auto	1400	1,287	120,000
CITY OF JACKSON						
Jackson, Ga.	1,150	1973	Auto	1400	1.287	30.000
GAS LIGHT CO. OF COLUMBUS (Ga.)						
Columbus, Ga.	6,000	1951	Man	1350	1.277	240,000
Columbus, Ga.	14,400	1957	Man	1350	1.277	900,000
CITY OF LAFAYETTE						
LaFayette, Ga.	2,250	1971	Auto	1400	1.287	60,000
CITY OF LA GRANGE, GA.						
La Grange, Ga.	7,200	1958	Man	1200-1340	Varies	240,000
CITY OF LAWRENCEVILLE GAS DEPT.						
Lawrenceville, Ga.	2,400	1970	Auto	1400	1.287	90,000
MADISON GAS DEPT.						
Madison, Ga.	1,150	1972	Auto	1400	1.287	30,000
CITY OF MONROE						
Monroe, Ga.	164,600	1972	Auto	1400	1.287	120,000
CITY OF SYLVESTER						
Sylvester, Ga.	600	1972	Auto	1400	1,287	30,000
CITY OF TOCCOA, GAS DEPT.						
Toccoa, Ga.	4,608	1970	Auto	1400	1.287	90,000
WATER GAS & LIGHT COMMISSION						
S.E. Albany, Ga.	8,700	1970	Auto	1350	1.29	810,000
West Albany, Ga.	5,200	1970	Semi Auto	1350	1.29	540,000
HAWAII						
HONOLULU GAS CO., LTD.						
Honolulu, Hawaii	1,000	1958	Man	1050	1.2	215,000
ILLINOIS						
CENTRAL ILLINOIS LIGHT CO.						
East Peoria, Ill.	10,000	1956	Semi	1350	0.885	960,000
Peoria, Ill.	5,000	1947	Semi	1350	0.885	660,000
CENTRAL ILLINOIS PUBLIC SERVICE CO.						
Desota, Ill.	5,000	1970	Semi Auto	1350	1.29	240,000
Hoopestown, Ill.	5,750	1972	Auto	1400	1.287	
Hoopeston, Ill.	2,000	1962	Semi Auto	1350	1.29	96,000
Marion, Ill.	7,000	1964-1970	Semi Auto	1350	1.29	270,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
ILLINOIS (Continued)						
Quincy, Ill.	15,000	1957	Semi Auto	1390	1.31	480,000
Quincy, Ill.	16,800	1972	Auto	1400	1.287	18,000
CITY OF FAIRFIELD, ILL.						
Fairfield, Ill.	1,440	1966	Auto	1350	1.277	60,000
CITY OF FLORIA						
Floria, Ill.	2,400	1973	Auto	1400	1,287	60,000
ILLINOIS POWER CO.						
Champaign, Ill.	15,000	1957	Semi	1380	1.31	720,000
Danville, Ill.	5,000	1947	Semi	1380	1.31	150,000
Galesburg, Ill.	9,500	1956	Semi	1380	1.31	540,000
Jacksonville, Ill.	12,000	1956	Semi	1380	1.31	150,000
Kewanee, Ill.	1,550	1949	Semi	1400	1.32	60,000
Monmouth, Ill.	500	1950	Semi	1380	1.31	60,000
INTERSTATE POWER CO.						
Savanna, Ill.	1,100	1962	Auto	1350	1.28	30,000
MID-ILLINOIS GAS CO.						
Rockford, Ill.	28,000	1949	Auto	1500	1.30	900,000
Lincoln, Ill.	1,000	1957	Auto	1350	1.25	60,000
Freeport, Ill.	1,000	1949	Auto	1500	1.30	60,000
MONARCH GAS CO.						
St. Elmo, Ill.	300	1949	Auto	1330	0.060	30,000
MT. CARMEL PUBLIC UTILITY CO.						
Mt. Carmel, Ill.	2,000	1968- 1969	Semi Auto	1400	1.29	180,000
NORTHERN ILLINOIS GAS CO.						
Skokie, Ill.	26,000	1962	Auto	1100	1.22	720,000
Mt. Prospect, Ill.	13,000	1949	Auto	1100	1.22	600,000
Forest View, Ill.	20,000	1949	Auto	1100	1.22	900,000
Matteson, Ill.	11,000	1948	Auto	1100	1.22	960,000
NORTH SHORE GAS CO.						
Libertyville, Ill.	29,500	1963- 1964 1966	Man Auto	1050	1.215	1,263,762
THE PEOPLES GAS LIGHT & COKE CO.						
Chicago, Ill.	N.A.	1934- 1938 1947 1957	Auto	1025	1.32	1,816,282
Chicago, Ill.	N.A.	1934- 1938 1947 1957	Auto	1025	1.21	1,776,015
UNION ELECTRIC CO.						
Alton, Ill.	6,000	1963	Auto	1450	1.2	180,000
INDIANA						
CENTRAL INDIANA GAS CO.						
Greenfield, Ind.	1,200	1954	Auto	1250		60,000
Marion, Ind.	7,200	1961	Man	1350	1.3	360,000
Muncie, Ind.	7,200	1950	Man	1350	1.3	360,000
CITIZENS GAS & COKE UTILITY						
Prospect Street	20,000	1948	Man	*	0.71	1,008,000
		(*Mixture of Natural Gas, Coke Oven Gas, L.P.G. and Air to a 900 Btu send-out gas.)				
INDIANA GAS CO.						
New Albany, Ind.	900	N.A.	Man	1350		30,000
Madison, Ind.	500	N.A.	Man	1350		30,000
New Castle, Ind.	2,400	N.A.	Auto	1350		120,000
Frankfort, Ind.	2,000	N.A.	Auto	1350		60,000
Greensburg, Ind.	2,400	N.A.	Man	1350		48,000
Lafayette #2, Ind.	5,650	N.A.	Auto	1350		210,000
Jeffersonville, Ind.	6,000	1978	Auto	1400	1.287	300,000
NORTHERN INDIANA FUEL & LIGHT CO.						
Auburn Junction, Ind.	2,500	1964	Man	1250- 1380	1.26- 1.28	120,000
NORTHERN INDIANA PUBLIC SERVICE CO.						
East Chicago, Ind.	28,000	N.A.	Auto	1300	1.27	550,000
South Bend, Ind.	21,580	N.A.	Auto	1300	1.27	275,000
Fort Wayne, Ind.	32,175	N.A.	Auto	1300	1.27	650,000
OHIO VALLEY GAS CORP.						
Winchester, Ind.	1,000	1950	Auto	1350	1.25	30,000
Connersville, Ind.	3,000	1964	Auto	1450	1.25	120,000
CITY OF RENSSELAER						
Rensselaer, Ind.	2,400	1972	Auto	1400	1.287	60,000
RICHMOND GAS CORP.						
Richmond, Ind.	1,800	1948	Man	1350		150,000
TERRE HAUTE GAS CORP.						
Terre Haute, Ind.	7,920	1947- 1953 1960	Auto	1260	1.26	300,000
IOWA						
CEDAR FALLS UTILITIES						
Cedar Falls, Iowa	9,600	1966- 1970	Auto	1360	.80	300,000
CENTRAL NATURAL GAS CO.						
Rock Rapids, Iowa	960	1966	Auto	1400	1.287	30,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
IOWA (Continued)						
COON RAPIDS MUNICIPAL UTILITIES						
Coon Rapids, Iowa	1,200			1350		24,000
EMMETTSBURG MUNICIPAL UTILITIES						
Emmettsburg, Iowa	1,728	1966	Auto	1400	1.29	49,000
GUTHRIE CENTER GAS DEPT.						
Guthrie Center, Iowa	1,200			1350		30,000
HARLAN MUNICIPAL UTILITIES						
Harlan, Iowa	1,200			1350		60,000
INTERSTATE POWER CO.						
Clinton, Iowa	4,000	1966	Auto	1400	1.29	240,000
Mason City, Iowa	12,000	1956	Auto	1350	1.28	450,000
IOWA-ILLINOIS GAS & ELECTRIC CO.						
Cedar Rapids, Iowa	14,000	1955	Auto	1350	1.28	450,000
Davenport, Iowa	12,600	1971	Auto	1400	1.287	360,000
IOWA POWER AND LIGHT CO.						
Des Moines, Iowa	31,000	1956- 1966	Man & Auto	1350- 1400	1.26	900,000
Des Moines, Iowa	2,400	1971	Auto	1400	1.287	60,000
Oskaloosa, Iowa	2,400	1972	Auto	1400	1.287	60,000
IOWA POWER SERVICE CO.						
Sioux City, Iowa	18,000	1967	Auto	1350	1.277	2,726,400
IOWA SOUTHERN UTILITIES						
Burlington, Iowa	6,860	1961	Man	1330	1.29	120,000
Grinnell, Iowa	860	1963	Auto	1330	1.29	66,000
Mt. Pleasant, Iowa	1,160	1961	Auto	1330	1.29	66,000
NORTH CENTRAL PUBLIC SERVICE CO.						
Algona, Iowa	2,000	1963	Auto	1280	1.3	60,000
Ft. Madison, Iowa	4,000	1964	Semi	1380	1.3	90,000
Perry, Iowa	1,000			1380	1.3	70,000
PEOPLES NATURAL GAS DIVISION						
Council Bluffs, Iowa	13,300	1949	Auto	1350	1.28	360,000
Dubuque, Iowa	12,000	1960	Auto	1350	1.28	360,000
Estherville, Iowa	3,300	1963	Auto	1350	1.28	90,000
Newton, Iowa	5,000	1948- 1969	Auto	1350	1.28	180,000
New Hampton, Iowa	1,000	1968	Auto	1350	1.28	30,000
CITY OF TIPTON						
Tipton, Iowa	600	1979	Auto	1400	1.287	30,000
KANSAS						
CITY OF HESSTON						
Hesston, Kansas	4,800	1972	Auto	1400	1.287	120,000
KENTUCKY						
CARROLTON LT. COMM.						
Carrolton, Ky.	4,350	1972	Auto	1400	1.287	90,000
COLUMBIA GAS OF KENTUCKY, INC.						
Lexington, Ky.	16,000	1947- 1961	Auto	1350	1.26	480,000
DELTA NATURAL GAS CO.						
Nicholasville, Ky.	2,400	1966	Auto	1400	1.287	30,000
ELIZABETHTOWN NATURAL GAS CO.						
Elizabethtown, Ky.	2,400	1965	Auto	1400	1.287	30,000
CITY OF HENDERSON						
Henderson, Ky.	4,550	1971	Auto	1400	1,287	150,000
UNION LIGHT, HEAT & POWER CO.						
Erlanger, Ky.	50,000	1961	Auto	1200	1.25	7,000,000
WESTERN KENTUCKY GAS CO.						
Mayfield, Ky.	2,500	1963	Auto	1350	1.277	90,000
Harrodsburg, Ky.	2,500	1965	Auto	1350	1.277	60,000
Stanford, Ky.	1,200	1969	Auto	1350	1.277	30,000
MAINE						
NORTHERN UTILITIES, INC.						
Portland, Maine	234	1968	Auto	1400	1.30	90,000
MARYLAND						
BALTIMORE GAS & ELECTRIC CO.						
Glen Arm, Md.	60,000	1963	Man Auto	1030	1.23	6,200,000
CHESAPEAKE UTILITIES CORP.						
Salisbury, Md.	700	1950	Auto	1300	1.3	60,000
COLUMBIA GAS OF MARYLAND						
Hagerstown, Md.	800	1949	Auto	1450	1.26	60,000
FREDERICK GAS CO., INC.						
Frederick, Md.	1,344	1962	Auto	1250	1.297	90,000
PENNSYLVANIA & SOUTHERN GAS CO.						
Elkton, Md.	1,000	1963	Auto	1400	1.29	30,000
WASHINGTON GAS LIGHT CO.						
Rockville, Md.	83,000	1952- 1960	Man	1025 1600	1.22- 1.34	3,460,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
MASSACHUSETTS						
BERKSHIRE GAS CO.						
Pittsfield, Mass.	10,000	1968	Auto	1350	1.277	300,000
Pittsfield, Mass.	4,800	1970	Auto	1350	1.277	300,000
Pittsfield, Mass.	1,200	1972	Auto	1400	1.287	30,000
Pittsfield, Mass.	2,300	1971	Auto	1400	1.287	
BOSTON GAS CO.						
Everett, Mass.	20,000	1969	Man	1400	1.30	1,050,000
Concord, Mass.	4,000	1953	Man	1400	1.30	150,000
Braintree, Mass.	6,960	1955	Man	1400	1.30	120,000
BROCKTON TAUTON GAS CO.						
Brockton, Mass.	21,900	1968	Rem Auto & Auto	1350	1.28	963,600
Tauton, Mass.	13,100	1964	Rem Auto & Auto	1350	1.28	401,500
W. Medway	4,400	1963	Rem Auto & Auto	1350	1.28	241,200
BUZZARDS BAY GAS CO.						
Cataumet, Mass.	2,800	1958	Semi Auto	1350	1.250	130,000
Chatham, Mass.	500	1960	Semi Auto	1350	1.250	60,000
Hvannis, Mass.	2,000	1967	Auto	1350	1.250	90,000
Hyannis, Mass.	3,000	1971	Auto	1400	1.287	
CAMBRIDGE GAS CO.						
Cambridge, Mass.	5,000	1948	Man	1350	1.25	180,000
CAPE COD GAS						
Cataumet, Mass.	4,800	1973	Auto	1400	1.287	
CENTRAL MASSACHUSETTS GAS CO.						
Southbridge, Mass.	2,750	1954	Auto	1400	1.3	180,000
FALL RIVER GAS CO.						
Fall River, Mass.	12,000	1947-1965	Man	1200	1.25-1.27	350,000
FITCHBURG GAS & ELECTRIC LIGHT CO.						
Lunenburg, Mass.	4,800	1960	Auto	1250	1.26	363,000
HAVERHILL GAS CO.						
Haverhill, Mass.	12,000	1960	Auto	1150-1400	1.23-1.28	542,500
HOLYOKE GAS AND ELECTRIC DEPT.						
Holyoke, Mass.	4,800	1976	Auto	1400	1.287	
Old Gas Works Holyoke, Mass.	4,440	1951	Auto	1380	1.3	270,000
Mueller Rd., Holyoke, Mass.	8,880	1965	Auto	1380	1.3	201,000
LAWRENCE GAS CO.						
Lawrence, Mass.	9,000	1949-1960	Auto	1350	1.28	390,000
LOWELL GAS CO.						
Lowell, Mass.	7,000	1956-1964	Auto	1300	1.28	450,000
Tewksbury, Mass.	4,940	1960	Auto	1300	1.28	120,000
Pepperell, Mass.	60	1958	Man	1330	1.30	30,000
MIDDLEBORO GAS & ELECTRIC DEPT.						
Middleboro, Mass.	60	1954	Man	1350	1.3	60,000
MYSTIC VALLEY GAS						
Malden, Mass.	12,000	1946	Auto	1250	1.26	390,000
Reading, Mass.	5,500	1951	Auto	1400	1.3	180,000
Revere, Mass.	5,500	1960	Auto	1400	1.3	180,000
NEW BEDFORD GAS & EDISON LIGHT						
Water St., New Bedford, Mass.	1,000	1947	Man	1225	1.25	30,000
Shawmut Ave., New Bedford, Mass.	5,800	1963-1969	Man	1225	1.25	150,000
Plymouth, Mass.	4,471	1951-1959-1960	Auto	1300	1.27	120,000
NORTHHAMPTON GAS LIGHT CO.						
Northhampton, Mass.	3,000	1960	Auto	1400	1.3	300,000
NORTHSHORE GAS CO. (MASS.)						
Salem, Mass.	3,000	1947	Man	1400	1.3	60,000
Davers, Mass.	4,800	1952	Auto	1400	1.3	150,000
Gloucester, Mass.	3,700	1952-1969	Auto	1400	1.3	120,000
NORWOOD GAS CO.						
Norwood, Mass.	4,400	1958-1970	Auto	1400	1.3	180,000
SPRINGFIELD GAS LIGHT CO.						
East Longmeadow, Mass.	12,000	1966-1967	Auto	1050-1400		720,000
West Springfield, Mass.	9,000	1960	Man	1050-1400		960,000
WACHUSETT GAS CO.						
Leominster, Mass.	3,080	1952	Auto	1400	1.3	120,000
WORCESTER GAS LIGHT CO.						
Worcester, Mass.	20,500	1970	Auto	1400	1.287	480,000
Framingham, Mass.	6,720	N.A.	Semi	1225	1.25	180,000
Hyde Park, Mass.	7,776	N.A.	Semi	1225	1.25	210,000
Hudson, Mass.	5,184	N.A.	Auto	1225	1.25	180,000
Marlboro, Mass.	2,496	N.A.	Semi	1225	1.25	90,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
MICHIGAN						
BATTLE CREEK GAS CO.						
Battle Creek, Mich. #1	3,400	1947	Man			360,000
Battle Creek, Mich. #2	3,600					
CITIZENS GAS FUEL CO.						
Adrian, Mich.	2,000	1947	Semi	2530	1.50	60,000
Adrian, Mich.	8,000	1973	Auto	1400	1.287	300,000
CONSUMERS POWER CO.						
Jackson, Mich.	3,800	1967	Auto	1400	1.287	180,000
CITY OF LANSING						
Lansing, Mich.	1,150	1974	Auto	1400	1.287	90,000
MICHIGAN GAS & ELECTRIC CO.						
Three Rivers, Mich.	240	1953	Man	1375		30,000
Marquette, Mich.	2,400	1955	Man	1375		60,000
MINNESOTA						
CENTRAL NATURAL GAS CO.						
Luverne, Minn.	1,800	1970	Man	1400	1.28	60,000
Pipestone, Minn.	1,800	1969	Man	1400	1.28	30,000
Waseca, Minn.	2,880	1957	Man	1400	1.28	120,000
Hastings, Minn.	3,360	1954	Man	1400	1.28	90,000
Hastings, Minn.	9,000	1972	Auto	1400	1.287	
INTERSTATE POWER CO.						
Albert Lea, Minn.	6,000	1966	Auto	1350	1.28	330,000
INTER-CITY GAS CO., LTD.						
Cloquet, Minn.	1,520	1970	Auto	1365	1.277	30,000
MINNEAPOLIS GAS CO.						
325 W. 61st St. Mpls, Minn.	24,000	1947	Man	1415	1.29	1,200,000
2601 N. Pacific Mpls, Minn.	24,000	1947	Man	1415	1.29	1,200,000
Golden Valley, Minn.	24,000	1953	Man	1415	1.29	1,200,000
New Hope, Minn.	24,000	1959	Man	1415	1.29	1,260,000
#2 Cedar Ave., Mpls, Minn.	36,000	1959	Man	1415	1.29	1,500,000
Dakota County, Minn.	48,000	1961	Man	1415	1.29	5,640,000
MINNESOTA NATURAL GAS CO.						
Mankato, Minn.	5,000	1948	Auto	1365	1.28	360,000
Minneapolis, Minn.	14,000	1971	Auto	1400	1.287	
Willmar, Minn.	3,700	1972	Auto	1400	1.287	
Brainerd, Minn.	3,600	1973	Auto	1400	1.287	
St. Peter, Minn.	3,600	1973	Auto	1400	1.287	
NORTH CENTRAL PUBLIC SERVICE CO.						
Coon Rapids, Minn.	5,000	1959	Semi	1380	1.3	150,000
NORTHERN STATES POWER CO.						
St. Cloud, Minn.	5,000	1963	Man	1350	Varies	120,000
St. Paul, Minn.-Sibley	48,000	1952	Auto	1350	Varies	7,160,000
St. Paul, Minn.-Maplewood	30,000	1957	Auto	1350	Varies	1,530,000
Mobil Plant	3,500	1964	Auto	1350	1.28	(see below)
Faribault, Minn.	(storage only)					90,000
Winona, Minn.	(storage only)					90,000
PEOPLES NATURAL GAS DIVISION						
Rochester, Minn.	11,500	1948	Auto	1350	1.28	360,000
