NATURAL GAS SPECIFICATION CHALLENGES IN THE LNG INDUSTRY

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ABSTRACT

LNG is primarily methane, but in most cases also contains ethane, propane and heavier components. Until recently potential buyers of LNG had some flexibility in the ranges of natural gas properties they were able to accept. In many cases the LNG buyer exported electric power and not natural gas, and therefore did not have to meet the needs of natural gas consumers. But now new markets are opening up where pipelines to consumers already exist, and the consumers have specifications which lie outside the range of the LNG produced at many locations. The challenge is to understand the requirements of the potential new markets and plan ways to meet the requirements of both the LNG exporters and importers.

This paper reviews the natural gas specifications of current consumers and those expected in the near future, such as North America and others. These specifications are compared to those of LNG being produced, and the options and problems for conditioning the natural gas on either end of the LNG trade are explored in detail.

INTRODUCTION

Natural gas contains methane, heavier hydrocarbons and inert components which affect burner performance. For this reason pipeline companies and LNG buyers specify allowable ranges of components and heating values. These requirements vary widely depending on the market location. But despite this, LNG product specifications have not been a major plant design issue. Historically, plant designs were based on long term contracts to few buyers, and there was little need for flexibility in the plant designs, either on the liquefaction or receiving ends of the trade.

However, the situation is changing as LNG trade becomes more global. The owners of liquefaction plants are targeting more than one market, and new markets have requirements that are not always compatible with existing trades. Furthermore the growing spot market for LNG provides opportunities for buyers and sellers who have the ability to be flexible on product specifications. As a result there is now considerable movement towards technical solutions for conditioning LNG on liquefaction and receiving ends.

WORLDWIDE NATURAL GAS SPECIFICATIONS

Natural gas specifications have several purposes, including corrosion prevention, avoiding liquid drop out in pipelines, and burner performance [1]. The specifications for corrosion prevention limit concentrations of CO_2, $H2_S$, mercaptans or total sulfur. LNG facilities treat gas down to 50 ppmv CO_2 to avoid freezing in the cryogenic processing unit, and therefore meet receiving end pipeline specifications. The sulfur specifications are typically consistent with the Japanese market which limits H_2S to 5 mg/Nm^3 maximum and total sulfur content to 30 mg/Nm^3 maximum. Meeting these specifications for Japan will also meet the requirements for Europe and the US (except California which limits total sulfur to 18 mg/Nm^3).

Acid gas is typically removed in an amine unit using the principal that an alkaline solvent will remove acid gas components. CO_2 is a weak acid compared to H_2S, and the process of taking CO_2 down to 50 ppmv often controls the acid gas removal design (H_2S is a stronger acid and therefore easier to remove). The exceptions are plants with high concentrations of mercaptans in the natural gas feedstock. Mercaptans are very weak acids and are removed by means other than straight chemical absorption.

To prevent liquid dropout, natural gas pipeline companies limit the amount of butane, pentane and heavier components. LNG plants must remove heavier hydrocarbon components to prevent freezing in the liquefaction process, and the heavies removed become a natural gasoline by-product. Heavy end specifications are therefore not difficult to meet in most liquefaction plants.

Specifications to prevent corrosion and liquids dropout are fairly consistent, and requirements of the LNG plant itself (i.e. avoiding freezing during cryogenic processing) tend to make those specifications almost universal. The requirements for heating value and gas interchangeability remain. Here the specifications vary significantly, as do LNG products from worldwide sources, and it is interchangeability that is presenting the greatest challenges.

Early LNG trade was primarily to Japan from Pacific Rim and Middle East export plants and to Europe from Northern Africa. The Japanese specifications vary depending on the importing utility company, but typically have a high heating value between 39.7 and 43.3 MJ/Sm^3 (Megajoules per Standard meter cubed at 1 atm and 15 degrees C, which converts to 1065 to 1160 Btu/SCF for a standard cubic foot at 14.73 psia and 60 degrees F). This relatively high range permits maximum use of infrastructure by moving greater combustion heat capacity for a given volume. European countries typically allow wider ranges. Spain, for example, allows a range between 35.0 and 44.9 MJ/Sm^3 (940 and 1205 Btu/SCF).

