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Development of a Small-Scale Natural Gas Liquefier

Final Report

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1. Executive Summary

Gas Technology Institute (GTI), with support from the U.S. Department of Energy (DOE) and Brookhaven National Lab, developed and tested a pre-commercial small-scale natural gas liquefaction system. This cryogenic technology has potential use in a variety of applications, ranging from producing fuel for LNG vehicles, remote gas recovery, biogas cleanup and recovery, natural gas peakshaving, and several other specialty natural gas and low-temperature refrigeration markets.

This report discusses the design, development, testing, and economics of a pre-commercial liquefaction unit sized to produce 4 m³/day (1,000 gallons/day, gpd). Figure 1 shows the pre-commercial unit prior to testing at GTI's facilities.

Figure 1: 1,000 Gallon/Day Liquefaction System

The GTI natural gas liquefier system uses a mixture of refrigerants operating in a single, simple refrigerant loop. The unit was skid packaged to be transportable to allow for rapid deployment and minimum installation costs.

The core value propositions for small-scale liquefaction systems include:

- Avoiding over-the-road costs for LNG trucking by placing production at the point of demand.
- Downsizing storage capacity and cost by more closely matching onsite production to demand.
- Having the ability to address smaller market segments that historically were unable to afford cryogenic liquefaction equipment (e.g., smaller gas utility peakshaving).

By being able to affordably own and operate small liquefiers, it is possible to envision concepts such as more highly distributed LNG peakshaving for utility operations, LNG peakshaving for smaller natural gas utilities or end users, or producing LNG onsite for fleets such as transit buses, refuse haulers, etc.

GTI developed a comprehensive economic model to examine the capital, operating, and maintenance cost factors for this system. The GTI Natural Gas Liquefaction Economic Model uses a Monte Carlo technique to address uncertainty and variability on cost and operating factors to obtain a representative life-cycle cost for making LNG.

We examined two different liquefaction system sizes—1,000 gpd and 5,000 gpd—and two different fuel scenarios. The fuel scenarios were intended to address conventional natural gas as well as using an “opportunity fuel” such as landfill gas. Table 1 shows summary results for the size and fuel choice matrix. Sensitivity of the LNG production lifecycle costs to scale—for systems between 1,000 and 5,000 gpd—is approximately \$0.15-\$0.2/gallon. The sensitivity to fuel choice between natural gas and an opportunity gas is significant—in the range of \$0.25-\$0.28/gallon. A 5,000 gallon per day (gpd) system operating on natural gas is projected to have a lifecycle cost of about \$0.50-\$0.55/gallon; using landfill gas would push this down to below \$0.30/gallon.

Table 1: Natural Gas Liquefier Lifecycle Cost

	Initial		Final
	1,000 gpd	5,000 gpd	1,000 gpd
Industrial Natural Gas	\$0.72	\$0.53	\$0.75
Opportunity Fuel	\$0.44	\$0.27	\$0.46

Figure 2 shows the results of a sensitivity analysis of a 1,000 gpd liquefaction system using natural gas or an opportunity gas such as landfill gas. For this scenario, the life-cycle costs—and the sensitivity of life-cycle costs—is clearly dominated by the cost of gas. Uncertainty or variability of the cost of major capital items such as gas cleanup have relatively minor effect on the lifecycle cost in both cases¹.

¹ For this analysis, gas cleanup cost was kept constant for the natural gas and landfill gas cases. In practice, the cost for natural gas cleanup would be less expensive; the analysis indicates the benefits of lower-cost cleanup systems are relatively minor unless substantially below the range shown.

Lifecycle Cost Using Natural Gas

Lifecycle Cost Using "Opportunity" Gas

Figure 2: Cost Drivers for 1,000 gpd Liquefier System

2. Introduction

LNG use is increasing worldwide. This is due in part to growth in the number of exporting countries (and liquefaction capacity), increasing international LNG trade, and expanding market applications. Figure 3 shows the trend in LNG international trade during the past three decades. LNG trade is expected to continue this upward trend as several countries seek to capture and export natural gas as a liquid to an ever-increasing number of customers (e.g., Japan, Korea, France, Spain, and the United States).

The LNG industry in the U.S. is mainly comprised of a select number of wholesale applications. This includes import terminals for pipelines supplies, natural gas utility peakshaving, and, to a lesser extent, baseload natural gas used to supply remote communities (Figure 5). These applications typically use the “form value” of LNG for stationary storage and ultimately vaporize the LNG to a gas to feed into natural gas pipelines.

Gas Utility LNG Peakshaving Plant

Baseload LNG for Remote Community

Figure 5: Traditional LNG Applications

Further expansion of the U.S. and North American LNG import infrastructure is expected as international supply sources compete to satisfy part of the growing U.S. natural gas demand. This LNG growth is likely to include throughput expansions at current terminals, changes in operating status of mothballed terminals, and construction of new terminal facilities (either in the U.S., in nearby locations such as Baja California, or at off-shore facilities).

Growth in domestic demand for LNG will arise from conventional as well as new market applications. These will help to meet interstate pipeline requirements or, in some cases, support specific large-scale power generation projects. Internal LNG growth within the United States is expected in new applications such as LNG-powered vehicles, use of LNG as an industrial/power plant standby fuel, and other non-utility applications.

The natural gas vehicle (NGV) market is growing in certain regions such as California and Texas, but generally is not sufficiently developed in most areas of the country to support large-scale LNG plant investments. The developing NGV market, however, provides an opportunity for considering smaller-scale liquefaction concepts that have lower facility costs.

According to DOE, NGVs in the U.S. have experienced steady growth in both CNG and LNG. Natural gas for vehicle use in the U.S. represents an annual displacement of about 120 to 130 million gallons of gasoline and diesel fuel. Of that, the LNG share is about 10% and growing. There are about 2100 LNG vehicles and 45 LNG fueling stations in the U.S.

Source: DOE EIA/GTI. Fuel use in gasoline gallon equivalent (gge)

Figure 6: U.S. LNG Vehicle Growth

The LNG vehicle market includes high-fuel-use vehicles such as refuse handlers, transit buses, shuttle buses, and over-the-road trucks. These vehicles are normally fueled with diesel. With a low fuel economy of approximately five miles per gallon, these vehicles consume large quantities of fuel on a daily basis. For these users, LNG represents a viable alternative fuel that can satisfy their range requirements better than compressed gas.

Recent growth in LNG demonstration programs and dedicated LNG fueled fleets has created a market demand for a small liquefier in the 1,000 to 10,000 gpd capacity range. The development of such a low cost liquefaction technology is the subject of this report. The approach taken by GTI to develop a lower-cost natural gas liquefaction system was made possible by the identification of new refrigerant mixture compositions that permit simple system configurations using off the shelf HVAC compressors and other components.

On the supply side, concepts such as converting “stranded” natural gas reserves and capturing opportunity fuels such as landfill gas or digester gas and turning them into value-added liquid fuels may provide economically viable opportunities for small-scale liquefaction. Along with other gas-to-liquids technology, small-scale liquefaction can play a role in developing and converting these resources into viable natural gas supply sources.

Cryogenic natural gas liquefiers are commercially available for many large ocean-transport and natural gas utility peakshaving applications. These units are custom-made, large capacity plants with substantial initial capital cost. By contrast, there is very little experience with the design and operation of small natural gas liquefiers (under 20,000 gpd).

The challenge with scaling liquefaction technology down to a small-scale involves specific capital cost and efficiency. Figure 7 provides a depiction of the typical scale and specific costs associated with large-scale natural gas liquefaction plants. By achieving reasonably low specific

costs at a smaller size, it is possible for these types of systems to achieve favorable economics by minimizing the cost associated with LNG transport.