Detroit Lakes, Minn.	1,000	1962	Auto	1350	1.28	60,000
PUBLIC UTILITIES COMMISSION						
Hibbing, Minn.	224	1946	Auto	1370		60,000
MISSISSIPPI						
MISSISSIPPI VALLEY GAS CO.						
Meridian, Miss.	1,920	1948	Auto	1350	1.28	210,000
UNITED GAS PIPE LINE CO.						
Philadelphia, Miss.	1,500	1964	Auto	1400	1.28	60,000
Newton, Miss.	1,200	1966	Auto	1400	1.28	60,000
MISSOURI						
TOWN OF BAY SPRINGS						
Bay Springs, Mo.	600	1972	Auto	1400	1.287	18,000
CITY OF BERNIE						
Bernie, Mo.	1,200	1976	Auto	1350	1.287	
GREAT RIVER GAS CO.						
Hannibal, Mo.	1,000	1959	Semi Auto	1350	1.28	60,000
KANSAS CITY POWER & LT.						
Kansas City, Mo.	7,200	1972*	Auto	1400	1.287	540,000
Kansas City, Mo.	2,150	1972	Auto	1400	1.287	150,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
MISSOURI (Continued)						
MISSOURI POWER & LIGHT CO.						
Jefferson City, Mo.	7,500	1958	Semi	1350	1.3	360,000
MISSOURI-EDISON UTILITY CO.						
Louisiana, Mo.	3,200	1972	Auto	1400	1.287	150,000
MISSOURI UTILITIES CO.						
Cape Girardeau, Mo.	4,500	1970	Auto	1350	1.3	180,000
Cape Girardeau, Mo.	3,000	1958	Auto	1350	1.30	150,000
Columbia, Mo.	3,000	1958	Auto	1350	1.30	150,000
NEBRASKA						
CENTRAL TELEPHONE & UTILITIES CORP.						
Columbus, Nebr.	1,000	1948	Man	1350	1.26	60,000
Lincoln, Nebr.	30,000	1949-1955	Auto	1350	1.26	1,080,000
Lincoln, Nebr.	4,550	1972	Auto	1400	1.287	150,000
Greenwood, Nebr.		(Leased cavern storage)				
HASTINGS UTILITIES						
Hastings, Nebr.	9,000	1964-1969	Auto	1300	1.267	390,000
KANSAS-NEBRASKA NG CO.						
O'Neill, Neb.	1,200	1978	Auto	1400	1.237	60,000
METROPOLITAN UTILITIES DISTRICT						
63, & Oak, Omaha, Nebr.	32,000	1961	Auto	1315	1.27	6,900,000
117 & Fort, Omaha, Nebr.	34,000	1964	Auto	1315	1.27	10,500,000
27th & Child's Omaha, Nebr.	10,000	1966	Auto	1315	1.27	660,000
NEBRASKA NG CO.						
Fremont, Neb.	1,800	1977	Auto	1400	1.287	120,000
NORTHWESTERN PUBLIC SERVICE VO.						
Kearney, Nebr.	6,000	1965	Auto	1250	1.26	180,000
North Platte, Nebr.	3,375	1968	Auto	1250	1.26	90,000
NEW HAMPSHIRE						
MANCHESTER GAS CO.						
Manchester, N.H.	12,200	1966	Rem Auto	1350	1.27	360,000
NORTHERN UTILITIES, INC.						
Portsmouth, N.H.	72	1966	Auto	1400	1.30	30,000
Exeter, N.H.	48	1964-1967	Man	1400	1.30	4,000
NEW JERSEY						
ELIZABETH GAS CO.						
New Village, N.J.	2,800	1969	Man	1350	1.27	60,000
NEW JERSEY NATURAL GAS CO.						
Atlantic Highlands, N. J.	2,400	1947-1951	Semi	1350	1.29	60,000
Long Branch, N.J.	9,600	1948	Semi	1350	1.29	60,000
Toms River, N.J.	3,000	1960	Semi	1350	1.29	180,000
Matawan, N.J.	3,000	1952	Semi	1350	1.29	90,000
Wall Township, N.J.	12,000	1970	Semi	1350	1.29	360,000
Ocean City, N.J.	2,400	1969	Semi	1350	1.29	90,000
Middle Township, N.J.	3,000	1969	Semi	1350	1.29	90,000
PUBLIC SERVICE ELECTRIC & GAS CO.						
Harrison, N. J.	48,600	1955	Auto	1600	1.35	1,224,000
Jersey City, N.J.	13,500	1947	Auto	1600	1.35	360,000
Edison Township, N.J.	33,900	1947	Auto	1600	1.35	900,000
Paterson, N. J.	5,600	1951	Auto	1600	1.35	180,000
SOUTH JERSEY GAS CO.						
McKee City, N.J.	12,000	1969	Auto	1350	1.26	450,000
Glassboro, N.J.	1,680	1948	Man	1350	1.26	90,000
Millville, N.J.	720	1962	Man	1350	1.26	60,000
NEW MEXICO						
SOUTHERN UNITED GAS CO.						
Albuquerque, N.M.	1,000	1951	Semi & Auto	1460		300,000
NEW YORK						
THE BROOKLYN UNION GAS CO.						
Greenpoint, N.Y.	30,000	1947	Man	1040	1.32	840,000
COLUMBIA GAS OF NEW YORK, INC.						
Binghamton, N.Y.	11,200	1946	Auto	1450	1.26	840,000
Olean, N.Y.	5,000	1947	Auto	1530	1.31	360,000
Johnson City, N.Y.	8,000	1956	Auto	1450	1.26	360,000
MUNICIPAL UTILITY COMM.						
Bath, N.Y.	2,400	1974	Auto	1400	1.287	60,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Inst-alled	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
NEW YORK (Continued)						
ORANGE AND ROCKLAND UTILITIES, INC.						
Orangeburg, N.Y.	16,000	1943	Rem Auto	1450	1.30	485,600
Suffern, N.Y.	6,048	1953	Man	1100- 1450	1.23- 1.30	180,000
Middletown, N.Y.	7,500	1960	Rem Auto	1450	1.30	312,400
NORTH CAROLINA						
CITY OF LEXINGTON, N.C.						
Lexington, N.C.	1,900	1963- 1964	Auto	1400	1.29	60,000
CITY OF SHELBY, N.C.						
Shelby, N.C.	1,920	1965	Auto	1400	1.27	90,000
GREENVILLE UTILITIES COMMISSION						
Greenville, N.C.	1,500	1940	Man	1400	1.32	102,000
NORTH CAROLINA GAS SERVICE CO. (Pennsylvania & Southern Gas Co.)						
Reidsville, N.C.	500	1947- 1970	Man	1060- 1390	1.23- 1.3	90,000
Reidsville, N.C.	1,200	1971	Auto	1400	1.287	
Reidsville, N.C.	6,200	1975	Auto	1350	1.287	
PIEDMONT NATURAL GAS CO., INC.						
Charlotte, N.C.	20,000	1959	Auto	1200- 1350	1.251 1.282	852,000
PUBLIC SERV. CO. OF NORTH CAROLINA						
Cary, N.C.	26,950	1972	Auto	1400	1.287	
Gastonia, N.C.	1,800	1971	Auto	1400	1.287	750,000
UNITED CITIES GAS CO.						
Hendersonville, N.C.	2,000	1968	Auto	1400	1.287	15,000
NORTH DAKOTA						
NORTHERN STATES POWER CO.						
Fargo, N.D.	5,000	1962	Man	1350	Varies	150,000
Grand Forks, N.D.	5,000	1967	Auto	1350	1.28	1,000,000
OHIO						
CINCINNATI GAS & ELECTRIC CO.						
Middletown, Ohio	50,000	1959	Auto	1200	1.25	7,000,000
Cincinnati, Ohio	65,000	1963	Auto	1200	1.25	8,000,000
CITY OF HAMILTON GAS DEPT.						
Hamilton, Ohio	14,000	1970	Man	1425	1.299	180,000
DAYTON POWER & LIGHT CO.						
Kemp, Ohio	20,000	1959	Auto	1400	1.3	660,000
Yankee, Ohio	22,000	1959	Auto	1400	1.3	79,200
Bellbrook, Ohio	28,000	1960	Auto	1400	1.3	79,200
OHIO FUEL GAS CO.						
Mt. Sterling, Ohio	35,000	1946	Auto	1530	1.3	2,100,000
Perrysburg, Ohio	27,600	1946	Auto	1450	1.3	1,800,000
OHIO GAS CO.						
Bryan, Ohio	6,000	1956	Auto	1450	1.25	210,000
Bryan, Ohio	20,300	1971	Auto	1400	1.287	120,000
Napoleon, Ohio	6,750	1972	Auto	1400	1.287	300.000
PIKE NATURAL GAS CO.						
Hillsboro, Ohio	1,844	1969	Auto	1400	1.287	30,000
Waverly, Ohio	1,844	1969	Auto	1400	1.287	30,000
WEST OHIO GAS CO.						
Van Wert, Ohio	4,500	1964	Auto	1430	1.296	90,000
St. Marys' Ohio	6,000	1965	Auto	1430	1.296	180,000
Lima, Ohio	14,000	1968	Auto	1430	1.296	360,000
PENNSYLVANIA						
COLUMBIA GAS OF PA., INC.						
Ellwood City, Pa.	7,000	1947	Auto	1530	1.31	480,000
Bradford, Pa.	4,000	1947	Auto	1530	1.31	630,000
PENN FUEL GAS INC.						
Bangor, Pa.	150		Man	1300	1.3	18,000
Lewisburg, Pa.	150		Man	1300	1.3	30,000
Lockhaven, Pa.	75		Man	1300	1.3	30,000
Oxford, Pa.	200		Man	1300	1.3	15,000
Pittston, Pa.	1,500		Man	1300	1.3	36,000
Pottsville, Pa.	250		Man	1300	1.3	30,000
Shamokin, Pa.	100		Man	1300	1.3	30,000
Stroudsburg, Pa.	250		Man	1300	1.3	30,000
PENNSYLVANIA & SOUTHERN GAS						
Sayre, Pa.	1,450	1971	Auto	1400	1.287	
PENNSYLVANIA GAS & WATER CO.						
Wilkes Barre, Pa.	5,000	1969	Auto	1400	1.287	180,000
Williamsport, Pa.	5,000	1969	Auto	1400	1.287	180,000
PHILADELPHIA ELECTRIC CO.						
W. Canshohocken, Pa.	23,000	1945 Enlarged 1946 1947 1949	Man	1525	1.32	520,000
Chester, Pa.	59,000	1947 Enlarged 1949 1952 1961	Man	1525	1.32	520,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
PENNSYLVANIA (Continued)						
UGI CORP. (Philadelphia Gas Works, Div.)						
31st & Passyunk Ave., Phila.	39,235	1951	Man & Auto	1500	1.32	780,000
3100 E. Venango St., Phila.	39,235	1951	Man & Auto	1500	1.32	780,000
UGI CORP. (Harrisburg Gas Division)						
Steelton, Pa.	4,800	1948	Man	1400	1.30	830,000
Steelton, Pa.	4,800	1956	Man	1400	1.30	
UGI CORP. (Lehigh Valley Gas Division)						
Bethlehem, Pa.	10,000	1953	Auto	1300	0.70	649,000
UGI CORP. (Reading Gas Division)						
Reading, Pa.	8,000	1948-1970	Auto	1300	1.5	300,000
RHODE ISLAND						
THE NEWPORT GAS LIGHT CO.						
Portsmouth, R.I.	4,800	1964	Rem Auto	1200-1400	1.226-1.266	180,000
PROVIDENCE GAS CO.						
Allens Ave. (Low Pressure Jet Plant)	26,750	1951	Semi	1400	1.30	1,215,000
Allens Ave. (High Pressure Plant)	12,600	1964	Semi	1400	1.30	184,200
SOUTH COUNTY GAS CO.						
Westerly, R.I.	3,984	1953 1970	Man & Semi Auto	1450	1.34	66,000
VALLEY GAS CO.						
Pawtucket, R.I.	15,000	1960	Auto	1350	1.277	210,000
Cumberland, R.I.	12,000	1960	Auto	Varies	Varies	717,800
SOUTH CAROLINA						
CAROLINA PIPELINE CO.						
Cheraw, S.C.	6,000	1962-1970	Auto	1400	1.287	120,000
Camden, S.C.	600	1963	Auto	1380	1.28	18,000
Dillon, S.C.	500	1963	Auto	1380	1.28	18,000
Abbeville, S.C.	800	1963	Auto	1380	1.28	18,000
Columbia, S.C.	18,000	1972	Auto	1400	1.287	
CLINTON-NEWBERRY NATURAL GAS AUTHORITY						
Newberry, S.C.	2,400	1958	Auto	1340	1.28	90,000
Clinton, S.C.	2,400	1977	Auto	1400	1.287	
CITY OF LAURENS GAS DEPT.						
Laurens, S.C.	1,440	1964	Auto	1400	1.28	90,000
COMM. OF PUBLIC WORKS						
Greer, S.C.	2,250	1971	Auto	1400	1.287	60,000
DEPT. OF PUBLIC UTILITIES						
Orangeburg, S.C.	9,100	1972	Auto	1400	1.287	210,000
CITY OF GREENWOOD						
Greenwood, S.C.	2,300	1973	Auto	1400	1.287	150,000
PEOPLES NATURAL GAS CO. OF S.C.						
Florence, S.C.	3,000	1968	Auto	1450	1.297	180,000
Sumter, S.C.	400	1948	Man	1325		48,000
SOUTH CAROLINA ELECTRIC & GAS CO.						
Charleston, S.C.	9,600	1965	Auto	1400	1.3	1,320,000
Charleston, S.C.	3,200	N.A.	Auto	1400	1.3	150,000
Columbia, S.C.	10,000	1970	Auto	1400	1.3	1,020,000
Columbia, S.C.	4,200	N.A.	Auto	1400	1.3	150,000
Charleston, S.C.	9,950	1971	Auto	1400	1.287	600,000
Columbia, S.C.	6,000	1971	Auto	1400	1.287	360,000
Columbia, S.C.	9,650	1972	Auto	1400	1.287	360,000
Aiken, S.C.	3,000	1974	Auto	1400	1.287	120,000
West Ashley, S.C.	3,000	1974	Auto	1400	1.287	120,000
SOUTH DAKOTA						
CENTRAL NATURAL GAS CO.						
Vermillion, S.D.	2,400	1961-1966	Auto	1400	1.287	120,000
Vermillion, S.D.	4,800	1971	Auto	1400	1.287	
CENTRAL TELEPHONE & UTILITIES CORP.						
Sioux Falls, S.D.	11,000	1950-1955	Auto	1350	1.26	480,000
NORTHWESTERN PUBLIC SERVICE CO.						
Aberdeen, S.D.	5,000	1959	Man	1250	1.26	210,000
Brookings, S.D.	1,500	1963	Auto	1250	1.26	60,000
Huron, S.D.	2,725	1952	Man	1250	1.26	180,000
Mitchell, S.D.	2,500	1953	Man	1250	1.26	180,000
TENNESSEE						
CITY OF ATHENS GAS DEPT.						
Athens, Tenn.	3,600	1968	Auto	1430	1.292	30,000
CHATTANOOGA GAS CO.						
Chattanooga, Tenn.	14,400	1954	Semi	1100	1.2	662,000
Cleveland, Tenn.	1,200	1965	Auto	1100	1.2	30,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
TENNESSEE (Continued)						
CLARKSVILLE GAS DEPT.						
Clarksville, Tenn.	7,700	1972	Auto	1400	1.287	240,000
COOKEVILLE, TENN. GAS DEPT.						
Cookeville, Tenn.	3,600	1963	Auto	1400	1.287	150,000
Cookville, Tenn.	1,150	1971	Auto	1400	1.287	
TOWN OF DICKSON						
Dickson, Tenn.	3,000	1973	Auto	1400	1.287	60,000
DYERSBURG GAS DEPARTMENT						
Dyersburg, Tenn.	6,000	1969	Auto	1400	1.287	120,000
ELK RIVER PUBLIC UTILITIES						
Tullahoma, Tenn.	2,400	1964	Auto	1400	1.287	90,000
GALLATIN NG						
Gallatin, Tenn.	4,800	1973	Auto	1400	1.287	60,000
GIBSON COUNTY UTILITY DISTRICT						
Trenton, Tenn. #1	2,400	1965	Auto	1350	1.277	120,000
Trenton, Tenn. #2	2,400	1970	Auto	1350	1.287	120,000
HUMBOLDT GAS & WATER						
Humboldt, Tenn.	2,300	1971	Auto	1400	1.287	
JACKSON UTILITY GAS DEPT.						
Jackson, Tenn.	5,400	1951	Auto	1375		300,000
JACKSON UTILITY DIV.						
Jackson, Tenn.	14,400	1974	Auto	1400	1.287	
KNOXVILLE UTILITIES BOARD						
Knoxville, Tenn.	11,000	1947	Auto	1350	1.27	620,000
Knoxville, Tenn.	30,300	1972	Auto	1350	1.287	
NASHVILLE GAS CO.						
Nashville, Tenn.	14,400	1972	Auto	1400	1.287	
OAK RIDGE UTILITY DISTRICT						
Oak Ridge, Tenn.	2,500	1960	Man & Auto	1400	1.29	120,000
CITY OF PARSONS						
Parsons, Tenn.	4,550	1972	Auto	1400	1.287	60,000
POWELL-CLINCH UTILITY DIST.						
Lake City, Tenn.	2,250	1972	Auto	1400	1.287	
UNITED CITIES GAS CO.						
Maryville, Tenn.	2,000	1965	Auto	1400	1.287	30,000
Bristol, Tenn.	2,250	1970	Auto	1400	1.287	
Nashville, Tenn.	2,250	1971	Auto	1400	1.287	
UNITED CITIES						
Morristown, Tenn.	7,200	1976	Auto	1400	1.287	480,000
Shelbyville, Tenn.	2,000	1966	Auto	1400	1.287	30,000
VOLUNTEER NATURAL GAS CO.						
Johnson City, Tenn.	3,000	1960	Man	1360	1.28	120,000
Kingsport, Tenn.	1,500	1965	Man	1360	1.28	120,000
Tennessee Tank (Storage)						30,000
TEXAS						
Lone Star Gas						
Dallas, Tx.	4,400	1971	Auto	1400	1.287	
VIRGINIA						
CITY OF CHARLOTTESVILLE GAS DEPT.						
Charlottesville, Va.	4,560	1970	Auto	1400	1.287	300,000
COLONIAL NG CO.						
Radford, Va.	4,550	1972	Auto	1400	1.287	
Blacksburg, Va.	4,650	1973	Auto	1400	1.287	120,000
Wytheville, Va.	6,550	1973	Auto	1400	1.287	120,000
COMMONWEALTH NATURAL GAS CO.						
Fredericksburg, Va.	1,200	1970	Auto	1400	1.287	30,000
Richmond, Va.	1.200	1970	Auto	1400	1.287	30,000
CITY OF DANVILLE, VA. GAS DEPT.						
Danville, Va.	5,000	1970	Auto	1400	1.3	270,000
PORTSMOUTH GAS CO.						
Portsmouth, Va.	9,000	1958-1963	Man	1200	1.5	198,000
ROANOKE GAS CO.						
Plantation Rd., Roanoke, Va.	3,600	1961	Man & Auto	1425		180,000
Main Plant, Roanoke, Va.	2,400	1956	Man	1425		240,000
SOUTHWESTERN VA. GAS CO.						
Martinsville, Va.	3,250	1973	Auto	1400	1.287	30,000
SUFFOLK GAS CORP.						
Suffolk, Va.	200	1945	Man	1350	0.65	36,000
VIRGINIA ELECTRIC & POWER CO.						
Norfolk, Va.	12,000	1959	Auto	1200	1.250	696,000
Norfolk, Va.	24,000	1964	Auto	1200	1.250	2,250,000
Newport News, Va.	6,000	1961-1970	Auto	1200	1.250	1,280,000

