Processes help turn rich LNG into lean gas

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- **LNG receiving terminals in some areas require less C_2 and C_3 and most LNG produced cannot meet gas specifications required in the US**
- **This complicates operations and marketing of liquefaction facilities because Asian-Pacific markets favor higher heating value LNG**

One of the challenges faced by both liquefaction facilities targeting the US market and import terminals is meeting the strict requirements for the composition of natural gas in terms of calorific value and quality.

This is particularly true for import projects targeting the large Californian market that must consider California Air Resource Board (CARB) regulations in addition to the local distribution companies.

An LNG receiving terminal with a C_{2+} or C_{3+} separation facility can flexibly receive LNG cargoes with varying compositions while producing a send-out gas meeting stringent calorific value specifications with modest capital and operating costs.

Foster Wheeler has recently developed C_{2+} and C_{3+} NGL recovery processes (patent pending) for LNG that allows import and processing of rich cargoes.

These receiving terminal processes allow effective and economic processing of the LNG because they leverage traditionally installed receiving terminal equipment, and maximize the use of the "cold energy" in the LNG.

The designs use low-temperature LNG as a cooling medium for the column overhead to achieve an economic separation.

The design does not need supplemental gas compression because the LNG is used as a refrigerant in a direct-contact condenser, which recondenses the C_1 vapor from the column overhead to produce a lean LNG for liquid pumps, vaporization, and send-out.

The vaporizers act as the heat source for the separation column. This process features low cost, no additional gas compression, and uses the equipment already present at a receiving terminal. It may also be retrofitted to existing facilities.

Compared to a system that requires send-out gas compression or an inert-gas injection system to dilute the send-out gas, C_{2+} or C_{3+} separation can reduce the capital investment by at least 40 percent and decrease operating costs.

LNG limitations

Sempra-owned local distribution companies such as Southern California Gas Co., San Diego Gas and Electric and Pacific Gas and Electric have raised concerns over CARB specifications that limit ethane as well as inert components in natural gas 1,2.

To add additional confusion, the CARB specification is currently under review and may well change to a Wobbe Index-based specification with additional limits on C_{4+} and inerts.

LNG receiving terminals on the US West Coast will require less C_2 and C_3 in the LNG than most existing baseload plants produce.

This will complicate operations and marketing of liquefaction facilities because Asian Pacific markets favor higher heating value LNG, potentially leading to a dual-specification production and also to shipping challenges.

This, and other integration and logistic issues mean it is often not cost effective to reduce C_2 and C_3 at the baseload plant.

This type of operation reduces LNG production at the baseload LNG facility and requires export of at least three products, so it is often better to install a C_{2+}/C_{3+} extraction process at the LNG receiving terminal.

Compared to a system that requires send-out gas compression or an inert-gas injection system to dilute the send-out gas, C_{2+} or C_{3+} separation can reduce the capital investment by at least 40 percent and decrease operating costs.

Most LNG produced worldwide cannot meet the gas specifications required in North America. Table 1 shows that CARB specifications for CNG have lower limits than typical pipeline gas (10 mole % of C_2, 6 mole % of C_3)3.

Table 1 also shows a specification for the Mexico pipeline gas. Only selected LNG suppliers provide light LNG to meet the CNG specifications. Table 2 shows typical LNG compositions.

If a terminal requires C_2 or C_3 for fuel, it will need to process LNG with a component extraction unit.

LNG buyers have different requirements; therefore, reducing C_2 and C_3 at the baseload LNG plant is not desirable because of:

- Less LNG produced
- Additional compression equipment required
- The desire to operate all LNG trains at the same conditions

After initial feed-gas treatment in an LNG baseload plant (acid-gas removal, dehydration, mercury removal, etc.), a scrub column removes benzene and C_{5+} components to ensure that they will not freeze in the main cryogenic heat exchanger.

To control LNG calorific value, the scrub column also removes some C_{2+} components.

The feed-gas pressure for most baseload LNG plants is greater than 60 bara. If the plant must remove heavier hydrocarbon components to meet a typical North American market calorific value of 1,070 btu/cu ft, the scrub column must operate at a pressure less than the critical pressure of about 40 bara.