Figure 7: Natural Gas Liquefaction Costs

The challenges, though, in achieving this type of economics at a small scale are real. The specific cost benefits of large plants—for example, allocating design, engineering, and permitting costs over a large investment—coupled with the purchasing power of obtaining natural gas as a large customer—are tangible. For small-scale liquefaction plants to be viable, the design needs to be:

- Simple to permit relatively low first cost, pre-packaged equipment that minimizes onsite construction costs
- Reliable to minimize the need for onsite operating and maintenance personnel.

The benefits of small-scale plants include the minimization or elimination of the need for over-the-road LNG transport, matching LNG supply with demand, minimizing at risk capital, and providing a value-added service (e.g., peakshaving) on a scale that normally would not be viewed as economic.

3. Natural Gas Liquefaction

Liquefaction Technology Options and Considerations

Natural gas can be converted from a gas to a liquid form (i.e., liquefy) at temperatures ranging from -220°F to -260°F (depending on gas composition and pressure). The sensible and latent energy required to transform methane from, for example, a gas at 60°F to a saturated liquid at -231°F (both at 30 psig) is approximately:

Sensible Heat (Btu/lb)	155.25
Heat of Condensation (Btu/lb)	204.75
Total Change (Btu/lb)	360.00

This is an idealized representation of the change between natural gas at these two conditions. As a practical matter, converting methane from a gas to a liquid results in inefficiencies due to the thermodynamic limitations of mechanical systems, heat transfer considerations, etc. As a point of reference, natural gas has a heating value of approximately 22,500 Btu/lb (higher heating value).

Table 2 lists some of the approaches that have been used to achieve natural gas liquefaction, with some discussion on potential trade-offs with each approach.

Table 2: Summary of Representative Liquefaction Approaches

LIQUEFIER TYPE	OPERATING PRINCIPLE	REMARKS
Precooled Joule-Thomson (JT) Cycle	A closed-cycle refrigerator (e.g. using Freon or propane) pre-cools compressed natural gas, which is then partially liquefied during expansion through a JT valve.	Relatively simple and robust cycle, but efficiency is not high. Used in Anker Gram onsite liquefier for LNG truck fueling (which is no longer operating).
Nitrogen Refrigeration Cycle (closed Brayton/Claude cycle)	Nitrogen is the working fluid in a closed-cycle refrigerator with a compressor, turboexpander, and heat exchanger. Natural gas is cooled and liquefied in the heat exchanger.	Simple and robust cycle with relatively low efficiency. Using multiple refrigeration stages can increase efficiency. Used in CryoFuel Systems Hartland LFG liquefier demonstration.
Cascade Cycle	A number of closed-cycle refrigerators (e.g. using propane, ethylene, methane) operating in series sequentially cool and liquefy natural gas. More complex cascades use more stages to minimize heat transfer irreversibility.	High-efficiency cycle, especially with many cascade steps. Relatively expensive liquefier due to need for multiple compressors and heat exchangers. Cascade cycles of various designs are used in many large-capacity peakshaving and LNG export plants.
Mixed-Refrigerant Cycle (MRC)	Closed cycle refrigerator with multiple stages of expansion valves, phase separators, and heat exchanger. One working fluid, which is a mixture of refrigerants, provides a variable boiling temperature. Cools and liquefies natural gas with minimum heat transfer irreversibilities, similar to cascade cycle.	High-efficiency cycle that can provide lower cost than conventional cascade because only one compressor is needed. Many variations on MRC are used for medium and large liquefaction plants. ALT-EI Paso Topock LNG plant uses MRC. GTI is developing simplified MRC for small plants (under 10,000 gpd).
Open Cycles with Turboexpander, Claude Cycle	Classic open Claude cycle employs near-isentropic turboexpander to cool compressed natural gas stream, followed by near-isenthalpic expansion through JT valve to partially liquefy gas stream.	Open cycle uses no refrigerants other than natural gas. Many variations, including Haylandt cycle used for air liquefaction. Efficiency increases for more complex cycle variations.

LIQUEFIER TYPE	OPERATING PRINCIPLE	REMARKS
Turboexpander at Gas Pressure Drop	Special application of turboexpander at locations (e.g. pipeline city gate), where high-pressure natural gas is received and low-pressure gas is sent out (e.g., to distribution lines). By expanding the gas through a turboexpander, a fraction can be liquefied with little or no compression power investment.	This design has been applied for peakshaving liquefiers, and it is currently being developed by INEEL in cooperation with PG&E and SoCalGas to produce LNG transportation fuel. Very high or "infinite" efficiency, but special circumstances must exist to employ this design.
Stirling Cycle (Phillips Refrigerator)	Cold gas (usually helium closed cycle using regenerative heat exchangers and gas displacer to provide refrigeration to cryogenic temperatures. Can be used in conjunction with heat exchanger to liquefy methane.	Very small-capacity Stirling refrigerators are catalog items manufactured by Phillips. These units have been considered for small-scale LNG transportation fuel production.
TADOPTR	TADOPTR = thermoacoustic driver orifice pulse tube refrigerator. Device applies heat to maintain standing wave, which drives working fluid through Stirling-like cycle. No moving parts.	Currently being developed by Praxair and LANL for liquefaction applications including LNG transportation fuel production. Progressing from small-scale to field-scale demonstration stage.
Liquid Nitrogen Open-Cycle Evaporation	Liquid nitrogen stored in dewar is boiled and superheated in heat exchanger, and warmed nitrogen is discharged to atmosphere. Counterflowing natural gas is cooled and liquefied in heat exchanger.	Extremely simple device has been used to liquefy small quantities of natural gas. More than one pound of liquid nitrogen is required to liquefy one pound of natural gas. Nitrogen is harmless to atmosphere. Economics depends on price paid for liquid nitrogen.

Adapted from USA Pro/California Energy Commission report (with modifications by GTI).

GTI Liquefier Technology Description

The approach used by GTI in designing and developing its small-scale liquefaction technology was to focus on use of commercially available and proven equipment to the maximum extent possible. Secondly, emphasis was placed on a simple system design to minimize equipment costs. Finally, within these constraints, the system was designed for maximum efficiency and flexibility to operate in warmer weather climates.

To satisfy these requirements, GTI devised a system that mimics standard commercial HVAC and refrigeration practices. The key differences are the use of five-component refrigeration mixture and a custom heat exchanger that permits natural gas to be taken from ambient conditions to a liquid state in a single pass.

Natural Gas Cleanup

Natural gas can be made to convert from a gas to a liquid at temperatures ranging from -220°F to -260°F (depending on gas composition and pressure). Natural gas is not pure methane. The composition of natural gas is predominantly methane, with a number of decreasing heavy hydrocarbons and trace species. Some of these components will be partially or wholly soluble or insoluble in liquid at cryogenic conditions, resulting in a possible three-phase situation involving gaseous, liquid, and solid compounds (Figure 8).

Insoluble compounds may remain largely as a gas (e.g., nitrogen) or may condense as a solid (e.g., carbon dioxide, water, heavy hydrocarbons). To address these concerns, natural gas cleanup systems are needed to nearly completely remove certain key actors such as carbon dioxide, water, and hydrogen sulfide.

Figure 8: Natural Gas Phase Considerations

There are several methods that can be used to address these minor to trace constituents. The approach used in this program included a two-step process. The first addressed sulfur removal, followed by combined removal of carbon dioxide and water. Both of these employed the use of solid adsorbents, a proven approach in natural gas liquefaction.