APPENDIX G (Continued)

Company and location	Max. Dly. sendout (Mcf)	Year Installed	Type of Controls Used	Btu/cf of gas made	Sp. Gr. of gas made	Storage at plant (gal.)
VIRGINIA (Continued)						
VIRGINIA GAS DISTRIBUTION CORP.						
Staunton, Va.	8,400	1961	Auto	1350	1.26	180,000
WASHINGTON GAS LIGHT CO.						
Ravensworth, Fairfax Co., Va.	133,000	1963	Rem Man	1025– 1600	1.22– 1.34	12,700,000
WASHINGTON						
WASHINGTON NATURAL GAS CO.						
Renton, Wash.	330,000	1965	Man	1470	1.3	1,035,400
Seattle, Wash.	20,700	1948	Auto	1470	1.3	102,000
Chehalis, Wash.	11,500	1948	Auto	1470	1.3	17,000
WEST VIRGINIA						
COLUMBIA GAS OF WEST VA., INC.						
Elk, W. Va.	24,000	1947	Auto	1350	1.26	1,140,000
Kenova, W. Va.	8,000	1947	Auto	1350	1.26	1,110,000
Charleston, W. Va.	16,100	1971	Auto	1400	1.287	
UNITED FUEL GAS						
Elk River, W. Va.	24,000	1970	Auto	1400	1.287	
WISCONSIN						
MADISON GAS & ELECTRIC CO.						
Summit, Wisc.	500	1958	Auto	1350	1.28	360,000
Patterson St., Wisc.	343	1966	Auto	1350	1.28	360,000
NORTH CENTRAL PUBLIC SERVICE CO.						
Monroe, Wisc.	2,000	1964	Auto	1380	1.3	30,000
NORTHERN STATES POWER CO.						
LaCrosse, Wisc.	5,000	1961	Man	1350		210,000
SUPERIOR WATER, LIGHT & POWER CO.						
Superior, Wisc.	2,500	1959	Auto	1420	1.29	180,000
WISCONSIN FUEL AND LIGHT CO.						
Manitowoc, Wisc.	3,000	1964	Auto	1350	1.287	90,000
Wausau, Wisc.	400	1955	Man	1350	1.287	91,000
WISCONSIN GAS COMPANY						
Sparta, Wisc.	410	1930	Man	1350	1.28	30,000
Tomah, Wisc.	410	1930	Man	1350	1.28	30,000
WISCONSIN NATURAL GAS CO.						
Ft. Atkinson, Wisc.	1,000	1948	Man	1350	1.28	144,000
Watertown, Wisc.	1,000	1948	Man	1350	1.28	216,000
Waukesha, Wisc.	4,000	1964	Auto	1350	1.28	180,000
WISCONSIN POWER & LIGHT CO.						
Beloit, Wisc.	1,300	1949	Semi	1350	1.1	240,000
Fond du Lac, Wisc.	1,500	1948	Semi	1350	1.1	240,000
Baraboo, Wisc.	240	1940	Semi	1350	1.1	30,000
Portage, Wisc.	240	1940	Semi	1350	1.1	30,000
Platteville, Wisc	240	1947	Semi	1350	1.1	30,000
WISCONSIN PUBLIC SERVICE CORP.						
Oshkosh, Wisc.	5,000	1930– 1945	Auto	1050– 1425	.605– .970	210,000
Sheboygan, Wisc.	7,800	1930– 1945	Auto	1050– 1425	.605	210,000
WISCONSIN SOUTHERN GAS CO., INC.						
Burlington, Wisc.	3,000		Semi	1200		180,000
Lake Geneva, Wisc.	1,500		Semi	1200		120,000