For a baseload LNG plant, refrigerant efficiency depends on the operating pressure of feed gas entering the main cryogenic heat exchanger.

A lower calorific value, therefore, would require recompression of feed gas from the scrub column to the main cryogenic heat exchanger, which is expensive.

In addition, multiple LNG train plants often operate at similar conditions and produce the same composition of LNG instead of multiple specifications.

The removal of C_2 and C_3 makes more sense for receiving terminals especially for the North American market. Although these additional facilities increase capital costs, they create an opportunity for competitive pricing because the plant can meet export specifications while having LNG from many different suppliers.

Terminal processing

Two process schemes can give the LNG receiving terminals more flexibility for processing the arriving LNG to meet the buyers' gas specifications.

Both of the process schemes eliminate the compressors in this; however, it is necessary to use a gas direct contact condenser where the overhead vapor from the heaviers extraction column is re-condensed by using cold LNG.

The column overhead condensers that are plate-and-fin or shell-and-tube exchangers utilize the low temperature of the LNG as the cooling medium.

Table 1: Typical pipeline specifications

Component, Mol%	California Air resources Board CNG		Mexico natural gas
	Minimum	Maximum	Maximum
C_1	88		
C_2	6		
C_{3+}		3	3.6

Table 2: Typical LNG compositions

Component, Mol%	Terminal location						
	Das Island, Abu Dhabi	Whitnell Bay, Australia	Bintulu, Malaysia	Arun, Indonesia	Lumut, Brunei	Bontang, Indonesia	Ras Laffan Qatar (RasDas)
Methane	87.10	87.80	91.20	89.20	89.40	90.60	89.60
Ethane	11.40	8.30	4.28	8.58	6.30	6.00	6.25
Propane	1.27	2.98	2.87	1.67	2.80	2.48	2.19
Butane	0.141	0.875	1.36	0.511	1.30	0.82	1.07
Pentane	0.001	--	0.01	0.02	--	0.01	0.04

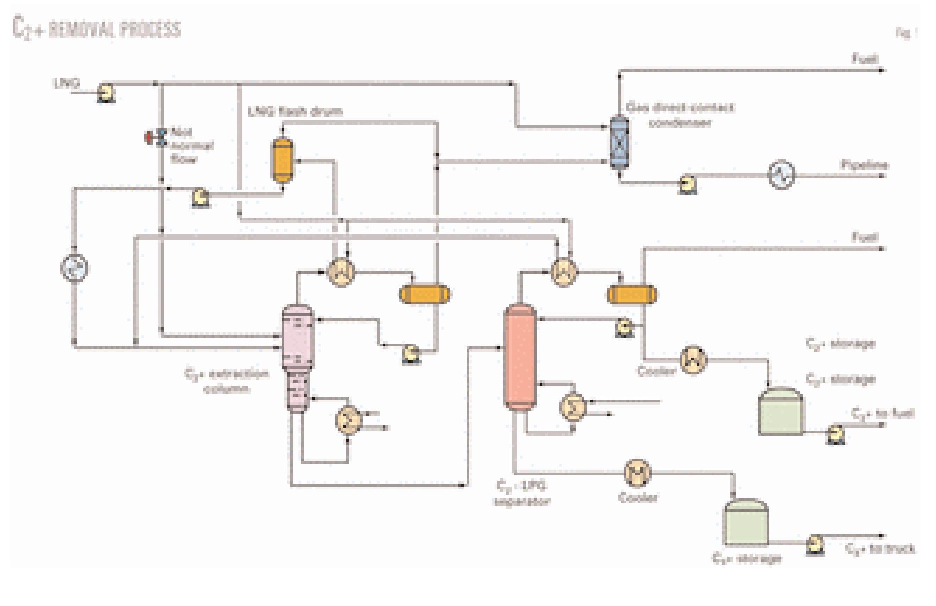

Figure 1: This shows a detailed process flowsheet (Patent Pending)