Natural gas can typically be expected to have less than 30 ppmv of total sulfur, mostly in the form of hydrogen sulfide. Trace levels of various natural gas odorants will also normally be found in natural gas—these are all sulfur compounds. Water content for natural gas is typically expected to be in the range of three to seven pounds per million standard cubic feet. This is about 100 to 150 ppmv.

Table 3 provides a summary of typical natural gas composition values for the major and minor components often found in this fuel. Carbon dioxide content for natural gas averages less than one percent, but higher levels are possible. Nitrogen tends to be around one to two percent, but can be substantially higher.

Table 3: Selected Natural Gas Composition Survey Data

	10%-tile	Mean	90%-tile
Methane	83.94	93.05	95.98
Ethane	5.62	3.47	2.14
Propane	0.99	0.67	0.36
C4 and higher	0.31	0.33	0.38
Nitrogen	6.28	1.67	0.53
Carbon dioxide	1.37	0.81	0.73

The compositional requirements for addressing landfill, wastewater treatment, and digester gases are substantially more challenging. These include very high levels of carbon dioxide as well as a greatly expanded list of trace contaminants.

4. GTI Liquefier System Design, Development, and Testing

Liquefier System Design

The liquefier design of the major components for the mixed-refrigerant-cycle-based, 1000 gpd unit is based on the same techniques utilized in the experimental 250 gpd mixed refrigerant liquefier that was assembled and tested in Phase I. The major difference is that a natural gas engine drives the compressor in the present 1000 gpd unit. The prior system used an inverter-driven electric motor.

A schematic of the major components of the 1000 gpd liquefier is shown in Figure 9.

Figure 9: Schematic of 1000 gpd LNG Liquefier

As seen in this figure, a natural gas engine drives a commercially available HVAC screw compressor through a 2:1 speed increasing gearbox and coupling. The engine is selected to run at a maximum speed of 1800 rpm, and the 2:1 speed increaser raises the input compressor speed to the rated 3600-rpm level of commercial screw compressors. The high-pressure refrigerant at the compressor output has significant oil entrained in it, which must be removed before the refrigerant enters the extremely cold sections of the main heat exchanger. This oil removal is accomplished initially by an oil separator, from which the oil is cooled and returned to the compressor suction. The gas output from the oil separator is then passed through two stages of coalescing-type oil filters. The oil removed in these filters is periodically returned to the compressor suction.

After oil filtration, the high-pressure refrigerant is cooled by an aftercooler, which was selected as an evaporative condenser in the 1000 gpd liquefier, versus the air-cooled unit of the experimental liquefier. The evaporative condenser is in effect a cooling tower and can reduce the temperature of the refrigerant gas lower than a typical air-cooled heat exchanger.

The cooled high-pressure refrigerant gas enters the top of the system's three-circuited main heat exchanger. The refrigerant gas passes downward through the heat exchanger and is condensed into liquid. The liquid refrigerant is then expanded into a two-phase mixture that enters the heat exchanger at its bottom. The extremely cold refrigerant then flows upward and evaporates, while accepting heat from the refrigerant condenser fluid as well as from the natural gas circuit. The low-pressure vapor exits the top of the heat exchanger and returns to the compressor suction.

The natural gas to be liquefied enters the third circuit of the main heat exchanger at its top and flows downward and liquefies, exiting the exchanger at its bottom.

Design Point Component Selection

The refrigerant selection is the same as in the experimental liquefier unit and is composed of a proprietary mixture of Nitrogen, Methane, Ethane, Iso-Butane, and Iso-Pentane. The design point cycle for the refrigerant is shown in Figure 10, and forms the basis for the liquefier's main component selection.

Figure 10: Refrigerant Pressure-Enthalpy Diagram at the System Design Point

At the design point, the compressor suction is chosen to be 10 psig (24.7 psia), and is controlled to that value in the system by movement of a motor-driven expansion valve. The high-pressure side of the system is assumed to be 200 psig, which is based on the system experience with the experimental liquefier and on system modeling with HYSIS. Note the major load on the evaporator section of the main heat exchanger is due to the condensing refrigerant, leaving a relatively small enthalpy available for LNG liquefaction.

The calculations associated with the system's engine and compressor requirements are based on condensing 1000 gpd of a 10-component mean U.S. natural gas mixture with major element molar composition: 92.9% Methane, 3.3% Ethane, 2.1% Nitrogen, and 0.8% CO₂. The inlet natural gas is assumed to be at 30 psig and 60°F.

Based on the assumed natural gas composition and inlet conditions above, a flow of 55.0 SCFM is computed, with refrigeration needs of 53,090 Btu/hr to produce 1000 gpd of 30 psig LNG at a saturated liquid temperature of -239°F.

The available enthalpy per pound of circulating refrigerant (13 Btu/lb) is computed based on a compressor suction of 10 psig, 50°F, a compressor discharge of 200 psig, with an adiabatic compressor efficiency of 65%, and an aftercooler refrigerant outlet temperature of 80°F.

In the absence of standards for rating points for small LNG liquefiers, we have chosen a 95°F, 40% RH ambient air condition as the rating point inlet to the system's evaporative-condensing aftercooler. For reference, an ambient dry bulb of 95°F is the ASHRAE standard for air-cooled air conditioning and refrigeration equipment. At 95°F, 40%RH, the air has a wet bulb of about 65°F, which results in a reasonable 15°F approach for the aftercooler to achieve the assumed 80°F outlet refrigerant temperature.

To achieve the 53,090 Btu/hr refrigeration for LNG liquefaction, a required refrigerant flow of 68 pounds per minute was computed, which corresponds to a compressor suction displacement of 477 cfm. The selected screw compressor is a model 2010 manufactured by Hartford Compressor, and has a displacement of 543 cfm at 3600 rpm. The actual refrigerant displacement of the compressor is 543 cfm multiplied by the volumetric efficiency of the unit, estimated to be about 90%, or very close to the design needs of the liquefier.

An adiabatic compressor efficiency of 65% requires about 200 horsepower at the compressor shaft input, which led to the selection of the Cummins GTA8.3 engine, which is rated at 200 horsepower.

A proprietary program of Chart Industries, LaCrosse, WI, was used to support the design of the custom-made main heat exchanger. Similarly, Baltimore Air Coil provided technical support in the equipment selection process for the aftercooler.

A list of equipment utilized in the system is given in Table 4 as procured by GTI.

Table 4. Liquefier Components, Manufacturers, and Costs

Component	Manufacturer	Model/Description	Cost, \$
Main Heat Exchanger	Chart	Custom Aluminum Plate Fin	41,830
Engine/Radiator	Cummins	GTA8.3	31,774
Compressor/Integral Oil Pump/Slide Ind.	Hartford	2010 Screw	17,340
Welded Steel Piping	Advance Mech. Systems	1.5-6"	9,000
Programmable Logic Controller, I/O Comp.	Allen Bradley	Micrologic 1500	8,700
Speed Increaser /Clutch/Assembly Labor	Twin Disc, Cummins	RM120D/BD290	7,500
After Cooler/Sump Heater	Baltimore Air Coil	VF1-009-12E	6,590
Tubing/Fittings	Swagelok, etc.	Various Sizes	5,000
Insulation	Luse-Stevenson	Heat Exchanger and Low Press. Piping	4,800
Steel Skid	Am. Metal Fabricators	8.5'x14'	3,770
Expansion Valve	Jordan	MK8000MV	3,674
Valves, Regulators	McMaster, Fluid Process Control	49415K34, 540-30BPLAA1E	2,500
Oil Filters	Finite Filter	HNOL-10DSJ,-4DSJ	2,448
Explosion Proof Solenoid Valves	Automatic Switch Co.	Various Sizes	1,500
Oil Separator	Carlyle	KH31ZZ212	1,377
Oil Heat Exchanger	Carlyle	KH51ZZ184	1,299
External Oil Pump	Viking Pump	3GJPM Ex. Proof Motor	728
Oil Sump	Henry Technology	RF-14084-800	571
Oil Pump Filters	Norman Filter	30MF2-10ML, 30MF224N-10ML	400
Total Cost =			150,801

Liquefier System Development

The components of the LNG system are standard parts purchased through a variety of vendors. With the exception of the gas clean up equipment, these components were placed on an 8-foot by 14-foot transportable skid to allow for servicing of components. The layout process started with a two-dimensional layout drawing, which aided in the early development of the location of the parts on the skid. After examining the number of interconnected parts involved in the system, it became apparent that a three-dimensional drawing would assist in the skid layout and in the plumbing of the components. However, the majority of the parts did not have 3D CAD-based models so they need to be created.