APPENDIX H

LPG INDUSTRY ACCIDENT HISTORY

APPENDIX H

LPG INDUSTRY ACCIDENT HISTORY

The following descriptions of representative LPG industry accident events are based on information from published literature and trade associations.

REFINERIES AND NATURAL GAS EXTRACTION PLANTS

- HUNTINGTON BEACH, CALIFORNIA. JUNE 2, 1934. A 5000-gallon truck was being loaded with butane by gravity through a closed system from a vertical 15,000-gallon storage tank by means of a three-inch line. The vapor spaces of the storage tank and the truck tank were interconnected by means of a one-inch pipe and one-inch hose so as to equalize the pressure, which was 65 psig, and permit the return flow of the gas. These tanks, however, had no excess-flow valves installed. The hose in question was of new synthetic rubber, but it had not been tested after the couplings had been applied, and appears to have blown off the coupling nipple in some unknown way. The valve at the base of the storage tank was shut, but the hydraulic internal valve of the tank truck unfortunately was not. The butane vapor that was released was at first like a heavy frozen fog. Five minutes later a flash occurred, igniting six surrounding derricks, five crude oil tanks, part of a small residence, the chassis of the tank truck and some fittings on the adjoining casinghead tanks. All of the tanks involved, other than those containing crude oil, were pressure tanks with adequate relief valves, frangible plug or fusible reliefs. These all functioned as designed, none of the tanks blew up, and the vent fires were extinguished by water shortly afterward. Estimated loss: $50,000 to $100,000.
- LAREDO, TEXAS. OCT. 11, 1945. A tank truck was discharging butane into a storage tank when sparks from an exhaust pipe ignited gas at a leaking connection between the tank and the truck. An explosion occurred which ignited a number of tank trucks near by, as well as the office

and warehouse building, 50 ft away. The driver of the tank truck ran fourteen blocks to telephone the alarm. The office telephone was not available as the office was locked.

When the first fire companies arrived, the fire was burning between the storage tank on a concrete base and the tank truck. Attempts to extinguish the fire with water from a fire department booster tank had almost succeeded when the water tank was exhausted and three severe explosions in rapid succession spread the fire throughout the property as a hose line was being laid to the nearest hydrant 4,000 feet away. The fire department reports that because of the hazardous nature of the product, this plant had been forced to locate on the outskirts of the city where no fire protection was available.

- DETROIT, MICHIGAN. DEC. 30, 1949. Structural failure of a sheet steel water settling tank (15 ft long and 3 ft in diameter) allowed approximately 780 gallons of a butane-propane mixture to escape. In the ensuing explosion and fire, five employees lost their lives, two were injured and a large area of the catalytic cracking plant was set afire. The fire was brought under control in approximately 45 minutes by closing all accessible valves and applying water spray and foam. However, the butane-propane mixture burned for several hours at pipe breaks until diverted by manipulation of valves to a large open field where the gases burned spectacularly but caused no damage.

The water settling tank and other equipment for cleaning the liquefied petroleum gases had been temporarily placed out of service for six hours and were being restored to operating status when the rupture occurred. Investigators were of the opinion that the settling tank was not engineered to meet the demands placed on it under the conditions found in this case, and that the failure was due to unusual stresses and vibrations occurring as the tank was being put back in service.

When valves were opened introducing liquid under high pressure into the tank, water hammer, turbulence at valve restrictions and vibrations set up when the hot liquid discharged into the cooler liquid all may have set up stresses that caused the tank to rupture.

- BAKERSFIELD, CALIFORNIA. JULY 21, 1942. Butane vapors spread over a 5-acre area when an earthquake toppled a butane storage tank from steel supporting girders, thereby breaking all connecting piping. Ignition is believed to have occurred at electrical equipment. After a delay of 45 minutes until 90 high pressure stills were cut off, Bakersfield firemen were permitted to enter the refinery and use hose streams to cool equipment. When leaking butane storage tanks were cooled, the fire went out. Loss was $1,500,000.
- WILMINGTON, CALIFORNIA. JULY 22, 1955. Rupture of a 6-inch pipe carrying butane under 500 psig pressure from the process tower to the cooling tower was believed to have preceded an explosion and fire that damaged all outside processing equipment including towers, pressure vessels, and piping. Congestion of equipment and lack of water spray protection contributed to the $400,000 loss which would have been still greater but for two employees who shut valves at considerable danger to themselves.
- MONTREAL EAST, QUEBEC. JANUARY 8, 1957. During filling of a sphere, butane overflowed into the diked area in which three spheres were located. Some of the butane remained as a liquid, but, unknown to employees, a large vapor cloud also formed that spread over a large area. At approximately 4:15 a.m. the vapor cloud ignited, presumably from a heater in a service station 600 feet away. An intense fire broke out in the diked area around the spheres, and explosions occurred in pump houses and other buildings. One employee was fatally burned. When two of the spheres exploded, debris was thrown over a large area, and several tank fires were started.

The refinery fire brigade with the help of other private brigades and the public fire department brought the fire under control within 24 hours and had completed extinguishment in another 24 hours. Property damage and business interruption totaled $3,164,000.

- WHITING, INDIANA. JANUARY 29, 1957. Three interconnecting large pressure tanks in which propane from an alkylation unit depropanizer was stored were destroyed by a violent explosion that occurred when a valve was opened to relieve pressure above the stored liquid propane. The explosion also occurred in the vent line which discharged into the refinery fuel-gas line. A high concentration of oxygen in the vapor space of the tanks was probably an important factor responsible for this explosion. No employees were killed, but the explosion damaged the alkylation unit, destroyed several small buildings and started several fires at storage tanks, many of which were destroyed before the fire was controlled. The loss was $3,000,000.

- ALMA, MICHIGAN. FEBRUARY 15, 1958. Overpressure, apparently the result of inadequate venting of an "80-type" butane container, caused it to rupture. The tank split at the end as well as longitudinally. The released contents drifted across the refinery yard as a visible fog. The fog ignited when it reached a source of ignition in the powerhouse 350 feet away. A severe fire was caused which killed one man and burned four others.

- ARDMORE, OKLAHOMA. APRIL 15, 1978. A propane mixture was being pumped from a storage tank to a tank car at the loading rack at an oil refinery when a flexible coupling between the pump and the storage tank failed. A workman noticed the vapors spraying out of the coupling, sounded the emergency whistle, and went through the vapors to attempt to close the valves. Before he reached the valves, the escaping vapors had found their way to a boiler house and ignited, blowing the employee into a ditch. The ensuing fire involved several buildings and storage tanks, including one 4,000-barrel tank of gasoline, which burned itself out. Damage was $100,000.

- MCKITTRICK, CALIFORNIA. MAY 28, 1959. A workman was starting to bleed water from a large horizontal LP-gas storage tank when a 1-inch by 3-inch screwed pipe nipple failed at the threads, allowing the LP-gas to escape through the 1-inch opening at about 100 psig. In a little more than a minute, the gas was ignited at a gas-fueled engine about 50 feet away. There were five other tanks nearby. The tank next to the tank with the broken connection was the first to fail. This occurred about ten minutes after the initial fire started. One head and about 15 feet of the shell of this tank traveled 2,250 feet. Within 35 minutes of the initial ignition, all six tanks had failed, some violently, and one with relatively small damage.

Of the six horizontal storage tanks of 10,000- to 26,000-gallon capacity, three contained butane, one propane, and two gasoline. Two of the butane tanks were almost full, but all of the others were less than one-quarter full. Apparently, the top dry shells became hot from flame impingement, softened, bulged, and ruptured. Miraculously, only two employees were injured.

- DETROIT, MICHIGAN. FEBRUARY 24, 1960. Refinery workmen had installed a pump on an LP-gas line leading to butane and propane receivers and had failed to tighten down all of the flange bolts. When the pump was placed in service, the propane leaked from the loose flange joint and was ignited by the open flame of a burnoff torch. Firemen allowed the spectacular blaze to burn under controlled conditions until the LP-gas supply was shut off and the remainder of the gas in the lines was consumed. The employee who started the pump was injured, and damage to the plant was estimated at $133,000.
- BAKERSFIELD, CALIFORNIA. SEPTEMBER 21, 1960. One head blew off a large natural gas blanketed caustic storage tank, presumably from overpressure caused by inadvertently introducing propane from a sour propane scrubber. The gas was ignited apparently by sparks from an electric panel board which was hit by the flying end of the tank. A piece of flying metal or other substance knocked loose a gage fitting and valve on another vessel about 100 feet away. This vessel contained butane which flowed to the ground where it was ignited by the original fire. This burning liquid flowed beneath four 10,000-gallon horizontal butane tanks.

The water spray system in the tank area was inadequate in both water delivery rate and design, and plant operators and the public fire department laid hose lines to cool the tanks. Fire was burning at the relief valves on all four butane tanks by the time the public fire department arrived, but the fires diminished rapidly as cooling water was applied. Under protection of water spray, the broken gage fitting was plugged and the original fire soon burned itself out. After the fire, it was found that one of the butane tanks was severely bulged on the bottom, this in opposition to the theory that vessels are usually self-protected below the liquid level line. Further inspection of the interior of the tank revealed a layer of rust and scale in the bottom of the tank about 3 feet wide and from 3 to 5 inches deep. This had served as an insulating blanket during the fire and allowed the tank plates to heat and bulge. The source of the residues was not known.

- NEAR PAMPA, TEXAS. FEBRUARY 12, 1962. Fire destroyed a major portion of a combination natural gasoline plant and refinery. The fire broke out after a horizontal tank failed, spilling 25,000 gallons of isopentane on the ground. The liquid quickly ignited when vapors reached a nearby heater in a catalytic reforming unit. One employee was injured; damage was $1,250,000.

In the relatively limited area in which the fire occurred, there were approximately 60 storage tanks of various sizes, containing light hydrocarbons from butane to gasoline. Flash points of the materials involved ranged from about -76°F to -45°F. Working pressures of the various tanks were from 30 psig to 200 psig. Fractionating columns, the reforming unit, and other production equipment were also in the small plant area.

A workman was standing near one of the 29 similarly constructed horizontal tanks of 25,000-gallons capacity and 30-psig working pressure, when the metal in the tank failed at the knuckle in one head. The opening extended through about 150° of arc around the lower portion of the head. The isopentane apparently was ignited by a heater in the catalytic reforming unit.

Because of the proximity of the storage tanks and other equipment in the area of the blaze, the fire was considered too dangerous to fight. Responding fire fighters from Pampa and Skellytown confined their efforts to extinguishing grass fires around the plant and protecting the company houses. At the height of the blaze, it was impossible to approach much closer than 200 yards from the burning plant.

All of the tanks were vented, but apparently the vent areas were not enough to protect them in a fire of such intensity. Many of the tanks opened up at the heads, and others collapsed. Three tanks which were in n-butane service with working pressures of 80 psig opened with some force longitudinally along the top. The failure of these tanks knocked two other tanks of the same size, but with 100 and 200 psig working pressures, off their foundations. The latter two higher-pressure tanks did not rupture. In all, 60 tanks and 800,000 gallons of flammable liquids and LP-gas were destroyed. A number of fractionating columns, pumps, and piping also were destroyed, and the catalytic reforming unit was heavily damaged.

- RAS TANURA, SAUDI ARABIA. AUGUST 4, 1962. A propane storage and shipping tank ruptured at the shell-to-roof joint because of over-pressure and caught fire. Both the tank and its contents were a total loss.