Figure 1 below shows representative heating value specifications for different countries. In one example a gas with HHV = 42.6 MJ/Sm^3 is suitable for the Japanese and Korean markets, but is too high for the US or UK. In the second example a gas with HHV = 37.2 MJ/Sm^3 meets US/UK specs but has an HHV too low for Korea or Japan. Both examples however are within the ranges allowed for France and Spain.
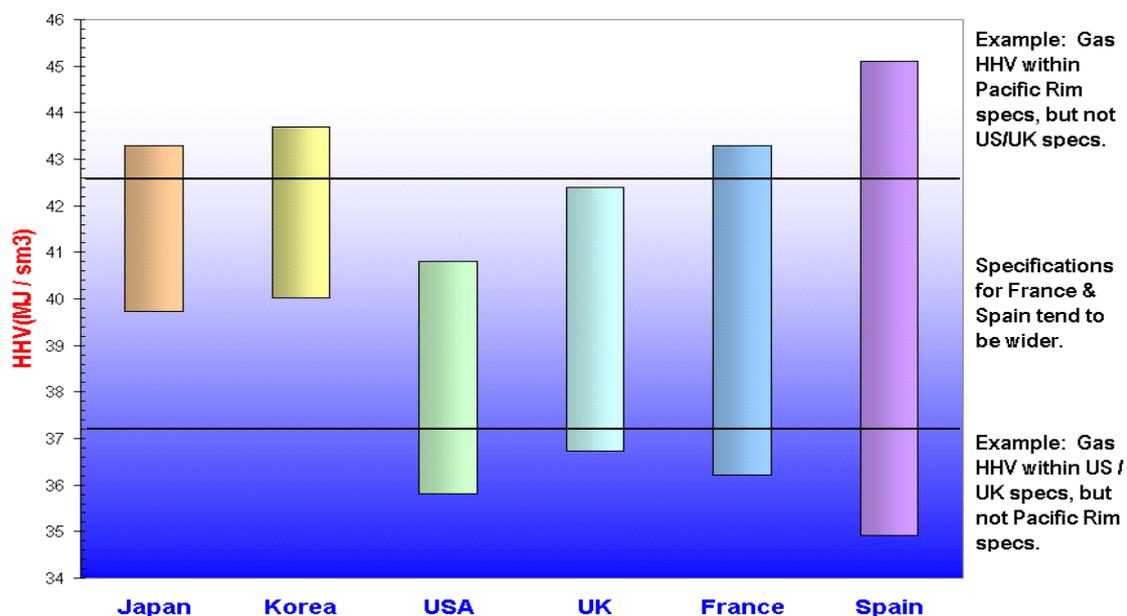

Figure 1 Worldwide Heating Value Specifications

In Europe a gas interchangeability index (the Wobbe Index) is also specified. The Wobbe Index is defined as HHV/ √Gas Specific Gravity (relative to Air) and is often more stringent than the heating value specification. British Standard 7859 provides all the details on standard conditions, component heating values, ideal vs. non-ideal gas calculations and more. The Wobbe Index is not a completely accurate way to predict the behavior of burners using alternate gases, but it is convenient and accurate enough for most purposes.

Generally speaking, a gas in the preferred ranges of HHV and Wobbe Index will burn without combustion problems such as flame lifting, flashback, excessive NOx and CO, or yellow tipping. Figure 2 below shows the preferred Wobbe Index region and the consequences of operating in objectionable ranges.
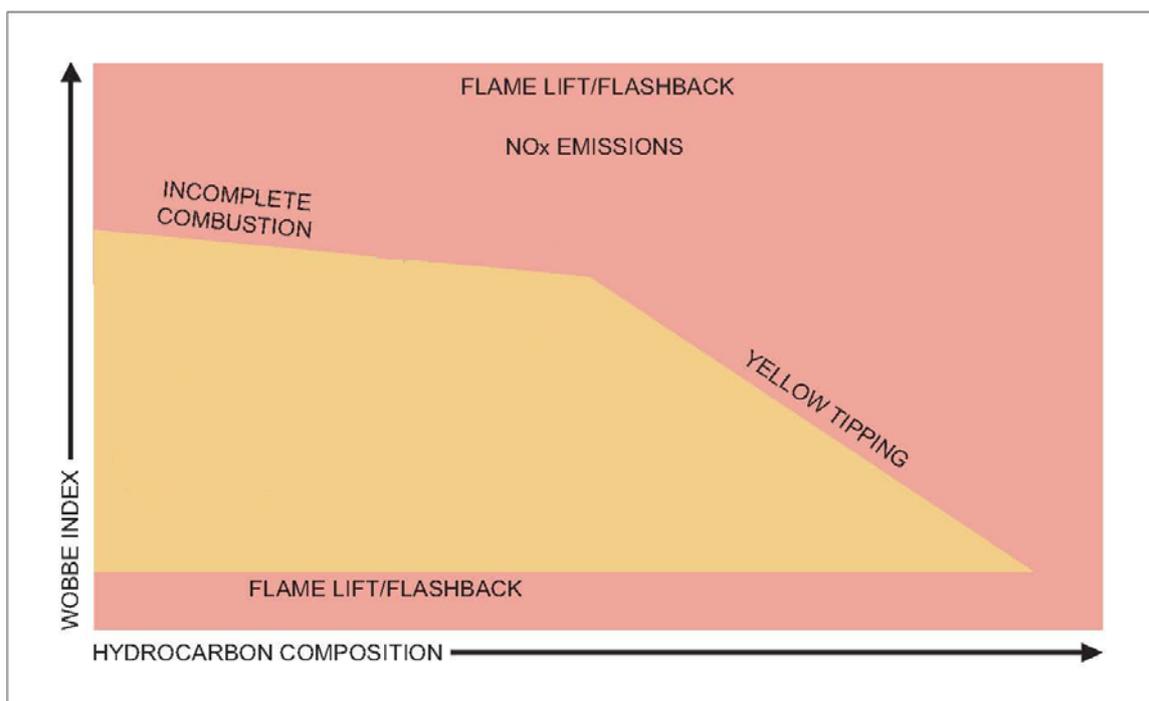
Source: National Petroleum Council