The first component placed on the skid layout was the compressor, since it is the heart of the system. One difficulty encountered with the compressor was its discharge port, which is below its associated mounting pad. The compressor was raised to allow for discharge port clearance. This also meant that the engine height would change since it was directly coupled to the compressor. Considering the need to raise both the engine and the compressor, it was determined a one piece-mounting fixture would work best. This mounting fixture had to support both the

engine and compressor, dampen the vibration to the skid, ensure proper shaft alignment, and allow for any service to the sub-system. The fixture was created from 'C'-channel, which maintains flexibility for changes.

Between some components, the size of the pipe had to change due to the different size connections on the variety of components. This was accomplished fairly simply with the use of reducer/expander fittings. Flexible pipe was used at the suction and discharge of the compressor. This was done to relieve any stress that may arise from vibrations and to allow for thermal expansion.

There are two circuits that needed piping, the refrigerant loop and the oil circulation piping. The refrigerant circuit is welded pipe, the majority of which is black pipe. The cryogenic side of the refrigerant system is stainless steel. The piping for the oil circuit is stainless steel tubing. Once everything was connected, pressure checks were made to insure a leak-free system.

Figure 11 shows a top view of the CAD skid layout and Figure 12 depicts an elevation view.

Figure 11: Top View of 3D CAD Layout of Skid Components

Figure 12: Elevation Perspective View of CAD Layout of Skid Components

Liquefier System Testing

Natural Gas Liquefier Test Facility

A complete test facility was constructed around the GTI LNG liquefier, including gas clean up and LNG tank storage, in order to measure the liquefaction performance of the basic refrigeration system. Following is a description of the system components and instrumentation that were assembled.

Figure 13 is a depiction of the complete natural gas liquefier test facility, from natural gas inlet to the cryogenic LNG storage vessel.

Figure 13: Liquefier System Components and Instrumentation

Pipeline natural gas is available at the GTI laboratory test location at about 160 psig. In the test set up, this gas was input to the GTI liquefier at two locations; one input was directly into the Cummins engine, through a pressure regulator, and the other to the top of the liquefier's main heat exchanger after passing through the gas cleanup system. The gas consumption of the engine is measured via an inlet volume type flowmeter. A MicroMotion Coriolis type meter measures the gas into the liquefier.

The Cummins GTA 8.3 engine drives the liquefier's Hartford model 2010 screw compressor through a Twin Disc 2:1 speed increaser gear set and a mechanical coupling. After starting at about 1000 rpm, the nominal operating speed of the Cummins engine was set, via an internal speed controller, to an operating speed of about 1800 rpm.

The refrigerant discharge gas from the screw compressor gas has considerable entrained oil in aerosol form. Virtually all of this oil must be removed before the refrigerant reaches the cold sections of the liquefier's main heat exchanger. Most of the oil is removed in a Carlyle cyclonic type separator and is returned to the compressor after cooling in a Carlyle air-cooled heat exchanger. The temperature of the returning oil is monitored by a thermocouple. The remaining oil in the refrigerant gas is removed by two stages of coalescing oil filters made by Finite Filter. Oil from these filters is also returned to the compressor periodically by solenoid valves.

The capacity (or effective displacement per revolution) of the screw compressor is adjusted by moving a suction cut-off slide valve through a series of solenoids (not shown). A pump located inside the compressor, driven by the compressor shaft, provides the pressure for this hydraulic actuation of the compressor slide. The position of the compressor capacity-controlling slide is available through an external potentiometer.

After passing through the coalescing filters, the high-pressure refrigerant gas is cooled by a Baltimore Air Coil evaporative condenser, which has an internal water circulation and distribution means. The unit has a water sump heater to prevent water from freezing.

The cooled, high-pressure refrigerant gas then enters the condensing channels of the main system heat exchanger at its top, and passes downward, being condensed into liquid at the lower cold end of the heat exchanger. The heat of condensation is transferred to the low-pressure, boiling refrigerant that flows upward in the exchanger's evaporator channels. After condensation, the refrigerant liquid is expanded into a two-phase state over a Jordan motor-driven cryogenic expansion valve. From there, the refrigerant passes upward through the heat exchanger and then to the compressor suction.

Transducers and thermocouples monitor the pressure and temperature conditions of the refrigerant entering and exiting the heat exchanger. The refrigerant temperature at the expansion valve outlet is the coldest temperature in the system, and should be about 20°F colder than the condensing temperature of the natural gas feedstock to produce an efficient liquefaction process.

Thermocouples were placed outside the main heat exchanger at five vertical locations. These measurements allow for the monitoring of the approximate refrigerant temperature distribution within the heat exchanger and are useful in determining the optimum refrigerant mixture composition.

The natural gas entering the liquefier's main heat exchanger passes through a desulfurizer and a series of three towers for water and carbon dioxide removal. These are described in more detail in a later section of this report.

The pressure and flow of the cleaned natural gas entering the main heat exchanger is controlled to the test condition by a combination of the flow areas manually set in valves V1 and V2.

The LNG exits the main heat exchanger at the bottom and flows through valve V2 to the bottom of a 1500 gal storage tank via a vacuum jacketed line. The level of the LNG liquid in the storage tank is measured by the difference in pressure between the top and bottom of the tank, after

computing the density of the liquid via tank pressure and liquid temperature measurements. The gallons stored in the LNG tank are determined by a LNG height versus tank volume computation.

A nominal 20-psig LNG tank relief valve was used in the testing. Flow through the relief valve during liquefaction is a combination of vapor formed due to heat infiltration into the tank, and vapor released from the LNG produced by the liquefier, if it is not a pure liquid after expansion over valve V2. The vapor flow through the relief valve is measured via a volume type flowmeter.

System Control and Data Acquisition

A complete data acquisition and monitoring system was developed to monitor and collect data on the system operation and performance. Data acquisition of all system measurements was accomplished via Labview software, with all signals stored in Excel files for analysis. An example copy of the monitoring computer screen is shown in Figure 14.

Figure 14: Data Acquisition Display on Laboratory Computer For Run 3/10/03-1

Instrumentation was employed throughout the system to provide real-time monitoring. Thermocouples were installed at various points including locations from the top to the bottom of the heat exchanger.

Control over the engine speed, compressor slide valve position, expansion valve position, various oil and refrigerant solenoid valves was implemented in a dedicated Allen Bradley Micrologic 1500 PLC. This unit was programmed by GTI.

A series of test runs on the LNG liquefier system were run during the period 2/19/03 and 3/20/03. Various ambient temperature conditions were encountered during that period, and to fully explore the liquefier characteristics, different compressor capacities, natural gas flows, and inlet gas pressures were imposed on the system.