The 80,000-bbl, dome-roofed, 49 by 108-ft tank contained 51,000 bbl of propane at a temperature of -41°F.

Two adjacent tanks of the same capacity--one containing 55,000 bbl of propane stored at -41°F and the other, 28,000 bbl of butane at 28°F--caught fire at pilot-operated relief valves on their roofs and were damaged. The propane was pumped out of the one tank and burned in a pit. All but 6,000 bbl of the butane from the other tank was shipped to a tanker at the company's Ras Tanura marine terminal.

One operator died from burns, and the fire hospitalized six other men. The estimated loss was placed at $1,250,000.

During the night preceding the fire, the entire refrigerated LP gas plant except the three storage tanks had been shut down for test and inspection. Pressure began to increase in the tanks soon after the shutdown, in spite of the operator's efforts to relieve it to flare.

The bursting pressure of the propane tank that eventually was destroyed was reached because liquid butane from one of the adjacent tanks leaked into and eventually completely filled the 24-in. vapor blowdown header leading to a vertical knockout drum at the base of a 150-ft flare stack. This shut off the normal path for excess vapor release from all the tanks.

It was later determined that a 6-in., remotely controlled, air-operated block valve in the butane recirculation to the butane storage tank had failed to close completely. This particular valve was designed to be held open by air pressure and closed by springs when the air pressure was released.

Investigation showed that a small accumulation of corrosion products, plus the protective coating of paint on the springs, caused the malfunction. Leakage through this valve was released to the blowdown header when the operating staff opened a manually operated valve in a 4-in. crossover line that had been installed for purging service.

- FAIRFAX, KANSAS. NOVEMBER 29, 1976. The gasoline processing unit at a petroleum refinery was damaged by a butane and propane explosion. It was the third explosion at the plant in recent months.
- EAST CHICAGO, INDIANA. AUGUST 13, 1978. Leakage of isopentane from the relief valve at the top of a storage tank resulted in a vapor cloud which drifted beyond the property line, ignited, and resulted in a fire which heavily damaged three homes. One person suffered minor burns, and several were treated for smoke inhalation.

The leak resulted from overfilling of the tank which occurred because of a malfunction in the tank's pressure gauge. The leak was stopped in 15 minutes, but it took five hours for the vapor to dissipate under the high-humidity, no-wind atmospheric conditions.

- OKLAHOMA. In a refinery, the flexible coupling of a propane supply line failed, resulting in an explosion which injured one person and caused about $200,000 damage.

PIPELINES

- PORT HUDSON, MISSOURI. 1970. Following rupture of an 8-inch propane pipeline, a large vapor cloud formed. This cloud ignited and resulted in an explosion estimated to be equivalent to the destructive force of 50 tons of TNT. There were no fatalities in this case, but 10 persons were injured.

  This particular pipeline had ruptured 12 times in the previous 6-year period, releasing a total of 39,000 barrels of LPG. These ruptures were all due to failure of the longitudinal welds.

- TEXAS. FEBRUARY, 1976. An 8-inch pipeline ruptured in the longitudinal weld, releasing some 5400 barrels of LPG, engulfing an area 3/4-mile long by 1/4-mile wide in flames, burning the occupants of two dwellings severely, and killing 5 people. This pipeline had suffered a series of 14 longitudinal pipe seam failures between January 1968 and February 1976, resulting in 6 fatalities and the loss of 60,000 barrels of LPG.

- CEDARDALE, OKLAHOMA. JANUARY 1976. In an accident at an LPG compressor station, five men were killed and two were burned seriously when natural gas liquids (NGL) were released into a ditch by an error in procedure.

- ROMULUS, MICHIGAN. AUGUST 1975. An 8-inch propane pipeline ruptured due to previous mechanical damage and released some 2400 barrels of propane. The resulting fire burned 9 persons, destroyed 4 houses and damaged three others, and burned 12 vehicles.

- DONNELLSON, IOWA. AUGUST 1978. An 8-inch LPG pipeline ruptured and released 3750 barrels of propane. The vapor cloud which formed ignited and burned 75 acres of cornfields; killed 2 persons and critically burned 3 others while fleeing their homes (1 of the critically burned persons later died); and destroyed a farmhouse and 6 buildings. The cause of the pipeline rupture is not yet known.
- RUFF CREEK, PENNSYLVANIA. JULY 1977. A 12-inch pipeline under 450 psig pressure ruptured and released 1800 gallons of propane which vaporized and settled like a fog over the bottom of a valley. Approximately 2 hours after the rupture, two men in a pickup truck entered the propane cloud, the truck stalled, and the propane gas ignited when an attempt was made to restart the truck. A flash fire approximately 10 yards wide followed a streambed located along the bottom of the valley and burned everything in its path for a distance of 1 mile. As a result of this accident, the 2 persons in the truck were killed, the truck was destroyed, 57 head of cattle were killed, overhead power and telephone lines were destroyed, a hay storage shed containing 450 bales of hay was burned, 1,800 barrels of propane burned, and a meadow and wooded area 1 mile long by 100 yards wide was burned. The National Transportation Safety Board determined that the probable cause of the accident was the failure by stress corrosion cracking of a 12-inch propane pipeline which had been subjected to earth subsidence caused by previous coal mining operations underneath the pipeline.

TANK TRUCKS

- DEER PARK, PENNSYLVANIA. 1959. A pipe on a tank truck containing LPG was damaged, the leaking LPG caught fire, and flames enveloped the rear of the tank. Firefighters cooled a neighboring building but did not cool the tank. After 45 minutes a BLEVE occurred, and the rear head the tank came to rest 900 feet away, after killing 11 people on the way.

- BERLIN, NEW YORK. 1962. Failure of the cargo tank of a tractor-tank semi-trailer unit resulted in the sudden and complete release of about 7,000 gallons of LP-gas. Subsequent ignition, by an unknown source, of the large vapor cloud resulting from vaporization of the liquefied gas culminated in the deaths of 10 persons, injuries to 17 others and property damage to 20 structures and 11 vehicles.
- FLORIDA. 1974. A hose from a 9,000-gallon tanker failed, resulting in a vapor cloud 40 ft high, which exploded. Two warehouses were destroyed, cars were crushed, and windows were broken in a four-block area. No one was injured.
- SCRANTON, NORTH DAKOTA. A full LP-gas transport was connected for unloading into a 20,000-gallon horizontal LP-gas storage tank. The truck driver was on top of the tank, preparing to open the discharge valve, when a small explosion of undetermined cause knocked him to the ground. After a fire at the truck had burned for about 10 minutes, one head of the tank truck failed in a BLEVE, and the tank shell rocketed 300 feet forward. A wave of burning gas moved rapidly in the direction opposite from tank travel about a quarter of a mile, igniting a grain elevator, an oil warehouse, and a pump house for an oil company. The driver was severely burned and subsequently died.
- A fire occurred when LP-gas liquid escaped from a broken 2-inch hose connection underneath a tank truck during delivery to a 500-gallon storage tank. The released LP-gas was probably ignited by gas-fueled kitchen equipment in a nearby mess hall. Approximately 15 minutes after the initial hose failure, the large cargo tank on the truck ruptured near the top of the front head. This failure was apparently due to localized heating, as the top of the head was in direct line with the vent discharge from a safety relief device on an LP-gas saddle tank. Three persons were burned when the tank ruptured. The fire destroyed the LP-gas tank truck and a nearby army truck.

RAILROAD TANK CARS

- PORTLAND, OREGON. 1954. A valve stem of a safety relief valve on a tank car was broken. The vapor cloud from the leaked LPG exploded, causing about $0.9 million damage.
- DALLAS, TEXAS. FEBRUARY 20, 1977. Twelve cars of a Santa Fe train derailed, and the coupler of a piggyback car punctured the head of a jumbo LPG car, releasing LPG and producing a vapor cloud over a large area. Approximately five minutes after the derailment, the vapor cloud ignited, presumably at one of the diesel units, and an extremely violent open-air explosion resulted. The shock was felt for miles. Casualties included only one civilian treated at the scene for minor injuries by fire department ambulance personnel. However, property damage estimates included $650,000 for railroad property and $3,500,000 for other property.
- CUMMING, IOWA. APRIL 29, 1969. A 112A340W car fabricated of A-212 Grade B flange-quality steel was derailed and sustained a long rail burn caused by the tank sliding along one of the rails. The rail burn started at the transition joint between the cylindrical shell and the head at one end of the tank. The tank was rotated by the wrecking crews so that the fittings were upright. Two days after the derailment and two hours after righting the tank, it ruptured violently. The fracture initiated at the rail burn in the transition weld. Examination of the fractured surfaces also showed that a significant dent existed at the fracture origin at the time of rupture.
- DES MOINES, IOWA. SEPTEMBER 1, 1975. Eleven tank cars carrying LPG were included in a derailment which totaled 13 cars. Several LPG cars exploded in rapid succession, and shrapnel was propelled as far as 1,000 feet. Because the area was isolated, the fire department made a formal "no attack" decision, and the tanks were allowed to continue to burn; one tank burned for 10 days. Two firemen received minor burns before the "no attack" decision was made, a passing

motorcyclist on the nearby interstate highway was seriously burned, a fireman who went to assist the motorcyclist received minor burns, and a woman motorist received radiant heat burns through the closed windows of her automobile.

- FERTILE, MINNESOTA. OCTOBER 23, 1975. As a train approached a bulk gasoline and fuel oil storage facility south of Fertile's downtown area, the train derailed, and the coupler on a freight car behind one of four propane tank cars punctured the head of the propane tank. In this case, the puncture propagated more than usual, resulting in a large enough opening so that issuance of the propane propelled the tank. The 30,000-gallon tank was propelled some 800 ft in a relatively straight line, shearing off two utility poles, destroying an automobile, careening off one house and passing through a second house. Several combustible buildings in the path were ignited, and buildings at the bulk petroleum plant were partially or totally knocked down and ignited. These fires threatened three other propane cars and the petroleum storage tanks, but application of cooling water prevented loss of these tanks. Fire damage was limited along the path the tank traveled because the entire contents of the propane tank were ejected and burned before the tank reached the end of its trajectory.

- DECATUR, ILLINOIS. JULY 19, 1974. A 30,000-gallon tank car containing liquefied isobutane was punctured in a humping operation in a railroad yard. Liquid was released, and a larger vapor cloud was produced. The vapor cloud exploded, and the effects were felt 45 miles away, and litter and debris covered a 20-block area. The explosion and ensuing fires caused 7 fatalities and injuries to 152 people. More than 500 freight cars were involved in the fires, and nearly 2,000 homes and commercial structures were damaged. The radius of blast damage was 2-1/2 miles. The total damage has been estimated at over $24 million.

- ONEONTA, NEW YORK. FEBRUARY 12, 1974. A train derailment involving 27 cars, including 7 LP-gas cars, resulted in explosions of 3 of the cars. Large portions of one tank car flew more than 1300 ft; some 50 people were injured, 2 of them critically.

- KINGMAN, ARIZONA. JULY 5, 1973. A propane tank car on a siding caught fire and exploded. The explosion killed 13 people and injured 95 others. The incident occurred at a bulk plant during unloading of a tank car. Leakage occurred in the transfer line, the vapor cloud ignited from an unknown source, and the resulting fire near the top of the car caused the tank car to explode within a few minutes. One-half the rail car tumbled end-over-end down the tracks for 1200 ft, and the other half flattened out in place. The explosion was followed by a flaming mushroom cloud several hundred feet high and 800-1000 ft in diameter. Most of the casualties were firemen who had arrived before the explosion, and who were within 150-200 ft of the tank car.
- CRESCENT CITY, ILLINOIS. JUNE 12, 1970. A train including 56 loaded cars and 46 empties derailed in the business district of Crescent City. The derailment was caused by a hot box on the journal bearing of a hopper car carrying sand. The derailment included a total of 15 cars, 10 of which were LPG tank cars carrying 34,000 gallons of propane each. Two other cars carrying propane remained on the tracks. It is thought that in at least 5 of the cars, the tanks failed by circumferential fracture. In the resulting fires, 6 propane tanks exploded, and the resulting inferno destroyed the business district and many homes. Although there were no fatalities, 64 people received treatment at area hospitals; several were hospitalized as long as two weeks because of the severity of their burns. Unofficial estimates of property damage were in the range of up to $3 million.
- LAUREL, MISSISSIPPI. JANUARY 1969. Four of fifteen tank cars containing LPG, which were involved in a derailment and fire, failed by circumferential crack fracture. These failures took place during a 40-min period following the derailment and resulted in the ends of the tank cars being projected up to 1600 ft from the site of the accident.
- MELDRIM, GEORGIA. 1959. In a derailment of a freight train, two LPG tank cars plunged down an embankment into a crowded picnic area. A coupler from one of the tank cars jammed through the head of the

other, thus releasing large amounts of LPG which subsequently ignited in a fireball explosion. Twenty-three persons died and many others were injured by burns.

- GALLUP, NEW MEXICO. MAY 29, 1981. A train at a refinery was being loaded with LPG when lightning struck the dome of one of the tank cars. The lightning must have ignited escaping LPG, causing fires that lasted several days. The fires burned from the pressure relief valves of six tank cars. No attempt was made to douse the tanks with water, no BLEVE occurred, and no deaths or serious injuries were reported. However, about 300 people living nearby had to be evacuated temporarily. At the time of this writing, the accident was still under investigation by the NTSB.
- WAVERLY, TENNESSEE. FEBRUARY 22, 1978. A 112A400W car fabricated of A-212 Grade B flange-quality steel was derailed, was separated from its trucks, and was wedged between two box cars. The tank car was dented and scraped for approximately one-quarter of its length. Forty hours after derailment and 24 hours after being moved, the car ruptured violently.

  The fracture initiated at the second girth weld from a head. The fracture propagated axially about 16 feet before it arrested and turned into the circumferential direction, producing two tubs that rocketed and a flattened center section. Examination of the fracture surfaces showed that a gouge was located at the origin of the fracture, and that the fracture initiated at the gouged outside surface of the tank.