Figure 2 Combustion Properties as a Function of HHV and Wobbe Index

The LNG market is changing because of two players in the marketplace: the US and UK. Ironically both new players were in the LNG business in the early to mid 1970's but a regulatory crisis in the US and North Sea discoveries in the UK resulted in both countries receiving only minimal quantities of LNG. Now the US is projecting a widening gap between gas production and demand that can only be satisfied with LNG imports projected to grow to 90 million metric tons per year (MTPY) by the year 2025 [2].

The UK interconnecting pipelines once exported gas from the UK to the continent, but recently the flow has reversed direction. UK gas imports via pipeline and LNG are now expected for the long term, and three LNG import facilities are now under construction with a combined capacity of about 16 MTPY [3]. LNG capacity increases via expansions of these projects or new projects should provide an additional 8 MTPY by 2013.

Clearly the US and UK demands will have quite an impact on the Atlantic Basin market which is currently at about 49 MTPY. The size of the US and UK markets are more than large enough to target, and dealing with gas quality specs is a requirement for participation. This is especially true in the Middle East where cargoes can go to either Pacific or Atlantic consumers. There are many options for modifying the quality of imported or exported LNG on both ends of the trade, and all parties involved have a high level of interest finding the best solutions for their own situations.

The specifications in the US remain somewhat fluid as there is no national standard for heating value or Wobbe Index. In March 2005 the Natural Gas Council (NGC) issued interchangeability recommendations to the Federal Energy Regulatory Commission (FERC) that included a Wobbe Index maximum of 1400 Btu/SCF (52.1 MJ/Sm^3). FERC made an announcement on 15 June 2006 where they recommended following the NGC

criteria on a voluntary basis. While this avoids some disruption of trade there is still the effect of market uncertainty; who takes the risk on quality if there is a chance of mandated requirements in the future?

Natural gas specifications were covered in good detail by Bramoulle, Morin, and Capellethe of Total in their paper "LNG Quality and Market Flexibility Challenges and Solutions" (Reference 1). This paper also covered adjustments in gas quality from a global perspective of the LNG chain. Since these topics have been well covered previously, the discussion below will place more emphasis on the impact interchangeability has on the liquefaction and receiving terminals and the thought process a plant designer goes through to select a solution.

LIQUEFACTION END SOLUTIONS

Modifying heating value at the liquefaction end usually means adding or extracting ethane, propane and butane (LPG), though nitrogen may also play a part. For natural gas supplies rich in LPG components in the Atlantic Basin, a lower high heating value (HHV) is preferred if the US and UK markets are to be the consumers. On the other hand, Pacific Rim consumers prefer a gas with increased HHV, and Pacific Rim sources that are lean in LPG components may require upward HHV adjustment. Natural gas sourced in the Middle East can go either direction to the Pacific or Atlantic markets, which raises the possibility of producing two product qualities.

Reducing HHV

The two methods available for reducing HHV on the liquefaction end are LPG extraction and adding nitrogen, as discussed below.

Plant Capacity and LPG Extraction. One of the first considerations a contractor faces when LPG extraction becomes a requirement is the effect the extraction has on the plant capacity. Because of the extracted LPG, the LNG production drops unless more power and capacity is added to the refrigeration systems, and this is one of the reasons published $/ton of LNG can be misleading. The cost of extracting LPG can be significant and yet the product is often not considered in published $/ton/yr numbers.

Table 1 shows a typical natural gas before and after extraction, based on extracting propane and butane products with the recoveries shown and the results are essentially independent of the selected liquefaction process technology. The comparison holds the LNG product rate constant on a tonnage basis. With LPG extraction the total product rate (LNG, propane, and butane combined) increases by 7.7% while total refrigeration power increases by almost 10%. Power thus increases at a greater rate than total product, indicating a greater energy requirement for extracting and fractionating LPG.