Start-up Testing Protocol

Before starting the engine, at an approximate speed of 1000 rpm, the compressor slide capacity is set to its minimum level by moving the screw compressor slide to its minimum position. After engine start, it was run from 10 to 15 minutes, before increasing compressor capacity to the test run value. The engine speed was then set to its operating level of about 1800 rpm. As the system pulls down the low temperature sections of the main heat exchanger to the test condition, the compressor suction pressure is controlled to about 10 psig, via PLC adjustment of the motor driven expansion valve position.

After the liquefier has established about -290°F at the expansion valve outlet, conditioned natural gas, from the system clean up tri-towers, is introduced into the main heat exchanger, and liquefaction begins. The natural gas flow rate and pressure into the main heat exchanger is then set to the test values via manual adjustments over valves.

Liquefier Testing Results

A summary of the measured test results on the LNG liquefier system is presented in Table 5.

Table 5: Measured LNG Liquefier Test Results

Test Run	Amb. Temp.	Compr Slide	Liquefier Gas Inlet Flow	Liquefier Gas Inlet Press.	After-Cooler Outlet Temp.	Compr. Disch. Press.	Suct. Press.	Engine Gas Use	LNG Tank Press. low/high	LNG Tank Relief Valve Flow	LNG Produced
	F	%	SCFM	psig	F	psig	psig	SCFM	psig	SCFM	gal/day
2/19-1	35	90	57.7	35	60.8	135	10.6	18.2	10.8/17.6	5.7	932
2/19-2	40	90	42.1	20	62.6	138	10.3	18.1	12.0/12.5	7.8	613
2/19-3	33	90	38.2	30	62.0	131	10.0	17.8	11.5/16.0	3.3	624
2/21-1	45	90	60.0	35	63.5	138	10.3	18.0	10.7/12.5	8.5	925
2/21-2	50	90	44.8	21	62.5	132	9.4	17.0	11.5/12.5	7.8	659
2/26	30	90	90.5	78	66.3	181	9.9	20.0	15.4/17.8	23.9	1190
2/27	35	100	81.7	77	78.5	163	8.4	26.0	14.5/19.8	5.1	1368
3/6	28	100	59.5	57	61.5	143	10.1	20.0	11.5/13.0	9.5	893
3/7-1	30	100	81.7	79	65.5	191	9.2	28.0	15.0/18.0	24.3	1028
3/7-2	34	100	79.9	35	69.3	175	8.6	27.0	17.8/19.5	29.4	905
3/10-1	13	100	90.0	80	63.3	169	8.2	26.5	12.7/17.3	10.8	1416
3/10-2	18	100	79.0	51	62.8	165	9.4	26.5	13.5/14.5	15.2	1141
3/18-1	41	100	83.3	25	78.0	145	8.7	26.0	12.0/14.0	35.2	860
3/18-2	44	100	55.2	25	90.0	155	9.2	26.0	12.5/12.5	34.8	364
3/20-1	49	100	NA	15	82.0	135	8.4	25.0	0/0	NA	886
3/20-2	55	100	NA	16	84.0	145	8.2	25.0	0/0	NA	394
3/20-3	62	50	16.1	28	67.0	132	11.8	13.0	0/0	4.8	201

As seen in Table 5, various ambient temperatures ranging from 13°F to 50°F were encountered during the test series. The compressor slide was set to 90% capacity level during the initial testing, as some failures were occurring in the clutch facing of the speed increaser assembly. After replacement of the clutch facing to a metallic composition, tests were continued at 100% slide position.

It was noted early in the testing that inlet gas flow rate was an important variable in determining the liquefaction rate of the system, and that the product of gas inlet pressure and gas inlet flow was the independent variable combination that provided the clearest system evaluator.

During testing, the compressor suction pressure was kept essentially constant at about 10 psig by controlling the effective orifice area of the motor driven expansion valve. This provided a low refrigerant temperature at the coldest part of the system, namely at the expansion valve outlet, which insured that a significant portion of the main heat exchanger was below the natural gas liquefaction temperature.

The highest liquefaction rates occurred at a combination of high gas inlet pressures and high gas flow rates. For example, run 3/10-1 produced a measured liquefaction rate of 1416 gpd with an inlet gas flow of 90 SCFM at 80 psig. Conversely, at the low inlet gas pressure of 30 psig and inlet flow of 38 SCFM during run 2/19-3, the system produced only 624 gpd.

Based on the measured results presented previously, Table 6 shows computed system performance variables of interest.

Table 6: Computed LNG Liquefier Test Results

Test Run	Engine Gas Use Per LNG Gallon	LNG Production /Total Inlet Gas Flow	LNG Production /Total Inlet Gas Flow (Assuming Full LNG Tank Relief Flow To Engine)
	SCF/gal	SCFM/SCFM	SCFM/SCFM
2/19-1	28.1	0.676	0.731
2/19-2	42.5	0.560	0.643
2/19-3	41.1	0.613	0.651
2/21-1	28.0	0.653	0.732
2/21-2	37.1	0.586	0.670
2/26	24.2	0.593	0.756
2/27	27.4	0.698	0.733
3/6	32.3	0.618	0.701
3/7-1	39.2	0.515	0.662
3/7-2	43.0	0.466	0.643
3/10-1	26.9	0.669	0.737
3/10-2	33.4	0.595	0.695
3/18-1	43.5	0.433	0.568
3/18-2	102.9	0.246	0.370
3/20-1	40.6	NA	NA
3/20-2	91.4	NA	NA
3/20-3	93.1	0.381	0.457

Figure 15 shows the dependency of the liquefier production versus the natural gas inlet pressure to the system's main heat exchanger. As seen, although the general trend is that higher

liquefaction rates are achieved with higher inlet gas pressure. However, there is considerable scatter of the data. This points to the influence that other system variables have on the LNG production rate.

The primary reason for an increase in liquefaction rate with higher inlet gas pressure is the higher condensation temperatures that accompany the higher gas pressures. The evaporator refrigerant channels in the main system heat exchanger set up a given temperature distribution in the exchanger's surface area which is exposed to the natural gas flowing downward in the gas channels. At higher gas pressures, condensation of the gas begins earlier in its flow passage downward, resulting in higher liquefaction rates. The highest liquefaction rate of about 1416 gpd (test 3/10-1) was achieved with about 80-psig natural gas supply pressure.

The higher liquefaction rates with higher pressures must be considered with the means necessary to achieve those pressures in applications where the natural gas pressure available is much lower. The initial and operating cost of gas booster compressors in low-pressure applications must be balanced against the increased liquefaction rates achievable by the liquefier.

Figure 15: LNG Liquefier Production vs. Natural Gas Inlet Pressure

A better correlation of the liquefaction rate of the system is achieved by considering the independent variable as the product of the natural gas pressure and flow. The result of this correlation is seen in Figure 16. Note the data scatter is greatly reduced relative to that shown previously.

Figure 16: LNG Liquefier Production vs. Product of Natural Gas Inlet Flow and Pressure

Clearly, the natural gas inlet flow rate is an important variable, and it is believed that higher flow rates, at a given gas pressure, increases the rate of liquefaction by scraping the liquid droplets off the heat exchanger surface areas in the condensing portion of the main heat exchanger and thereby forcing the liquid downward in the channels. At low gas flows, the droplets rely on gravity to pull them downward and may tend to get hung up in the tortuous channels. There would be a limit to increasing liquefaction by increasing the gas flow, as extremely high flows would tend to prevent liquid droplet formation. During these tests that limit was not achieved.

Figure 17 shows the refrigerant cycle on a pressure-enthalpy diagram for Test 3/10-1. The test measurements indicate that the refrigerant at the condenser outlet was not entirely condensed into a pure liquid. The cooling effect of the system's mixed refrigerant cycle is proportional to the difference between the enthalpy at the compressor suction and the enthalpy of the refrigerant at the after cooler outlet.