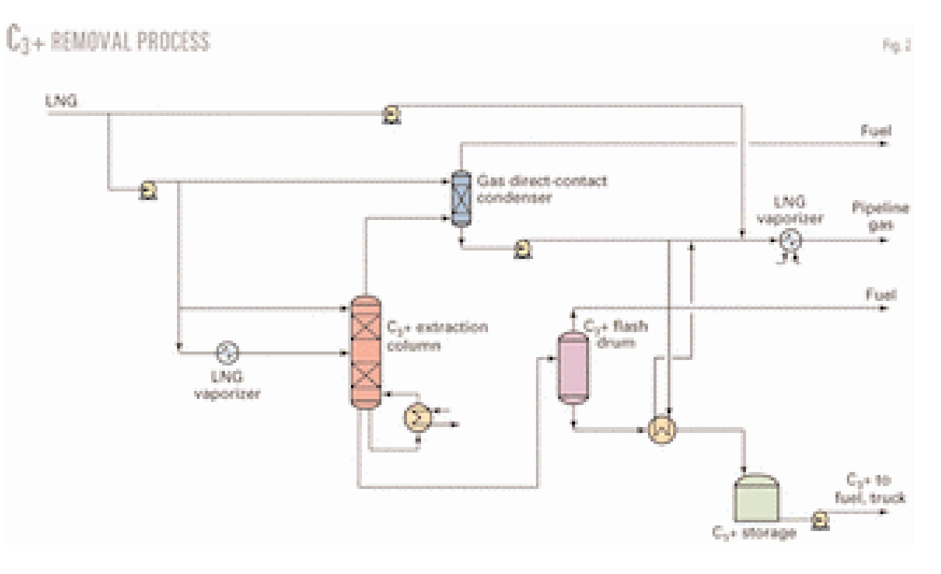

Figure 2: The C_{3+} extraction column processes about 8% of the 19% LNG fed to the C_3-removal system (Patent Pending).

The column feed is vaporized through the existing vaporizers. The condensed LNG is then pumped up to the gas send-out pressure (about 85 Barg) for regasification in the vaporizers and sent out via the export gas pipeline.

In the first scheme, the design is to meet a specification of a maximum of 6-mol% ethane using the C_2 plus extraction column for separating C_1 from C_2 and heavier components. The C_2 /LPG separator is used for C_2 extraction from the propane and heavier components if required.

In the second scheme, the design is to meet a specification of a maximum of 3.6 mol% propane and heavier components.

A packing-bed column is used for C_3 plus extraction. The top section is used to remove C_3 and heavier components from C_2 and lighter. The bottom of the column is used for C_1 extraction from C_3 and heavier components

C_{2+} hydrocarbon removal: In a typical example, rich imported LNG contains excessive heavier components: 87 mole % C_1, 11.4 mole % C_2, 1.3 mole % C_3, and some heavier components.

The fractionation section processes about half of the LNG from the send-out pump.

The other half of the cold LNG serves as the refrigerant and mixes with overhead vapor (richer in methane) from the C_{2+} extraction column to produce a warm condensed LNG in the gas direct-contact condenser.

Condensed LNG contains less than 6 mole % C_2, which is pumped to the vaporizers for send-out to the export pipeline.

Total installed cost of the C_{2+} hydrocarbon removal system is $35-40 million for an LNG receiving terminal with a capacity of 1,000 MMcfd send-out gas.

Fractionation section

The C_{2+} component removal section also produces LPG for export and a C_2-rich cut for plant fuel or export.

The fractionation section consists of two fractionation columns: a C_{2+} extraction column and a C_2 -LPG separator.

A gas direct-contact condenser recondenses vapor from the extraction column reflux drum.

The extraction column receives vaporized LNG from the existing LNG vaporizers and cold LNG from the existing LNG feed pumps.

The cold LNG is also the cooling medium in the column overhead condensers. Extraction column products include a leaner LNG vapor stream overhead and a C_{2+} stream from the bottoms.

Overhead vapor from the extraction column recondenses against cold LNG in a gas direct-contact condenser.

Condensed vapor is pumped up to about 85 barg, regasified in the existing LNG vaporizers, and sent to the export gas pipeline.

Extraction column bottoms go to the C_2 -LPG separator, which produces an LPG stream from the bottoms and a C_{2+} cut from the overhead.

Both streams are cooled and stored in tanks.