Table 1 - Effects of LPG Extraction

		Without LPG Extraction	With LPG Extraction	Delta %
LNG Rate	MMTPY	4.6	4.6	0.0%
Propane Product Rate	MMTPY	0	0.21	
Butane Product Rate	MMTPY	0	0.17	
Total Product Rate	MMTPY	4.6	5.0	7.7%
LNG High Heating Value	MJ/Sm3	42.4	40.6	-4.6%
LNG Wobbe Index	MJ/Sm3	52.9	51.9	-2.0%
Nominal Compression Power	MW	Base	Base + 9.8%	9.8%
Specific Energy	MW/MMTPY LNG	Base	Base + 9.8%	9.8%
	MW/MMTPY LNG + LPG	Base	Base + 2.3%	2.3%

Basis: Feed Composition (%)

Nitrogen	6
Methane	83
Ethane	7
Propane	2
Butane	1
C5+	1
Total	100

Propane/Butane Recovery in Extraction Case = 66% and 83% respectively
Nitrogen Remaining in LNG = 1 mol %
HHV and Wobbe Index calculated on ideal gas basis

Just to make a point, if the cost of liquefaction is assumed to be linear with power then the $/ton of LNG goes up by the same percentage as the specific power. The $/ton of LNG then appears 10% higher for the plant that extracts LPG even though the benefit of LPG product sales is hidden.

For this reason it is more appropriate to use $/ton of total product including LPG for comparisons to overcome some of the inherent differences between plants with and without extraction. However most published figures on plant costs and capacities only provide the LNG product for reference. A similar thought process should also be applied to specific energy which shows only a 2% difference when based on total product compared to the 10% difference obtained using only LNG production.

Table 1 also shows the magnitude that product properties change as LPG products are extracted. The HHV drops by 1.8 MJ/Sm^3 while Wobbe Index drops by only 1.0 MJ/Sm^3. The desired Wobbe Index often controls the extent of extraction required and the index is not as sensitive to extraction as heating value (however, as discussed later Wobbe Index is *more* sensitive than heating value to nitrogen injection).

Another factor a design engineer considers is the optimum amount of LPG extraction. Natural gas specifications in the design basis are usually given as a range with maximum and minimum HHV. It is possible to target a lower HHV limit with deeper extraction, or a higher HHV by allowing more LPG to remain. Usually the LPG netback to the exporter has a higher value as separate LPG products as opposed to components in the LNG product and in such cases the optimum design tends towards deeper extraction and a lower HHV LNG. Also, once LPG extraction is part of the design, there are many more facilities required for the plant, such as LPG storage and loading; so higher extraction tends to be required to justify the extra cost of these new facilities.

If we extract 95% of the propane and 100% of the butane in the above example, the HHV becomes 40.0 MJ/Sm^3 and the Wobbe Index becomes 51.6 MJ/Sm^3. If lower values are needed then ethane must also be extracted. Ethane extraction has its own difficulties and is discussed later.

Process Options for LPG Extraction. If LPG extraction is necessary several processing options exist. The best process for a given application depends on factors such as the feedstock composition, the degree of extraction needed, and capacity. The more basic process options are as follows:

LPG Recycle. The principle behind recycling LPG is similar to that of the lean oil plants used in gas processing; adding a cold liquid to a natural gas stream absorbs lighter components and increases the amount of recovery beyond what could be achieved solely by chilling. In LNG plants the most common "lean oil" is butane, but in some cases propane and even ethane can be recycled to increase LPG recoveries. Recycling the natural gasoline stream is not done because of the possibility of sending heavy hydrocarbons into the cryogenic exchangers and freezing as a result.

A typical LPG recycle flow diagram is shown in Figure 3. The advantage of LPG recycling is that the equipment configuration changes very little from a liquefaction process without LPG extraction. The LPG extraction process is essentially integrated with liquefaction, and it is not difficult to design for operation in both extraction and non-extraction modes. It is often possible to retrofit an existing plant for extraction even if the original plant did not include extraction, depending on the size of the fractionation equipment.

Figure 3 - LPG Recycle Process

The LPG recycle process is effective up to a point, but if deep extraction is required then the turbo-expander process costs less. If 60-70% of the propane must be extracted LPG recycle works well, but if 80% or more of the propane must be extracted then the turbo-expander process should be considered.