  Investigation also revealed that during the 40 hours after the accident, the ambient temperature increased from 30° to 51°F at the time of rupture. This change in temperature is believed to have caused the pressure in the tank to increase 50 percent from 55 to 82 psig. The pressure may have even been higher, since a bright sun was out at the time of failure, and the liquid could have been locally hotter than the ambient temperature.

The sequence of events in the failure of the tank car was: (1) denting of the shell during derailment, (2) partial recovery of the dent by pressure in the tank when the denting object was removed, (3) further recovery of the dent when the pressure increased, (4) plastic tensile strain created by the recovery, and (5) cracking of the cold-worked material.

- CRESTVIEW, FLORIDA. APRIL 8, 1979. A 105A300W tank car fabricated of TC 128 B steel was derailed, causing a small leak in a head of the car. Because of a steel jacket over the insulation, it was not known that other damage existed, and the car was put on a siding and allowed to leak. Eight days later it ruptured. The fracture initiated at the circumferential weld between the first two shell rings due to mechanical damage sustained in the derailment. The crack propagated a few feet in each direction circumferentially, and then both fractures turned and propagated longitudinally toward the head. Other than the fact that this accident represents another incident of a delayed failure of a tank car due to mechanical damage, it is also important from the viewpoint that the damage was concealed by the steel jacket after the accident.

SHIPS/BARGES

- TOKYO BAY, JAPAN. NOVEMBER 9, 1974. While underway in Tokyo Bay, the Japanese LPG tanker YUYO MARU No. 10, carrying refrigerated LPG in her main tanks and naptha in her wing tanks and forward reserve cargo oil tank, was struck approximately at a right angle on her starboard bow by the bow of the Liberian cargo vessel PACIFIC ARES. As a result of the collision, the outer plating of the forward reserve cargo oil tank and the #1 starboard wing tank (both containing naptha) was broken. This allowed a large amount of naptha to flow out onto the PACIFIC ARES and onto the water. The naptha caught fire, killing 5 of the YUYO MARU crew members and 28 on the PACIFIC ARES.

Fire-fighting efforts began about an hour after the collision. About two hours later, all external fires aboard the PACIFIC ARES were extinguished. The fire aboard the YUYO MARU was attacked with fire-fighting foam but, in spite of these efforts, the fire continued to spread to more of the naptha tanks. The heat from the fire caused the LPG tanks to vent and reportedly melted one relief valve and gasket and packing materials at joints in several vent and gage lines leading to the LPG tanks, resulting in a series of small fires where the LPG vented from the tanks. Eventually all naptha fires on the YUYO MARU were extinguished; only the LPG venting from the relief valves and heat damaged piping continued to burn. For the most part, these were small, localized fires with an occasional larger flare-up.

Five days after the accident, the decision was made to tow the YUYO MARU out of the day. During the towing operation, naptha was spilled and fire again broke out. Towing was suspended at this time, the ship now being about 23 nautical miles from the shoreline. The ship was subsequently towed further out to sea and was then sunk by the Japanese Defense Agency.

Because this is the largest and most dramatic incident involving fire aboard an LPG ship, it has often been cited as an example of the hazards posed by such ships. However, the tanks that contained the naptha onboard the YUYO MARU cannot be used to carry flammable liquids when the ship is in a U.S. port. Since the collision did not damage any of the LPG cargo tanks, it is very likely that no fire would have occurred if the naptha tanks had been empty or were filled with ballast water. The fire was essentially a naptha fire. The only part of the LPG cargo that was involved was the portion that vented and fed the small fires around the cargo tank hatch areas; the naptha fire never breached the integrity of the cargo tanks but only damaged relief valves, packing, gaskets, etc. on the cargo piping system. And, as a final note, the relative integrity and stability of the LPG ship was demonstrated by the fact that the Japanese Defense Agency was able to sink the ship only with great difficulty using shells, bombs, and torpedoes.

- BORDEAUX, FRANCE. MARCH 23, 1967. As LAVOISIER, A French LPG ship, was off-loading butane gas into a dockside storage tank a leak occurred. Officials said a fracture in a safety valve may have caused the leak. Firemen dispersed hundreds of tons of butane gas which leaked into the harbor. Authorities, fearing an explosion, instituted strict security precautions as firemen attempted to stop the huge white cloud drifting over the city. The vessel had loaded 3,000 tons of butane at Argentina.

- DONGES, FRANCE. OCTOBER 10, 1978. A deck pump on DANIAN GAS caught fire while discharging propane. The vessel had discharged about 900 tons of propane at the time of the fire; 14,000 tons of propane were on board. A refinery fireman extinguished the fire in 30 minutes. For safety's sake, the vessel was towed to sea. Void spaces were inerted, the defective pump repaired, and the vessel returned to port and completed discharging.

- HOUSTON, TEXAS. JUNE 16, 1976. An accident resulting in LPG cargo leakage occurred when the tank barge CC114, CG-006304, broke away from its moorings in Sims Bayou, Houston, Texas. The barge subsequently struck an unknown object, shearing off two aft port cargo tank fittings, with subsequent partial sinking and collision with the SS ARCH ENTERPRISE, O.N. 276911, moored at ARCO Dock #3, Houston Ship Channel.

  From the marine casualty report, the proximate cause of the casualty was an abnormally fast rate of flow of Sims Bayou caused by torrential rains during the previous 24-hour period. This resulted in the CC114 breaking away from its morrings at Fleeting Service, Inc., striking an unknown object and shearing off two fittings on the top of the port aft tank, allowing butane to be sprayed into the atmosphere. Also, at this time the stern rake and aft hopper space were flooded. The barge then entered the Houston Ship Channel, where it floated downstream and struck the ARCO ENTERPRISE, loading at ARCO Dock #13, on the port side in the vicinity of the #5 cargo tank. A Port of Houston fire boat, a U.S. Coast Guard 30-foot utility boat, and commercial towboats arrived at the scene and took the barge under control about two hours after it had broken away from its moorings.

The reason the barge partially sank is unknown. Small holes were found in the bottom of the stern rake after the casualty. These may have occurred during the sequence of events after entering the Houston Ship Channel or during the off-loading operation. The most probable reason the barge partially sank was that when the barge struck the object which sheared the cargo tank fittings, the stern rate was forced under the surface, which allowed water to enter the aft hopper space and stern rake vents, further sinking the aft end of the barge.

The fittings on the cargo tank which were sheared off were a 1-1/2 inch ullage pipe which emitted liquid and gaseous butane and a 2 inch nipple on a butterworth plate which emitted gaseous butane. The barge was reportedly loaded with 1250 long tons (approximately 13,500 bbls) of butane. No estimate is given in the report of the amount of butane released.

Repairs to the <u>CC114</u> were made. There was no apparent damage to the SS ARCO ENTERPRISE.

- GOOD HOPE, LOUISIANA. AUGUST 30, 1979. The LPG barge PANAMA CITY was docked on the lower Mississippi River near Good Hope. It had been loaded with about 283,500 gal of butane in its six cylindrical pressure vessel cargo tanks. No cargo was being transferred when it was struck by the Peruvian freighter INCA TUPAC YUPANQUI, which had lost its steering. The collision severed the barge into two pieces and ruptured at least one of the cargo tanks. The LPG vapor escaping from the ruptured tank created a vapor cloud that engulfed the freighter. This vapor cloud was ignited almost immediately by an unknown source. A fireball formed which was hundreds of feet high and lasted less than one minute. The cloud did not detonate and none of the tanks underwent a BLEVE. The fireball ignited combustibles on the dock, the shoreline, the freighter, and a towboat, and burned several people in the immediate vicinity. The forward half of the barge hull and the most severely damaged tank both quickly sank. The aft portion of the hull, containing three tanks, and two tanks from the forward half all floated downstream, with all tanks releasing burning LPG from cracks, broken pipes, and/or relief valves. The two separate tanks and the aft portion of the barge were beached at various locations downstream of the dock. The fires were allowed to burn out over the next 24 hours, during which time the Coast Guard closed that section of the river, and some local residents were evacuated. Once beached, the burning tanks did no further damage.

A total of 34 people were hospitalized for burns; 9 died. Three others died from drowning. Ten of the victims were from the freighter; two were from a towboat that was standing by to move the barge. All of the people that were injured were in the immediate vicinity (i.e., on the freighter, barge, dock, or towboat). No one from the general public was injured.

STORAGE

- FEYZIN, FRANCE. 1966. A leak occurred from a 528,340-gallon storage vessel containing propane at a pressure of about 125 psig. The leak ignited, and a fierce fire burned underneath the sphere. The firefighters were advised to use water for cooling neighboring vessels. It was believed that the relief valve would protect the vessel which was on fire. Indeed, the relief valve soon lifted, but after 1-1/2 hours, the vessel burst, and a wave of burning propane wiped out the firefighters. The fire then spread unchecked.
- MICHIGAN. Overpressure in a 21,000 gallon storage tank at a refinery caused failure of the tank. The vapor cloud eventually ignited 350 ft from the source, killing one person and causing $1 million damage.
- LOUISIANA. 1975. Overpressure caused a split in a 6-in. underground line from a butane storage well, releasing about 600,000 lbs of butane. The resulting vapor cloud was estimated to have been 30 to 40 ft high and to have extended about 1 mile downwind. No ignition occurred.
- MARCUS HOOK, PENNSYLVANIA. JANUARY 1978. A 10.6 million-gallon butane storage cavern at a refinery in Marcus Hook was overfilled despite three monitoring methods and a safety relief valve. The

cavern was excavated in granite more than 300 feet underground. The main device to measure the butane level in the cavern, a manometer, malfunctioned and indicated that the cavern contained only 2.6 million gallons of butane. A vapor pressure gauge at the wellhead also malfunctioned, and company bookkeeping records, showing how much butane was pumped into and out of the cavern, were very inaccurate. The safety relief valve, set at 175 pounds per square inch, never operated because the cavern pressure never reached that level. The escaping butane caused fires that destroyed five houses only 100 feet from the outer wall of the cavern.

- MISSISSIPPI. 1974. Overfilling of a salt dome storage well with butane created a vapor cloud 1.25 miles in diameter. Two explosions occurred. The second one was 800 to 1,000 ft above grade and was more severe. Houses were damaged up to 900 ft away and glass breakage occurred up to 7 miles away. Twenty-four people were injured.
- NEW BRUNSWICK, CANADA. JANUARY 1973. Leaking gas from an underground LPG storage site seeped through the ground into a nearby cave. When a light switch was thrown, an explosion occurred that seriously injured two boys.
- BARBER'S HILL, TEXAS. SEPTEMBER 1980. A pressure loss was observed in the giant Barber's Hill salt dome LPG storage cavern. Some gas had leaked to the surface and forced temporary evacuation of 68 homes located above the cavern. Only one minor explosion has been reported with no injuries. The evacuated people are now suing the owner of the LPG storage facility in a class action suit.
- TEWKSBURY, MASSACHUSETTS, FEBRUARY 9, 1972. The chain of events started about 10:30 p.m. February 9, 1972 when a Lowell Gas Company employee backed a tank truck, T-24, loaded with propane that later exploded, into position to unload. The tanker apparently struck a pipe used to fill two 60,000 gallon LPG horizontal storage tanks. Whether the pipe was broken off or cracked is not known, but this pipe was connected to the fill pipe that a second tanker, T-25,

was using to unload its 8,500 gallons of liquid propane into the same storage tanks at a rate of about 68 gallons per minute. Although it was not known at the time, the investigation showed that the gas was flowing from the unloading tanker and was escaping from the damaged pipe.

As the two drivers attempted to shut off the flow of gas, the vapors spread against the wind, estimated at 10 to 15 mph, to vaporizers about 50 feet away, where they were ignited. The vaporizers were not then being used, but the investigators concluded that a pilot flame was the ignition source.

Liquid propane from Truck T-25 was burning at the break at a rate of probably nearly 100 gpm in the confined area between the ends of the storage tanks and the cargo tank of Truck T-24 and contacting the cargo tank extensively on the exposed side and rear head. Combustible cab materials on both trucks were ignited almost immediately.

An ambulance and pumper (Engine 5) were dispatched by the Tewksbury Fire Department on first alarm. Engine 5 had its normal company manning of a captain and four fire fighters. After a delay of a few seconds for the ambulance to clear with the badly burned gas company employee, Engine 5 parked at the plant roadway alongside the Propane-Air Mix Building at 10:42 p.m. The captain immediately radioed a second alarm and asked that hose lines be laid to supply Engine 5.

A deluge gun was set up just inside the gateway, and by the time two 2-1/2 inch lines were hooked up to the gun, another unit (Engine 4) had arrived at the scene and had completed an 800 ft lay of two 2-1/2 inch supply lines to Engine 5 from the town hydrant near the plant entrance driveway. At 10:44 p.m., this deluge gun was in operation, playing water on the ends of the storage tanks and T-24's cargo tank in the fire area at a rate of 400 to 500 gpm. The stream was being directed from side to side to alternately cover the storage tank ends and the cargo tank. The men on Engine 4 took its deluge gun to a position east of the storage tanks and waited for lines to be run in from the next closest hydrant by other units responding to the alarm.

At about 10:46 p.m., one of the two relief valves on T-24 operated, sending a burning torch about 60 feet into the air. At this time, the men working the operating deluge gun were directed to leave the gun fixed in position and take cover behind the corner of the Propane-Air Mix Building--coming out periodically to reposition, if necessary. The relief valve remained open steadily. The second relief valve on this cargo tank did not operate nor did the relief valves on the storage tanks.

About 10:48 p.m., a stream from a 2-1/2 inch hand line was directed from Engine 1 of the Billerica Fire Department to T-25 and attempting to reach T-24. However, the initial hose lay was too short to reach T-24 effectively. Water for the second deluge gun had not yet been received.