Operation description

The exact amount of cold feed LNG that the fractionation section processes (roughly 50%) depends on the required C_{2+} specification. The exact LNG pipeline pressure will depend on the buyer's specification.

The extraction column usually operates at 40 bara. A lower operating pressure improves separation efficiency, but also increases column size and reduces the fractionation column overhead vapor condensing capacity.

This pressure setting is reasonable because it is less than the system critical pressure needed to achieve separation.

The C_2 -LPG separator operates at 20 bara. The operator can change C_2 and C_3 specifications based on fuel-quality requirement or other quality requirements for using or selling C_2 and C_3.

Further process

In the second example, LNG feed contains excessive heavy components: 87.8 mole % C_1, 8.3 mole % C_2, and 3.9 mole % C_{3+}.

In this example, the LNG produced satisfies a Mexican pipeline gas C_{3+} specification of less than 3.6 mole%.

The C_{3+} removal system processes about 19% of the pumped-out LNG. Remaining LNG feeds directly to the vaporizers.

Total installed cost of the C_{3+} hydrocarbon removal system is $5-8 million for an LNG receiving terminal with a gas send-out capacity of 1 billion cubic feet per day.

The C_{3+} extraction column processes about 8% of the 19% LNG fed to the C_3 -removal system as shown in Figure 2 (Patent Pending).

The remaining 11% of the LNG enters the gas direct-contact condenser as an absorbent and refrigerant. It mixes with the C_1-rich overhead vapor from the extraction column to produce a condensed LNG with a maximum 3.6 mole % of C_{3+}.

To ensure that condensed liquid stays in the liquid phase, LNG leaving the direct-contact condenser is sub-cooled at least 5° C.

This sub-cooling requires about 11% of the cold pump-out LNG to recondense and refrigerate the extractor overhead vapor.

Condensed LNG is pumped up to about 85 barg, regasified in the existing LNG vaporizers, and sent to the gas pipeline.

Process description

The C_{3+} removal process also produces LPG products for export or for terminal fuel use.

Approximately 30% of the LNG that enters the packed-bed extraction column feeds directly to the top as an absorbent.

The other 70% first goes to existing LNG vaporizers; the vapor then enters the column between the two packed beds.

The extraction column separates C_{3+} components from the LNG feed. Vapor leaving the extraction column mixes with cold LNG in the gas direct-contact condenser.

Extraction column bottoms stream flows to the C_{3+} flash drum, in which light components flash to the top as fuel.

The C_{3+} stream from the bottom of the flash drum first depressurizes to atmospheric pressure, is cooled with cold LNG, and feeds to the C_{3+} storage tanks.

Liquid from the direct-contact condenser pumps to about 85 barg, flows through existing LNG vaporizers and to the export gas pipeline.

Pressures

The extraction column operating pressure is 40 bara. A lower operating pressure improves separation efficiency, but increases column size because this is less than the critical pressure for separation.

A process simulation determined the need for four theoretical stages between the liquid and vapor feed and three stages between the vapor feed and bottoms for the C_1 and C_{3+} separation.

In the extraction column, 90% of the C_3 flows to the column bottoms, which contains no more than 10 mole % of C_1.

Vapor leaving the extraction column is recondensed when mixed with cold LNG in the gas direct-contact condenser. To ensure that the condensed liquid is at least 5° C. sub-cooled for easy pumping, cold LNG flow to the condenser is at least 20% more than vapor flow.

Conclusions

This study shows that the example process schemes for extracting C_2 and/or C_3 NGL from rich imported LNG are potentially an effective and economical way to meet US send-out gas specifications.

Integrating a C_{2+} or C_{3+} NGL separation facility into an LNG receiving terminal allows it to receive various LNG feed compositions while meeting stringent calorific value export gas specifications.

The minor modifications to the typical LNG receiving terminal equipment avoids the need for additional compression and maximizes utilization of normally required equipment.

In comparison to a system requiring send-out gas compression or an inert gas injection system for diluting the send-out gas, it should be possible to reduce the capital investment by at least 40% while realizing significantly lower operating costs.

This is an updated version of an article first published by PennWell Corp.

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