Turbo-expander extraction. A typical turbo-expander extraction process is shown in Figure 4. There are many variations possible, but the key feature is a dynamic expansion of the natural gas to produce liquids followed by re-compression. Thermodynamically, the expansion follows an isentropic path which makes both high recoveries and high efficiencies possible. This is in contrast to static expansion through a valve which follows a less efficient isenthalpic path.
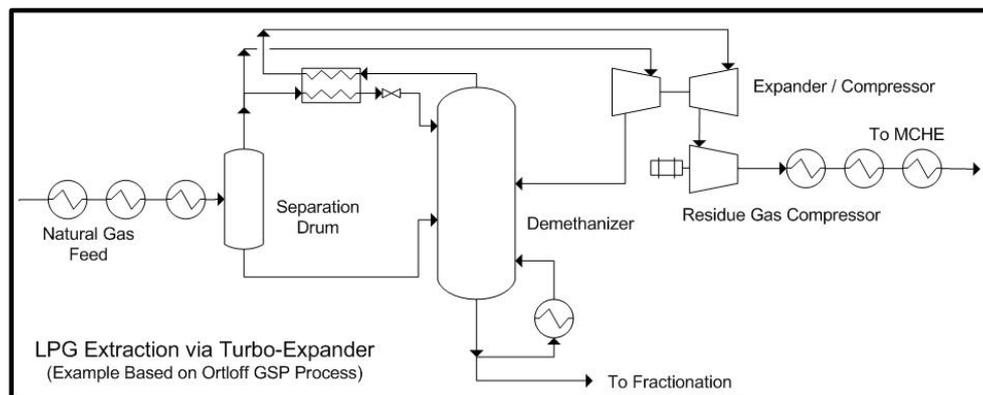
Figure 4 - Turbo-Expander Process

One objective of the process design engineer is to make the process flexible enough to operate without the expander if the expander or residue gas compressor is down for planned or unplanned maintenance. The goal in this case is to make on-spec product at reduced capacity because operators can run at 5-10% less capacity for a short time. Most operators cannot tolerate running down off-spec product even for a limited time when there is a possibility of having a full storage tank of LNG that cannot be shipped to the intended customer.

Another requirement is that the heavies removal is sufficient in J-T mode to prevent freezing of heavies in the main exchanger.

Scrub Column Modifications. One more method to increase LPG recovery is to lower temperatures in the scrub column [4]. This can be done by increasing reflux and reducing (or eliminating) the scrub column reboiler. Using a lower temperature refrigerant (such as using mixed refrigerant instead of propane) in a modified scrub column condenser increases the reflux.