At 10:55 p.m., the cargo tank on T-24 ruptured violently and came apart in two sections. Of the 6,500 gallons of propane in it at this time (about 2,000 gallons having been discharged as vapor through the relief valve), about 3,600 gallons flash-vaporized, mixed with air and was ignited by the fire--creating a large ball of fire. The remaining cold propane was atomized and was flung in all directions in burning and unburned form.

The rear portion of the tank flattened out, and its contents spewed laterally in a nearly 360 degree arc. As a result, the men at the second deluge gun, about 150 feet away, and at the hand line, about 120 feet away, were knocked around and burned. Ironically, at this time, water reached the gun.

The front portion of the T-24 tank remained in cylindrical form and, still attached to the tractor, was propelled forward, jack-knifing the rig. The combination moved along the ground and impacted a vaporizer 30 feet away. At this point, a fifth wheel pin sheared, separating the tractor and front portion of the cargo tank. The tank portion became airborne and sailed through the air into the wooded area. After shearing off three 8-inch diameter trees several feet off the ground, it came to rest about 300 feet from its original position.

The tank, spewing its burning contents, passed almost directly over the men at the operating deluge gun. They were tossed about and burned. At 10:56 p.m., a third alarm was called. At this time, other piping in the area was broken, resulting in propane and natural gas fires at the propane process pump area alongside the northernmost storage tank; in the propane vaporizer area; at the insulation in wall and roof panels in Propane Plant structures; and on the roof of the LNG Plant Boil-off Compressor Building and the redwood portions of the Cooling Tower. Propane from T-25 continued to feed the fire.

Burning propane vapor and liquid ignited hose, tires and other combustibles on Engines 4 and 5 and on the two nearer empty propane transports parked in the Plant Roadway. Grass and fir tree needles and small branches were ignited for a radius of about 400 feet from the original position of T-24.

The shock and pressure wave and subsequent vacuum phase popped wall and roof panels in the Propane-Air Mixing Building, cracked concrete block walls at the mortar joints and blew out windows and doors in the Boil-off Compressor Building and LNG Plant Control Room. Burning liquid propane entered these openings and started small fires in Class A combustibles inside these buildings.

Between 11:00 p.m. and 11:15 p.m., as more help arrived, cooling water was reestablished on the storage tanks, propane vaporizer area and T-25 and the numerous structure and grass fires were extinguished. Fifteen fire companies from fourteen municipalities and an Air Force Base were ultimately required to handle the fire.

At about 11:30 p.m., the excess-flow check valve in the liquid withdrawal connection on T-25's cargo tank closed when the rubber unloading hose burned off--thus shutting off the major fuel source. Shortly after this, gas company employees closed valves at the top of the storage tanks and controlled broken natural gas lines. By midnight, the situation was under control, although another couple of hours was needed to mop up.

- NEWARK, NEW JERSEY. JULY 7, 1951. A severe fire followed by a number of explosions of propane tanks occurred at an import terminal. The incident occurred while the NATALIE WARREN was transferring cargo at the loading dock. Evidence indicates that the fire was caused by a break in a liquid pipe or fitting near the ground in a group of 70 pressure storage cylinders of 30,000-gallon capacity. A minor explosion occurred three minutes after the leak was first discovered, which was followed by a violent explosion 10 minutes later. The NATALIE WARREN's transfer lines were disconnected and the vessel moved to a safe location. For two hours, tanks exploded at 2- to 3-minute intervals. All 70 tanks in the one bank were demolished, and three tanks in another bank of 30 located 350 feet away were also involved, although no explosions occurred in the second bank of tanks.

  The tanks in each bank were arranged in two rows about 15 feet apart, with a clearance of about 15 feet between rows and a separation of about 5 feet between tanks. The tanks (70 feet in length and 10 feet in internal diameter) were mounted on conventional concrete cradles.

  One of the exploding LPG tanks landed on a filling station over 1/2 mile away. Another broke a water main, depriving the area of water until fire boats arrived and pumped water into the mains. Firemen cooled the 30 tanks located in the second bank of storage tanks with fire hoses, thereby limiting damage to these tanks.

  The first alarm was sounded at 1:10 p.m. on July 7, and fire units were still at the scene four days later. The total loss at the facility was estimated at $1.05 million.

- ANTOFAGASTA, CHILE. JANUARY 13, 1965. Four 800-gallon propane tanks on skids and one hundred 72-pound containers of propane were part of a general cargo aboard a freighter being unloaded at dockside. Unloading had begun, and three of the propane tanks had been safely lowered from the ship to the dock. As the fourth was being swung over the side, the 1/2-inch diameter steel cable sling, which was attached to two steel eyes on the tank, broke. The tank fell on top of a railroad car and then to the edge of the pier against the open hatch of a hold in the ship. The impact tore a small hole in the tank and allowed propane to escape. The vaporizing propane began to envelope the ship as the approximately 60 dock workers and crewmen jumped into the sea or fled the area. Within two minutes after the accident, the vapor cloud ignited, probably at a stove in the ship's galley. Seven ship's officers, who stayed aboard, were killed.

The ensuing fire in the ship was so intense that it was five hours before a cable could be attached to the ship to tow it into the bay, where it burned for 27 hours before sinking.

- BARGEDDIE, ENGLAND. In a fire at a propane filling plant, the propane cylinders, which were of 29 pounds, 42 pounds and 104 pounds capacity, were fitted with pressure relief valves. Nevertheless, firefighting was hampered by explosions and debris projected by exploding cylinders. The subsequent investigation showed that many complete cylinders were projected (probably by BLEVEs) up to 100 ft, and cylinder fragments were recovered at distances in excess of 330 ft. A comparison of the debris distribution patterns from the two major incidents showed that, although the pressure relief valves fitted to the cylinders did not have the capacity to prevent explosion, they had sufficient capacity to markedly reduce the severity of the explosions.
- MITCHAN, ENGLAND. Fire swept through a storage depot and involved several hundred full and partly full cylinders of LPG which ranged in size from 2 pounds to 100 pounds capacity. Many cylinders exploded violently, and debris, which included full and empty cylinders, as well as cylinder fragments, was projected up to 1970 ft over the surrounding residential and industrial area and caused considerable damage. There were times during the progress of the fire when the danger from explosion and debris was such that temporary withdrawal of the fire services was necessary. During the subsequent examinations of the scene, it was noted that the least damaged of the cylinders remaining on the site had been fitted with pressure relief devices.

- HARRIN, ILLINOIS. While an LPG tank truck was parked at a bottled gas dealer's property, a leak in the tank filling line permitted LP-gas to spread through a residential area. It eventually ignited explosively in a dwelling 175 feet from the leak. The dwelling was destroyed by the explosion, and three other dwellings were ignited and destroyed when flames flashed back through the vapor cloud. Nine persons standing around watching the leaking LPG were killed in the explosion and fire.

CONSUMER USE

- CALIFORNIA. 1955 In a gasoline plant, a 6-in. line (at 500 psi) carrying liquid butane ruptured, causing an explosion and fire which destroyed all equipment in the vicinity. Estimated damage was about $1 million.
- LOUISIANA. 1965. In a chemical processing plant, liquid propane was accidentally vented into a flare line, which ruptured in the ensuing explosion, injuring 12 workers. Damage was about $6.4 million.
- CALIFORNIA. 1948. The bottom connection on a 4300-gallon butane storage tank failed, creating a 1/4-mile diameter cloud. Five people were killed.
- QUEBEC, CANADA. 1957. Overfilling of a storage sphere with butane created a 1200-ft diameter vapor cloud. The storage sphere subsequently ruptured, causing explosion damage of about $7 million. One person was killed.
- MONTANA. 1972. A butane leak from a 6-in. flange was ignited at a furnace 100 ft away. The resulting fire caused the vessel to rupture in a BLEVE, killing one person and causing $3.3 million damage.
- ILLINOIS. 1951. A bottom discharge pipe on a butane tank ruptured due to freezing of collected water, resulting in the release of about 9,500 gallons of butane. No ignition occurred.

- The owner of a trailer had been using his oven when he ran out of LP-gas. Later in the day, he installed a full tank of gas but had apparently forgotten to turn off the oven valve. Gas leaking from the oven was eventually ignited by a spark from a refrigerator motor in the trailer, and the explosion ripped out one side of the trailer and burned the occupant.

- Instead of calling on trained and qualified men to make an LP-gas installation in his home, the homeowner decided to do the work himself. Gas escaping from a loose connection in the kitchen ignited explosively on reaching the coal-burning furnace in the basement. Three of the four occupants asleep in the second story bedrooms were suffocated in the ensuing fire.

- In making preparations for moving from the house the following day, the occupant disconnected a stove in the basement and attempted to seal the end of the tubing from the LP-gas supply by mashing it against the basement wall with a hammer. This "seal" did not hold, however, and LP-gas leaked from the tubing all afternoon and part of the night, until the gas was ignited by a spark from an electric pump or the flame of a water heater. The resulting explosion severely injured a family of five who were in the dwelling at the time and blew out the outside walls of the house.

- While butane was being transferred from a delivery truck to a tank outside a 35-family, 3-story wooden apartment building, a failure of the hose or connections at the tank allowed gas under pressure to escape. The gas settled in the basement or in a kitchen, where it found an unknown source of ignition. The explosion and ensuing fire demolished the building, injuring five persons.

- UTAH, 1956. Fifteen people in a 1-story concrete block building housing a restaurant were killed when an LP-gas explosion that originated in the basement demolished the building. On the previous day, a changeover had been made at the restaurant from LP-gas to natural gas. After completing the natural gas installation, a workman shut off the LP-gas supply by closing a valve at the 1,000-gallon outdoor tank. He then cut the LP-gas supply line in the basement of the restaurant, neglecting to cap or plug the line.

APPENDIX I

PRELIMINARY ANALYSIS OF LPG POOL FIRE RADIATION DATA

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PRELIMINARY ANALYSIS OF LPG POOL FIRE RADIATION DATA

INTRODUCTION

The DOE test program on liquefied gaseous fuels (LGF) includes studies on radiation fluxes from both liquefied natural gas (LNG) and liquefied petroleum gas (LPG). This report deals primarily with LPG, and is specifically aimed at beginning analysis of a portion of the fire radiation data obtained at the Naval Weapons Center, China Lake, California. During late August and early September, 1977, eight LPG burn tests were run. In four tests, ignition occurred immediately following the start of the spill, so that the fire burned primarily as a pool fire. In the remaining four tests, a vapor cloud was allowed to form before ignition, so the fire was primarily a vapor cloud fire. The spill rates and total spill quantity were about the same for all tests. This summary report covers only the pool fire tests.

SUMMARY OF TEST RESULTS

Table I.1 lists the basic test conditions for each test. The spill quantity is the total amount spilled during the test and the spill duration is the time the flow valve was open. Some small quantity of LPG remained in the discharge line after the spill valve was closed, and that LPG drained out after valve closure. The fuel was directed downward onto the surface of the spill pond, where it spread and vaporized. Deep penetration of the water by LPG was prevented by a splash plate mounted just below the water surface.

Figure I.1 shows the flame dimensions of the pool fire in Test LPG-1 as measured from motion pictures of the fire. The measured wind speed was essentially zero, although some slight flame tilting early in the test indicates a very low wind speed at the start. The fire behavior for Test LPG-1 is not typical of most pool fire tests. As viewed from the crosswind camera, the flame diameter remained approximately constant until the fuel flow was finished. Then, both the flame diameter and the flame height increased

TABLE I.1. China Lake LPG Pool Fire Test Conditions

Test	Date	Time	Type	Quant m^3	Spill Duration (sec)	Air Temp (deg C)	Rel Humid (%)	Wind (Kts)	Wind Dir (deg)
LPG-1	23 Aug 77	1348	Pool	5.3	65	39	23	0	---
LPG-3	25 Aug 77	1320	Pool	4.8	82	36	30	2	45
LPG-4	29 Aug 77	0956	Pool	5.11	57	33	35	1	340
LPG-5	30 Aug 77	1031	Pool	5.5	57	33	38	5	120

for a short time before the fire died out. The upwind view of the flame showed an increasing flame diameter during much of the spill, but film was exhausted before the large increase in flame height was reached. The behavior points out some of the problems in analyzing the data. One is that the fuel burning rate is not well known. Fuel is discharged at a more or less constant rate during a period of about one minute for these LPG tests. In the pool fire test ignition is immediately after start of the spill, and the burning rate increases to a quasi-steady-state rate. When the fuel is turned off, burning continues, and the flame size may increase. Figure I.2, which shows radiant fluxes for LPG-1, shows that the radiant flux increases after fuel is turned off. The increase in flux corresponds to the time the size of the flame increases, as can be seen by comparing Figures I.1 and I.2. In order to calculate the fuel burning rate per unit of flame base area, both the base area and the burning rate must be known. Figure I.1 shows that even for nearly calm tests, the flame base may be irregular in shape, so that finding the base area may be difficult. Determining the burning rate of fuel may be even more difficult. In LPG-1, for example, 5.3 m^3 of LPG was discharged in 65 sec. However, the fire burned for more than 20 sec following fuel shutoff, and the fire was larger during most of the time following fuel shutoff than it was before fuel shutoff. Although the fire size did not increase after fuel shutoff in other LPG tests, the additional burning did occur.