High ambient temperatures will tend to reduce system liquefaction capacity, as shown in the test of 3/18-2, where airflow was restricted to the after cooler and the outlet refrigerant reached 90°F. This reduced the liquefaction rate to the 364 gpd level at the 25 psig natural gas inlet set during this run.

Refrigerant Pressure-Enthalpy Diagram Test 3/10-1

Figure 17: Pressure-Enthalpy Diagram For the Refrigerant Cycle in Test 3/10-1

Figure 18 shows the natural gas consumption of the Cummins engine expressed as a function of the LNG liquefaction rate. At the highest liquefaction test points the engine uses about 27 SCF per gallon of LNG produced.

Figure 18: Natural Gas Consumption of Engine

Figure 19 shows the LNG production efficiency achieved in the GTI liquefier, presented as the ratio of the LNG produced (in SCFM units) divided by the total inlet gas flow (SCFM) to the system, i.e., the flow to the liquefier plus the flow to the gas engine. As seen, at the higher liquefaction rates, this ratio is about 0.70, indicating that 70% of the total gas input to the system results in LNG liquid in the storage tank. Note this storage tank liquid was created and stored at about 15 psig.

Figure 19: LNG Production Efficiency

In the tested system, the flow through the relief valve on the LNG storage tank was vented to the atmosphere. There was no attempt to route this flow to the system's engine. In an actual commercial application, this gas would be recovered and used to help fuel the engine.

Figure 20 shows the resulting increase in the liquefier's LNG production efficiency with recovering vented gas to fuel the engine. Flatter and somewhat higher production efficiency results from this system improvement, with increases to the 0.75 SCFM/SCFM level. Since the LNG tank is pressurized to about 15-psig and since the engine requires only several inches of water gas pressure, it is felt that this system improvement would be entirely feasible.

LNG Production Efficiency With Full LNG Tank Relief Flow To Engine

LNG Production/Total Inlet Gas Flow (SCFM/SCFM)

Figure 20: LNG Production Efficiency with Full LNG Tank Relief Flow To Engine

Natural Gas Cleanup System Design and Development

The natural gas cleanup system was primarily designed to address removal of sulfur compounds, carbon dioxide, and water. These are key actors that will form solids and plug up narrow passages and valves within a system.

The sulfur removal used a packed tower, filled with a solid adsorbent that GTI previously evaluated for this type of duty. The tower was approximately 4 inches in diameter and 6 feet tall and packed with a product called NUSORB FC-3, an activated carbon impregnated with metal oxides. The nominal particle size is three millimeters, with a surface area of 1000 m²/gram. The recommended minimum contact time is 1.5 seconds. The sulfur holding capacity, on a weight basis, was estimated to be in the range of 5 to 15%. The material was obtained through a company called Selective Adsorption Associates (www.selectiveadsorption.com). The product is manufactured by Nucon International, Inc. (www.nucon-int.com). This material was obtained in 25-pound containers at a cost of approximately \$7 per pound.

Figure 21 shows a picture of this tower (smaller black vessel). This was constructed by GTI using Schedule 40 pipe with 300 psig flange end caps. The tower was pierced at certain points to provide thermocouple temperature measurements.

Figure 21: Sulfur Removal Tower

The water and carbon dioxide removal system was based on a zeolite molecular sieve. This was specified by GTI and procured from Zeochem (www.zeochem.com). The material is identified as Z10-02; this was based on a 9x16 mesh bead size. The material is relatively inexpensive at about \$2 per pound.

The molecular sieve was used to charge a three-tower assembly that was obtained by GTI through an arrangement with a company that had previously designed and tested the unit for pre-treating natural gas for liquefaction service (Figure 22).

Figure 22: Tri-Tower Gas Treatment System

The towers are designed with an external pressure vessel and an inner pressure vessel that held the adsorbent material. Each inner pressure vessel was approximately 16 inches in diameter and 6 feet high.

Figure 23 provides a view of the inside of an individual tower. The inner vessel was charged with the Zeochem Z10-02 molecular sieve adsorbent. This included the use of screens for controlling dust and ceramic balls to act as supports and a method for flow distribution.

The annular space between the inner and outer vessel was filled with a material to insulate the inner vessel. This helps facilitate the heat-driven regeneration process. The top of the tower is hinged to provide access. Each tower is designed to withstand 150 psig (MAWP). The hinged top used a gasket for sealing in the gas and was bolted down to contain the pressure.

Several problems were encountered by GTI during the assembly of these towers. In particular, several attempts were needed to properly gasket and seal the inlet and outlet ends of each individual tower. These problems were eventually resolved.

Figure 23: Inside View of Tri-Tower System

The tri-tower system was originally intended operate with one tower active (adsorbing water and carbon dioxide) while the other two towers were in various states of regeneration (heating to drive off adsorbed carbon dioxide and water as well as cooling off to prepare for the next adsorption cycle). This process is part of an otherwise conventional temperature swing adsorption (TSA) process.

The tri-tower system, and variations on this concept, is desirable because it permits continuous operation. The system was designed to advantageously use the gas engine drive. Figure 24 shows the logic for this system, outlining the integration of regeneration process with the natural gas prime mover used for refrigerant compression. In this graphic, the three towers are depicted over time, with the logic for the first time interval shown. In this design, the first interval is approximately 4 hours and the total cycle time is 12 hours. Note that the natural gas that ultimately goes to the natural gas prime mover is carrying a higher concentration of water and carbon dioxide and is consumed as fuel gas in the system. The logic for the subsequent two time sequences is similar, but shifted to begin with tower 2 and tower 3, respectively. Heat from the engine exhaust is used to help drive the thermal regeneration process.

Figure 24: Original Operational Logic for Tri-Tower System

While the above concept is a valid and appropriate approach for a small-scale liquefaction system with a gas engine drive, there were a series of problems with the equipment and control system provided by the vendor. This was further complicated by the financial difficulties of this company that made it very difficult to get them to complete their scope of supply to GTI. This culminated in GTI being virtually unable to contact anyone from the firm. GTI was ultimately required to complete the effort on the gas cleanup system.

During this period, significant time was lost in testing the liquefier system. Due to several problems with the clean-up equipment as provided, it was ultimately decided to simplify the tri-tower operation. GTI removed the piping, instrumentation, and controls and designed the system so that all three towers operated in series. This allowed extended operation during periods of liquefaction, followed by periodic off-line regeneration. Regeneration was accomplished using heated nitrogen. This approach was appropriate for the operational strategy used during system development and testing. In practice, the unit was able to achieve approximately 24 hours of run time before regeneration was required.

Natural Gas Cleanup System Testing

The desulfurizer and molecular sieve tri-tower system were operated during all periods of the natural gas liquefaction tests. With only a few exceptions, the operating performance of the gas cleanup equipment was very good.

Test samples were taken of natural gas before and after the desulfurizer and the molecular sieve towers. A copy of a typical major component analysis on the inlet natural gas appears in the Appendix.

The following is a summary of gas sampling before and after the adsorbent systems.

	Inlet	Outlet
Sulfur Compounds ¹	0.99 ppmv	BDL (<0.05 ppmv) ²
Carbon Dioxide ³	7,950 ppmv (0.79%)	29.7 ppmv
Water ⁴	Variable	Variable dewpoints of -60 to -100°F

1. Using AQSTM D6228-98. Standard conditions are referenced to 14.696 psia, 0°C. The mercaptans measured included t-butyl mercaptan and dimethyl sulfide.
2. BDL = below detectable limit.
3. ASTM D-1945 using gas chromatography.
4. Measured using online aluminum oxide moisture sensor from Panametrics.