Eliminating the reboiler sends a greatly increased amount of methane to fractionation, and in most cases a demethanizer is required to separate methane from ethane as shown in Figure 5.
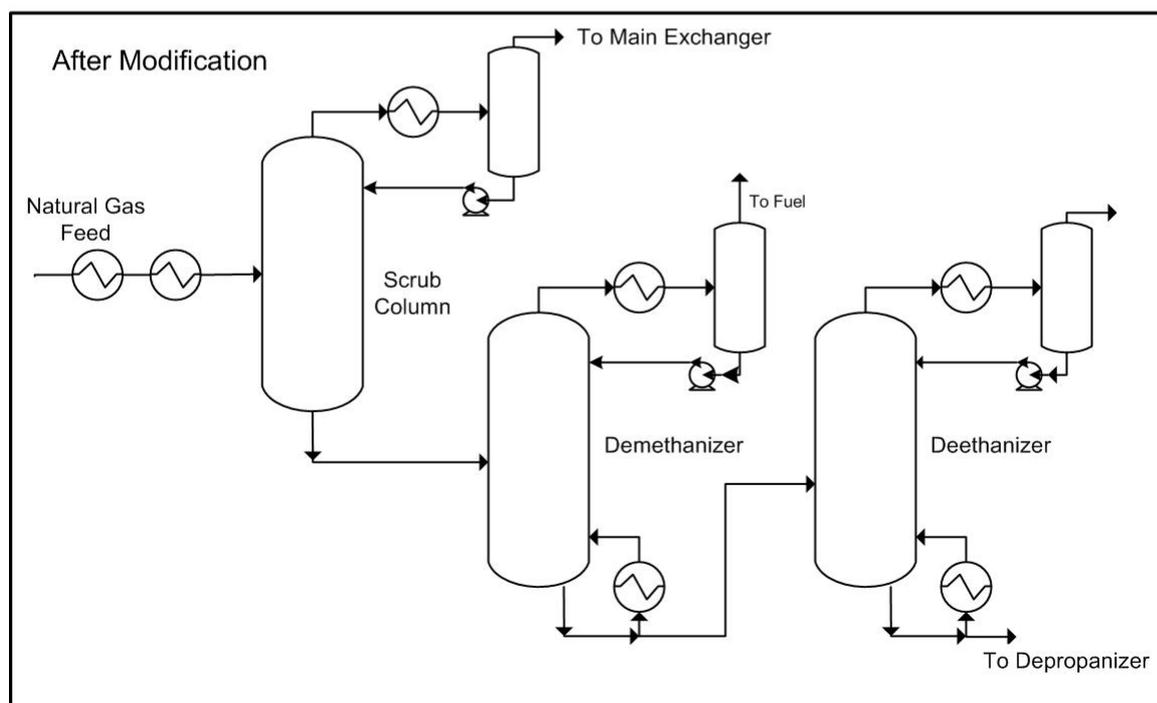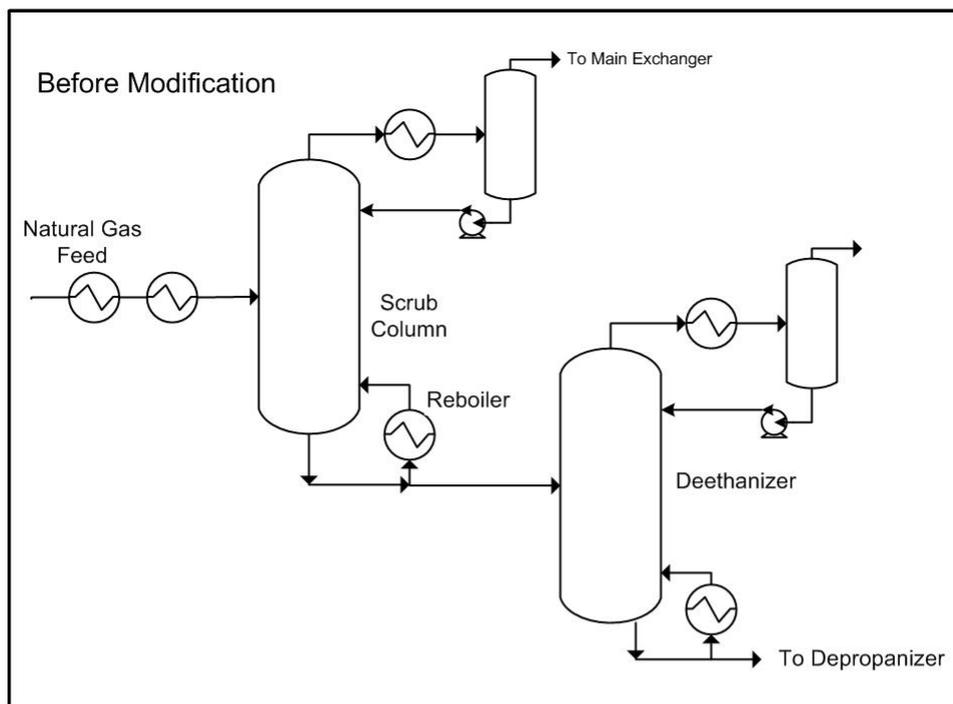
LPG Extraction Via
Scrub Column Modification

Figure 5 - Scrub Column Modification

All three of the methods above may be applied to existing plants, with various degrees of modifications required. The turbo-expander option may actually involve very little change to the existing equipment since it can function similarly to a "saddle" plant on a gas pipeline. Such an arrangement is sometimes called "front end

LPG recovery" where the LPG is extracted upstream and essentially independent of the liquefaction unit. The natural gas diverts to the expander plant and then returns to the LNG train at the same conditions less the extracted LPG. Similarly, the LPG recycle arrangement is sometimes called "integrated LPG recovery" because the recovery takes place within the liquefaction unit.

With front end LPG recovery the capacity of the existing fractionation unit can be fully utilized by design, with fractionation columns in the new extraction unit overcoming existing fractionation capacity limits. The LPG recycle concept on the other hand is likely to require significant modifications or equipment replacements in the fractionation unit to handle the increased hydrocarbon liquid flows. Scrub column modifications also require additional equipment and modifications, especially if the existing design has no demethanizer.

Depending on the basis of design and the capabilities of existing equipment, rotating equipment can also change significantly. As shown in the above example, at constant LNG production the total compression power increased by almost 10%. Accommodating this power increase generally requires compressor and driver modifications as a minimum and in many cases new machinery.

Nitrogen. Another way to reduce heating value is to add nitrogen. In fact, if achieving Wobbe Index is the goal, adding nitrogen can be more effective than extracting LPG. Typical US pipeline specifications allow for a maximum of 3 mol % nitrogen.

At many sites the nitrogen is not "added." The nitrogen is present in the natural gas and in some cases the concentration is over 5 mol %. To keep the nitrogen with the LNG it is only necessary to chill the LNG product colder, but this does take power. The power is significant: on a constant tons/hr basis, it takes roughly 4% more power for a 1 mol % increase in nitrogen concentration.