Table I.2 shows the burning rate for each of the four pool fire tests calculated on two assumptions of elapsed time: 1) the fuel burned as fast as it was discharged, and 2) the burning occurred uniformly over the period from

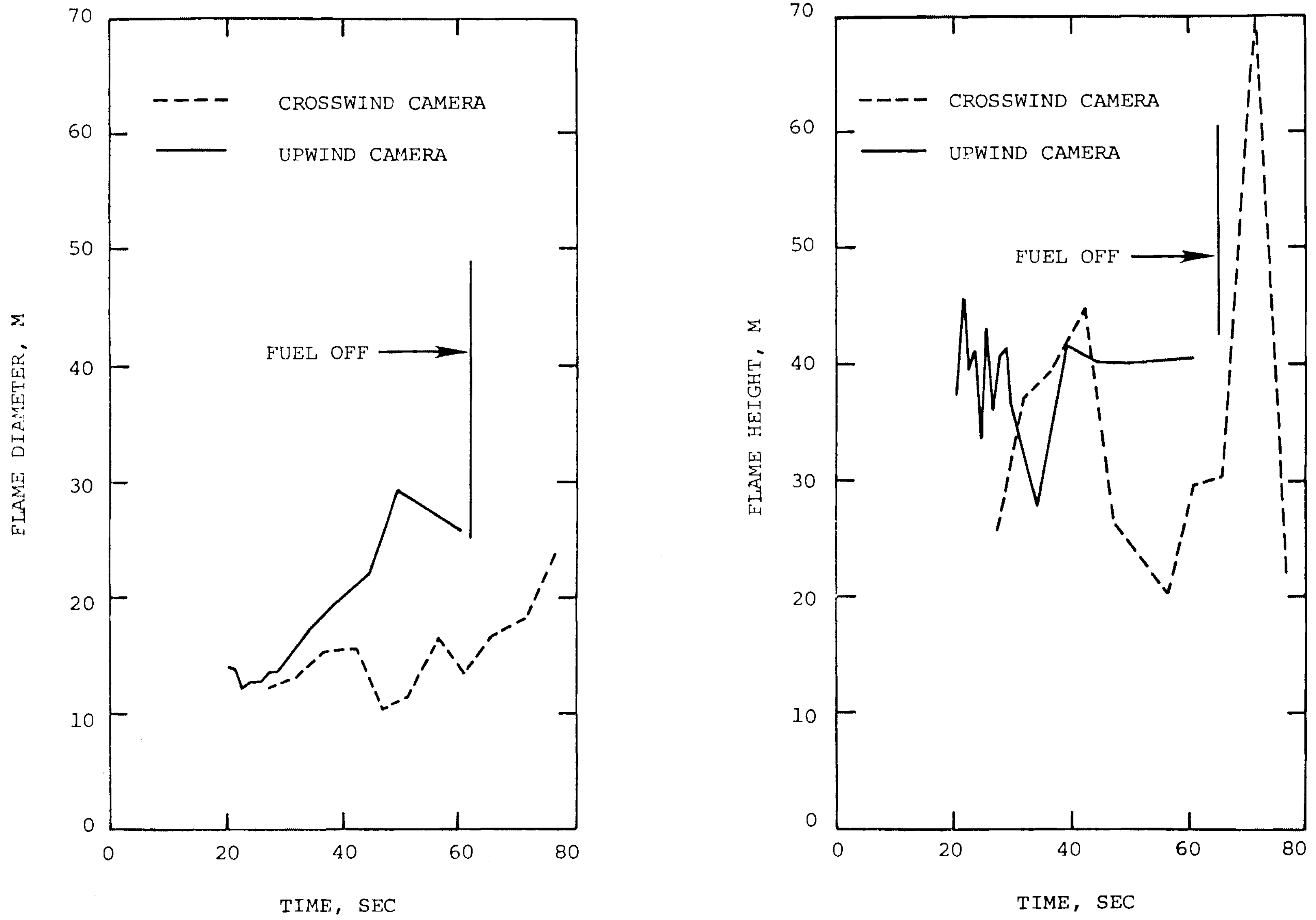

FIGURE I.1. Flame Dimensions for Test LPG-1

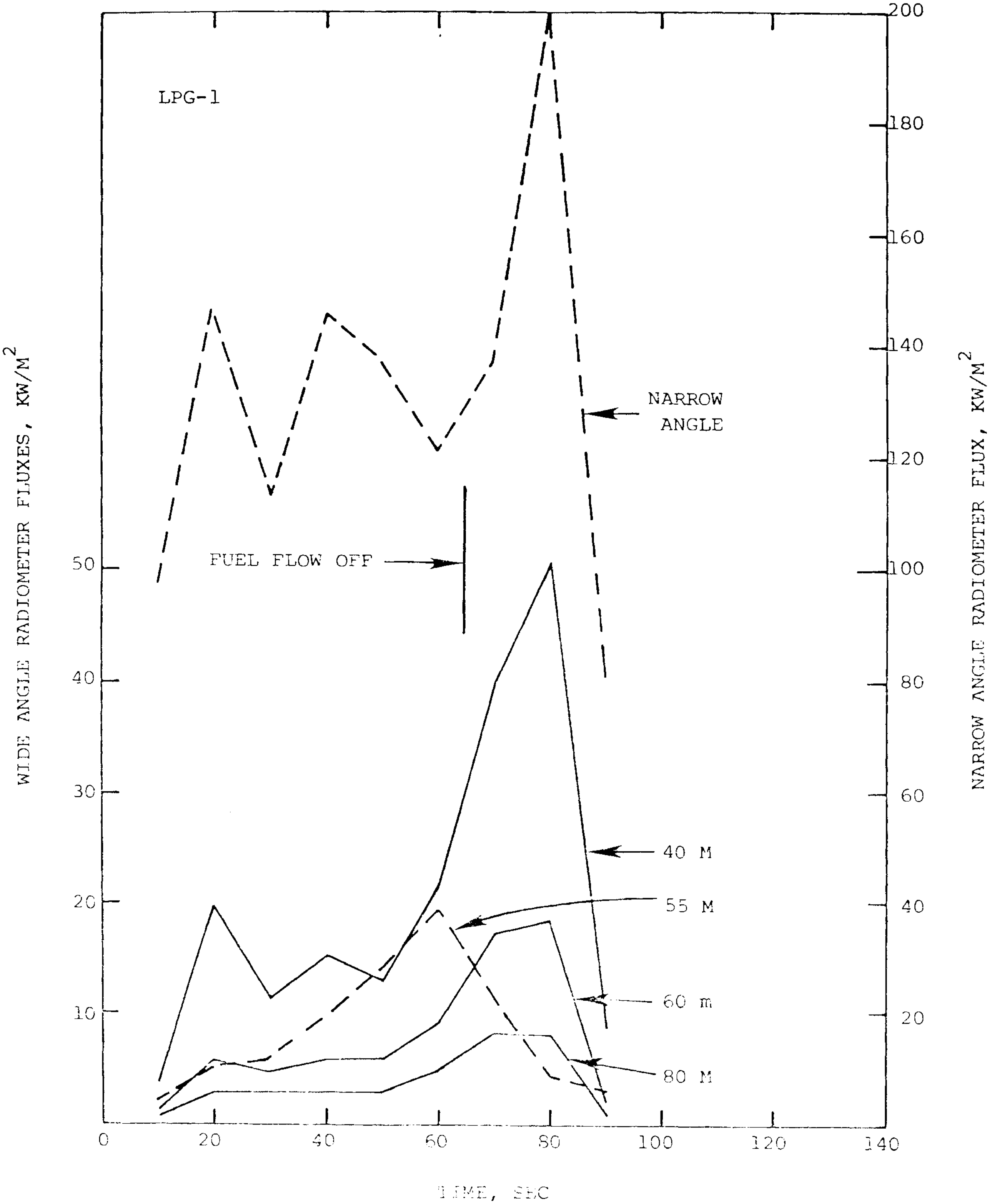

FIGURE I.2. Radiant Fluxes Measured by Wide Angle and Narrow Angle Radiometers for Test LPG-1

TABLE I.2. Flame Sizes and Angles for LPG Pool Fire Tests

Test	Time Interval (sec)	Burn Rate (kg/m^2-sec)	Flame Diam (m)	Flame Length		Flame Angle	
				Meas (m)	Calc (m)	Meas (deg)	Calc (deg)
LPG-1	0-65	0.30					
	12-88	0.26			45(a)		
	20-40		14	36		0	0
LPG-3	0-82	0.18					
	10-120	0.13			38(a)		
	30-80		16	41		13	10
LPG-4	0-57	0.35					
	8-88	0.25			43(a)		
	20-80		14	51		4	4
LPG-5	0-57	0.22					
	12-80	0.20			47(a)		
	20-60		18	47		27	29

(a) Calculated from average burning rate of 0.21 kg/m^2-sec and calculated flame diameter for the time interval shown.

ignition until the fire died out, as judged from radiometer data. The burning rate calculated from spill time is obviously too high. Burning rates calculated from the duration of the radiometer curves are more reliable, although the average thus calculated is somewhat lower than the true instantaneous value at any time during a test.

The burning rates are shown as a mass flux, and the area for the flame base is taken from motion picture records. The flame base diameters cannot be measured more accurately than a meter or two. Assuming the diameter to be 15 m, a 2-m error in base diameter corresponds to about 25 percent error in area, and therefore in burning rate. The average burning rate for all tests based on the active burning period was 0.21 kg/m^2-sec. Steady-state burning rates for propane burning in precooled test pits on land are about 0.11 kg/m^2-sec (Johnson et al. 1980), which is about half the average value measured for the tests on water. The difference corresponds to an effective boiling heat transfer coefficient of about 0.8 kW/m^2-°K based on the superficial pool area.

Flame lengths were calculated from the Thomas (1963) equation to compare with those measured from the motion pictures. Table I.2 lists those results for the four pool fires. Measured flame lengths are shown for each of the fires, as recorded on motion picture film. Flame lengths are measured as the average of the length of the flame from the base to the top of the continuous portion of the flame. If a portion of the flame breaks away, it was not taken as part of the flame in determining the lengths. Such parts of the flame are usually rather small by comparison with the main body of the flame.

Flame bending angles were measured utilizing all the motion picture records and the recorded wind direction. Corrections were applied to the angles measured directly from the pictures to account for the fact that the cameras did not always view directly perpendicular to the wind direction. Flame angles were calculated from the correlation of Welker and Sliepcevich (1970) for comparison. Table I.2 lists both the measured and calculated flame angles. There is a very close agreement. The calculated angles were based on the measured flame diameter and average measured wind speed.

Table I.3 lists the radiant fluxes measured during the tests. The measured fluxes are corrected for the effects of radiometer window transmissivity but not for atmospheric transmissivity. Corrections for window transmissivity were based on the spectral emission curves for LNG for ealier tests at China Lake (Raj et al. 1979). Radiant fluxes were calculated from the flame model developed from confined pool fires burning on land (Welker and Cavin 1981). The only input from the tests involving spills on water used in determining the calculated fluxes was the average burning rate, wind speed, and wind direction. The flame length, flame diameter, flame angle, and surface flux were all taken from correlations or models. The surface flux was

TABLE I.3. Comparison of Calculated and Measured Radiant Fluxes

Test	Time Interval (sec)	Calc Flame Diam (m)	Calc Flame Length (m)	Narrow Angle Rad[a] (kW/m^2)	Wide Angle Radiation			
					Rad 1 at 40 m (kW/m^2)	Rad 2 at 55 m (kW/m^2)	Rad 3 at 60 m (kW/m^2)	Rad 4 at 80 m (kW/m^2)
LPG-1	12-88	16	45	158	16.5	9.8	8.8	5.3
	20-40	(14)[a]	(36)	(136)	(15.5)	(7.0)	(5.4)	(3.0)
LPG-3	10-120	13	38	158	12.4	8.3	6.3	3.8
	30-80	(16)	(41)	(128)	(13.2)	(9.2)	(3.4)	(2.1)
LPG-4	8-88	15	43	158	14.9	9.4	7.9	4.8
	20-80	(14)	(51)	(158)	(25.7)	(7.1)	(7.0)	(3.0)
LPG-5	12-80	17	47	158	26.4	11.6	14.0	7.9
	20-60	(18)	(47)	(151)	(39.3)	(9.0)	(9.7)	(4.6)

158 kW/m^2 for the calculations, a value that is equal to the highest measured narrow angle flux and less than 25 percent higher than the lowest measured narrow angle flux.

Calculated fluxes are included in Table I.3 for comparison with the measured radiant fluxes. On the average, the calculated fluxes are about 30 percent higher than the measured fluxes. If the average surface flux for the calculations was about 120 kW/m^2, the average of the calculated and measured fluxes would be about equal. However, there are some anomalies in the correlation that make that comparison suspicious. For example, the measured narrow angle fluxes are all greater than 120 kW/m^2 and the average narrow angle flux is 143 kW/m^2. Thus, the difference between calculated and measured radiant fluxes is not entirely attributable to differences in surface flux. In addition, the calculated flux for radiometers near the fire (40 m) are generally lower than the measured fluxes, while at farther distances (60 and 80 m) the calculated fluxes are greater than the measured fluxes. The slope of the line drawn through a logarithmic plot of the measured flux data versus distance is much steeper than expected. In particular, the data points measured at the 40-m location show much higher fluxes than expected.

There might be some atmospheric absorption of radiation that would cause some of the behavior noted, but the radiometers are so close together that it is not likely. In addition, the narrow angle radiometers are located at the 60-m position, which minimizes apparent differences in measured fluxes caused by atmospheric absorption.

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16. Abstract
The purpose of this project was to develop and provide information on R&D needs in LPG safety and environmental control for use in future planning and implementation of a LPG safety R&D. This report includes a general review of LPG characteristics, hazards and risks followed by discussion of LPG production, transportation, storage and consumer use. Two sections of the report summarize the extent of available knowledge and capabilities that minimize the consequences of LPG hazards. The first of these reviews current knowledge of phenomena involved in LPG releases; the second summarizes the state-of-the-art in LPG release prevention and control. The last section documents the conclusions resulting from PNL's assessment. These conclusions are presented in the form of specific recommendations for R&D projects that address gaps in current knowledge of LPG release behavior and needed improvements in LPG release prevention and control

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