Based on these data, the system was performing adequately to ensure low levels of sulfur compounds and carbon dioxide. As a general rule, it is believed that CO₂ levels need to be below 50 ppmv to avoid freeze-up problems. Measured CO₂ values were below 30 ppmv.

The measurements for water used a real-time aluminum oxide moisture sensor. The unit never measured below a -100°F dewpoint during the testing program. This may be due to limitations in the molecular sieve or the sensitivity of the instrumentation or sampling system at these extremely dry conditions. The absolute humidity at these conditions is very low.

As noted, the operation of the gas cleanup system during the testing program was generally reliable and satisfactory. Problems that did arise included:

- An incident that resulted in substantial pressure drop across the desulfurizer.
- An incident of impeded liquid natural gas flow that was likely attributed to solids plugging a valve or the heat exchanger.

The desulfurizer pressure drop issue was traced to an inadvertent problem with insulation from the molecular sieve towers being blown loose during regeneration and plugging the discharge of the desulfurizer. This problem was readily addressed by a change in the valve and piping system.

The plugging problem was attributed to breakthrough of carbon dioxide or water in the molecular sieve system. A regeneration of the towers after this incident removed the problem and no further occurrences were noted during operation of the gas cleanup system or liquefier.

Our conclusions are that the materials used for desulfurization and removal of carbon dioxide and water in this program are suitable for treating natural gas for liquefaction service. The system was reliable, though real-time instrumentation on the outlet conditions may be helpful. The aluminum oxide moisture sensor measured dewpoints as low as -100°F. No attempts were made to attempt to validate the accuracy of this measurement.

5. Liquefier System Markets and Economics

Small-Scale Natural Gas Liquefier Markets

There are several markets that would benefit from the increased availability of LNG supplies that could be provided by strategically placed small-scale liquefier units. Among these markets are transportation markets, including transit buses, waste hauling and short distance goods transport. Other potential transportation markets include long distance and interstate goods hauling, rail transport and marine applications such as ferries, tugs and service vessels. Non-transportation markets that would also benefit from small-scale liquefiers include supply (exploit shut in gas sources), liquefaction and use of landfill gas sources. Local gas distribution companies could also make use of systems for peak-shaving supplies as well as providing gas to isolated areas where pipelines are not practical or economical.

Transportation Markets

Current U.S. energy demand totals nearly 100 quadrillion British thermal units (quads) and is growing about 1.2% per year. Of this total, 38% comes from petroleum, 25% from natural gas, 23% from coal and the remaining 14% from nuclear, hydroelectric and other sources. The transportation sector accounts for 27% of our total energy needs, and more than 68% of the petroleum consumed in the U.S. Since more than 50% of our petroleum is imported, this means that most of today's vehicles are fueled by imported, petroleum-based fuels. Conversely, nearly 90% of the natural gas consumed in the U.S. is domestically produced, with most of the balance coming from Canada. Greater use of natural gas as a transportation fuel offers significant benefits to the U.S. balance of trade while also addressing issues related to energy security and policy.

For the Natural Gas Industry, the transportation sector continues to represent a key potential growth market. In 2002, total transportation natural gas usage (CNG and LNG) increased by 17.6% from 2000 levels to more than 124 million gasoline equivalent gallons (GGE). This is in comparison to the combined total for gasoline and diesel of more than 170 billion GGE, which grew 4.5% over the same period.

Since 2000, the number of light-duty CNG vehicles has increased by 25.3% to 99,568 and the number of heavy-duty vehicles rose by 25.8% to 26,773. For LNG, heavy-duty vehicles in use grew by 52% to 2,735. Continued expansion of CNG and LNG vehicles into key fleet markets such as transit buses, refuse hauling, and medium and heavy-duty transport vehicles can expand gas use while addressing local environmental concerns and reducing dependency on foreign oil.

A key factor in the use of LNG for transportation applications is obviously availability of sufficient quantities of LNG supply. Many of the current LNG fleets depend on supplies that are fairly remote from the point of use resulting in significant cost and availability issues. In California, having the largest number of LNG fleets and vehicles, most LNG is currently delivered from a liquefaction plant in Topock, Arizona (Figure 25). The Topock plant is roughly 250 miles (one way) from the Los Angeles and San Diego areas, and more than 500 miles from Sacramento, making the fuel trucking distances to most California LNG fleets significant. Other sources that are used for the California fleet market come from far away plants in Wyoming, Colorado and Kansas.

Past history shows that the cost of delivering LNG (typically in 10,000 gallon LNG tank trucks) is in the range of \$1.50 to \$3.00 per mile (one way). Costs may be higher for short distance or smaller or sporadic deliveries. Based on these approximations, a 200-mile delivery of 10,000 gallons at \$2.50 per mile results in an additional cost of \$0.05 per LNG gallon, for example.

Figure 25: Current Sources of LNG Delivered to California

When considering the market for small-scale liquefaction systems for the transportation market, both the LNG vehicles as well as the CNG vehicle markets should be considered. While LNG is required for the LNG vehicles, there are also definite economic advantages to using LNG to make CNG (L-CNG). Natural gas can be compressed more efficiently in liquid form than in gaseous form. This inherent advantage means that smaller motors can be used to compress LNG into high-pressure CNG than are used to compress equal amounts of pipeline gas. Smaller motors translate into lower capital, energy and maintenance costs, reduced noise and other benefits for station developers. In addition, much higher fill rates are available with L-CNG and with less extensive storage tanks (cascades).

Transit Market

The transit bus market is a significant market for natural gas sales. Over the past several years, the number of natural gas transit buses in operation has significantly increased. According to a recent NGVC survey utilizing data collected by the American Public Transportation Association (APTA) and the NGVC, there are more than 6,200 transit buses in operation at more than 85 transit agencies around the United States. This represents more than 11 percent of all transit buses and 97 percent of all alternative fuel buses. As of January 1, 2002, an additional 1,313 new natural gas buses were on order. There also are hundreds of other natural gas buses in operation at airports, universities and other locations, that are not included in APTA's annual survey of transit bus operators.

Figure 26: Orange County California LNG Transit Bus

The number of natural gas buses now represents approximately 11 percent of all transit buses and 97 percent of all alternatively fueled transit buses are powered by natural gas. Natural gas buses include dedicated, bi-fuel, and hybrid applications. More than 90 percent of the existing natural gas buses are fuel by compressed natural gas (CNG). Liquefied natural gas (LNG) bus fleets are concentrated mostly in Arizona, California, and Texas.

Waste Hauling

Over the past 30 years, municipal solid waste (MSW) generation per capita has doubled in the United States, from 2.3 to 4.7 pounds per day. There are currently an estimated 136,000 refuse collection trucks, 12,000 transfer vehicles, and 31,000 dedicated recycling vehicles in use today. A recent survey by Inform, Inc. found there to be 26 U.S. waste hauling fleets operating natural gas vehicles. Of these, 227 are CNG vehicles and 465 are LNG vehicles. The study further showed that at the time of the study, the companies planned to add 2,221 future natural gas vehicles. Of these planned vehicles, 436 will be CNG and 1,785 will be LNG. This is a very significant addition of LNG fleet vehicles, and will definitely stress the currently already tight LNG supply network. Small-scale liquefaction units would be an economically viable means of serving this new load, particularly for the larger fleets being planned.

Figure 27: Waste Management LNG Refuse Truck

Centrally Fueled Heavy-Duty Urban Trucks

Another relatively untapped market is the short distance return to base goods hauling. These fleets can be categorized as return-to-base fleets, with a centralized operation mode where vehicles begin and end their delivery routes at a facility operated by the fleet owner. This is an extremely large market. The American Trucking Association reports that more than 1.6 million class-eight trucks are operated for business in the U. S., consuming nearly 10,000 gallons of diesel per year. In a 2001 study of California fleets, there were 55 operating LNG trucks (39 in grocery distribution and 16 other return-to-base) with 47 vehicles on order, and another 87 vehicles projected.