The table below shows the effect of adding nitrogen compared to extracting LPG which is also shown graphically in Figure 6.

Table 2 – LPG Extraction vs. Nitrogen Injection

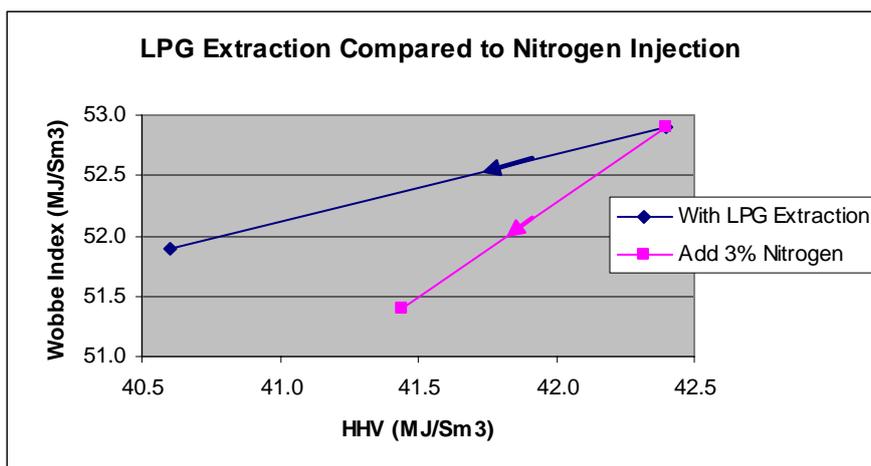|		Unadjusted	With LPG Extraction	Add 3% Nitrogen
LNG High Heating Value	MJ/Sm3	42.4	40.6	41.4
LNG Wobbe Index	MJ/Sm3	52.9	51.9	51.4

Figure 6 - LPG Extraction vs. Nitrogen Injection

Adding 3% nitrogen (instead of extracting LPG) does not change heating value as much as extraction however the nitrogen has a greater effect on Wobbe Index. Since Wobbe Index is often a more stringent requirement than heating value, increasing nitrogen content may be a good way to meet specifications in some cases.

When considering using nitrogen on the liquefaction end to meet natural gas specs, a design engineer must confirm that the shipping and receiving end can take a high nitrogen LNG because the boil off gas from a 3 mol % nitrogen LNG contains over 50 mol % nitrogen. Also, the receiving terminal might have to send out boil off gas that meets the pipeline specifications prior to recondensing in the LNG send out. In general the nitrogen solution is handled at the receiving end.

Handling Ethane. One of the biggest challenges on the liquefaction end can be the disposition of ethane. The UK specifications (and California specifications in the US) have the effect of limiting the ethane to as little as 6 mol % in the LNG. If the natural gas feed to the LNG plant contains 8-9% ethane then ethane must be used within the plant or exported.

The problem with exporting is that the liquefaction plants tend to be in remote locations where there is little industrial demand for ethane. The value of the ethane usually does not support the cost of the extraction.

Using ethane as fuel within the plant is not a trivial matter; beyond possible problems with excess fuel, there are also limits on how much ethane can be contained in gas turbine fuel while still meeting environmental restrictions.

The amount of fuel an LNG plant requires depends on many factors such as process efficiency, compressor and driver efficiency, and site specific factors such as domestic gas or electric power exports. However for the sake of argument, we can assume a plant requires 10% of the heating value contained in the feed to be consumed as fuel.

Also assume for this example that the plant has a feedstock containing 9 mol % ethane, and the LNG product must be 6 mol % ethane. The resulting fuel composition would be 64% methane, 36% ethane, with a high heating value of 47.90 MJ/Sm^3.

Fuel containing 36% ethane may be a problem depending on the type of gas turbines used for compressor drivers and NOx emissions requirements. If NOx emissions up to 160 ppm are allowed then Frame 5 gas turbines are capable of burning this amount of ethane using lean head combustors. Frame 6 gas turbines can also be designed to meet this requirement with a high ethane fuel.

Frame 7 gas turbines, on the other hand, require dry low NOx burners and methane must be at least 85% of the reactive material in the fuel gas. This precludes burning a 36% ethane stream in the Frame 7 gas turbines. However, power generators and other process services are often driven with Frame 5 or Frame 6 gas turbines, and segregating the fuel system into high and low ethane systems can help to achieve the amount of ethane "disposal" required while meeting environmental restrictions.

In some cases an ethylene plant may be considered as a solution to the ethane problem. This would be a last resort option because operating an ethylene plant adds such significant scope to the simple objective of transporting natural gas. The numbers work out well on paper at least; in our example if *all* of the ethane is extracted from the nominal 4.5 MTPY LNG train feed the ethylene plant capacity would be about 1.3 billion lbs. per year, which is a competitive world scale plant.