Figure 28: Raley's LNG Grocery Distribution Fleet Truck

Potential Transportation Vehicle Markets

A logical extension of the return-to-base trucking market is the long distance, over-the-road hauling market. Over-the-road class-eight tandem-axle tractors are the highest fuel consumption vehicles. Unlike centrally fueled urban fleets, many of these trucks consume up to 35,000 gallons of diesel annually. Moreover, many travel fixed routes, fueling at only one or two captive stations, so LNG might be supplied to fleets of long-haul units via relatively few LNG stations. According to EIA statistics, on-highway diesel fuel usage accounted for 33 billion gallons in distillate fuel oil sales in 2001, or more than 55% of total US sales. The major obstacle to this market is the lack of LNG sources along major interstate routes. Small-scale liquefiers strategically placed along major trucking routes would open up a tremendous market potential for LNG trucking. The low cost units could take advantage of interstate pipelines which many times run along the major roadways as their source of natural gas. The units could take additional economic advantage of the high-pressure gas source to more effectively produce LNG.

Another major oil consuming market is rail transportation. In 2001, railroad use accounted for more than 3 billion gallons of US distillate fuel oil sales according to EIA statistics. There has been past efforts to enter this market, but with little success. The rail market would seem to be well suited to LNG with fixed routes, and the ability to carry significant amounts of LNG in insulated rail cars. This would tend to require fewer LNG refueling sources across the country, and could be well served by small-scale liquefier technology.

Other minor markets for LNG as a transportation fuel include various marine applications. The various types of vessels used in service applications in ports and waterways including tug boats and off-shore oil drilling platform service vessels. It is likely that a single small liquefier could effectively serve as a refueling station for an entire port. LNG has also been successfully used for a 100-car, 300-passenger ferry in Norway since February, 2000, running 19 hours a day, 7 days a week without interruption.

Other Markets

Additional markets that represent applications for the small-scale liquefaction unit include non-traditional gas supply exploitation and energy distribution uses. Gas supply market potential includes shut-in gas wells and other sources such as landfill gas. Energy distribution application potential includes LDC peak shaving, supplying gas to utility markets that are isolated from the distribution pipeline network.

Supply Applications

Stranded gas is in a reserve that has little or no access to a market, and therefore has little prospect of being produced in the near term. Almost 60% of roughly 5,000 tcf of proven natural gas reserves worldwide can be categorized as remote or stranded. LNG technology provides one method of transporting these reserves to market. Of the roughly 2,200 stranded fields, less than 100 contain the minimum of 5 to 6 tcf of gas needed to justify a large scale LNG plant.

Another source of gas that can be considered a "stranded" reserve are the many landfill gas facilities. Landfills are a good potential source of energy because they produce a methane-rich gas (approximately 55% methane, 45% CO₂) soon after startup and for up to 25 years after closure. There are approximately 1,500 municipal landfills in the U.S. The average LFG production rate

is approximately 1.4 million SCF per day of methane, this resulting in an average LNG yield of 5,400 gallons per day. Landfills are also typically located near large metropolitan areas, making them near the desired points of distribution for vehicular LNG and other natural gas end uses.

Peak Shaving

Supplemental gas (also called "peakshaving" and "standby gas") is a means of reducing or eliminating the need for industrial, commercial and municipal gas consumers to purchase "firm" gas supply or other fuels for backup when supply capacity is short. A small liquefier would liquefy and store gas offline for vaporization during periods when supplies are short or system capacity is inadequate. Use of LNG for peak period supply would allow LDCs to avoid costly upgrades to pipelines and distribution systems.

Isolated Distribution Markets

Remote communities typically are captive customers. Pipeline economics are poor because of the pipeline length and the market size. A small liquefaction plant may be viable to communities that are isolated from gas pipelines, but are within several miles of either stranded gas reserves or a pipeline source. About 400 U.S. communities (with populations between 250 and 10,000) have no access to gas pipelines. There are several trucked-in LNG including: West Yellowstone, MT; Jackson, WY; Wendover, UT; and 15-20 proposed projects, mostly in Alaska. Competition to LNG includes propane, electricity, wood and heating oil. Fuel prices often exceed \$10 per MMBtu.

Natural Gas Liquefaction Economics

Natural gas liquefaction plants are typically made to produce large quantities of LNG. A large plant producing over 1,000,000 gallons per day can have capital costs in the range of \$1.5 to \$2 billion. This is derived from GTI analysis of plants in this size range having a specific cost of about \$100-200/LNG gallon per day of capacity (Figure 29). Normally, due to negative scaling effects, the cost of a small-scale liquefier could easily fall in the range of \$600 to \$900/LNG gallon per day.

Figure 29: Scaling Issues for Natural Gas Liquefaction

Whether a small-scale liquefier is economically viable depends on several factors, including its first (fixed) cost and the (variable) cost of transporting LNG from a large central production source to a remote LNG user. The economic proposition for the small-scale liquefier under development by GTI relies on cost-effective components and packaging techniques to reduce the first cost. Locating LNG production near demand also helps users to avoid over-the-road transportation and storage costs. The program goal was to develop a small-scale liquefaction plant that had attractively low initial capital costs—in the range of \$200 to \$400/gallon/day.

The construction, operation, and maintenance of a small-scale liquefier involve a variety of cost elements. Overall economic performance is heavily dependent on the system configuration, utilization, and feedstock cost (i.e., natural gas price). With such a wide variety of factors, it is often difficult to identify those characteristics having the greatest effect on the system economics. GTI developed a Natural Gas Liquefaction economic model to address the small-scale liquefaction system.

Natural Gas Liquefier Economic Model

Investors considering a small-scale liquefier may have quite different investment evaluation criteria and constraints. Some users may evaluate capital projects using an internal rate of return or net present value (NPV) measure. Others may use cash payback, accounting rate of return, return on investment, or impact on profit or contribution. Fleets will need to evaluate costs and benefits in comparison to their existing practices. Investors need to be aware of the role of possible tax effects, incentives, and evaluation methods when deciding among several alternatives. To address these complexities, several financial evaluation methods and incentive structures are included in the economic model.

The “output” of the economic model is the after-tax life-cycle cost. This amount provides enough income over the specified equipment life to recover all first and recurring costs on a

present value basis. Any desired profit margin would be additive to the after-tax life-cycle cost result. Fuel excise taxes are not included as they "pass through" to the end consumer.

In analyzing the economic performance of the liquefier, the first step is to establish first costs (equipment, installation, etc.), recurring costs, and the intervals between events giving rise to recurring costs. A cost element will be economically influential either because of its:

- Expected magnitude
- Frequency
- Uncertainty or inherent variability, or
- A combination of magnitude, frequency, and variability.

To gauge the impact of variation or uncertainty, a Monte Carlo simulation is employed. This simulation approach offers the ability to vary any or all desired cost elements simultaneously within individually specified ranges. Each range is represented by a probability distribution. Figure 30 and Figure 31 show sensitivity charts of LNG fuel price to several system parameters for the 1000 gpd ($4 \text{ m}^3/\text{day}$) liquefier system. The first figure is generated using U.S. average industrial gas rates while the second assumes that an opportunity fuel, such as landfill or digester gas, is available at approximately one-third the cost of industrial gas.

Figure 30: Cost Drivers for 1000 Gallon/Day ($4 \text{ m}^3/\text{day}$) System, Industrial Gas