Transportation of the ethylene is a problem however, and most owners would also build a polymer plant to convert the ethylene into a product more easily shipped. Although ethylene is moved by ships in some cases, the ships are expensive and the ethylene becomes uncompetitive because of transportation costs.

Operations of the ethylene plant would be dependent on the LNG plant for feedstock and this would require a cooperative environment between the upstream LNG plant and downstream chemical plant operations. Ethylene plant operators sometimes run 4-5 years before turn-around for maintenance, while the LNG plants have gas turbine maintenance requirements more frequently (typically annually).

Such inherent differences in technologies are not typically close-coupled together and this would certainly put the "cooperative environment" to the test. The ethylene and LNG parties would both have to face technical realities while responding to production pressures.

Given all the challenges of the ethylene plant option it does not seem likely to be a common solution for future plants. However, in Skikda, Algeria Sonatrach has an ethylene plant with ethane feedstock extracted from LNG trains. The concept has worked at that location at least, and perhaps there are other niches to be discovered in the future.

D. Chretien of Total has proposed another alternative at the World Gas 2006 Conference which converts ethane to methane. The process begins with steam reforming ethane to H_2 and CO, followed by catalytic conversion to methane. For every four molecules of ethane seven molecules of methane are produced, along with water and CO_2: ($4 C_2H_6 + 9H_2O \rightarrow 21 H_2 + 7 CO + CO_2 \rightarrow 7 CH_4 + 7 H_2O + CO_2$). The conversion process is called CRG for "Catalytic Rich Gas" conversion which is commercially proven and licensed by Davy Process Technology [5].

As with the ethylene option, the cost is significant and the investment may not pay out on a localized basis. However, if there is a need to remove ethane to make the LNG product marketable this process has the advantage of eliminating by-product streams and the cost associated with purifying and storing them (propane, butane and heavier hydrocarbons can also be converted, but hydrocarbon efficiency drops as the feedstock gets heavier).

Although projects are proceeding with processes modified to extract ethane for export or as fuel within the plant, there are also projects on the receiving end where ethane will be extracted. It is probably a safe bet that ethane extraction will be done on both ends for the foreseeable future.

Increasing HHV

At some locations very lean gas is available for export to Japan or Korea. Historically this trade began with Alaskan LNG with heating value adjusted upwards on the receiving end by way of LPG injection. Currently there are several projects in the Pacific Rim under development with relatively lean feedstocks. If the LNG product does not meet minimum heating value specifications LPG must be injected on the liquefaction or receiving end.

Looking at the problem globally, it would seem the best technical solution is LPG injection at the receiving end. The shipping temperatures of propane and butane are at least as warm as -40 degrees C and thus do not require expensive special materials such as 9% Ni, aluminum or stainless steel. All things being equal, it would cost less to ship and store LPG separately from the LNG, until the LPG is injected into the natural gas send out stream. If the LPG is injected at the liquefaction end then the LPG is transported in special materials suitable for cryogenic service. At the same time, land constraints on the receiving end may still lead to a solution where LPG is injected into the LNG prior to export.

If LPG must be purchased and injected at the liquefaction location the first challenge is finding a local source with sufficient quality and then installing the facilities for unloading and storage. Then more refrigeration is required because the LPG must be chilled from -40 degrees C (in the case of propane) down to -160 degrees C. This requires energy which will be applied as additional boil off gas compression or refrigeration compression. If the LPG is injected at -40 degrees C into the LNG then the LPG chilling is accomplished by vaporizing methane. The vaporized methane must be compressed to high enough pressure to be used as fuel or material for the natural gas circuit of the liquefaction process.

Simply injecting -40 degrees C LPG into the LNG is equivalent to pure component refrigeration with methane at -160 degrees C in a single stage, which is not very efficient. A better approach is to stage the injection into liquid methane at several pressure levels, or better yet use refrigeration to chill the LPG to LNG temperatures prior to injection. This can be easily accomplished by adding an LPG injection pass to the MCHE, or adding another exchanger in parallel to the MCHE, using liquefaction level refrigerant to provide the necessary cooling duty.

While there are technical problems and solutions for LPG injection, the commercial side is not as easy to overcome. The refrigeration for chilling the LPG to LNG temperatures is expensive and the LPG itself generally sells for more as a separate product than as a component of LNG. Add to this the more expensive transportation when the LPG is shipped at -160 degrees C, and the prospects for purchasing LPG and injecting on the liquefaction end do not look favorable.

Multiple Products

At some facilities there may be a desire to market multiple products, such as a high HHV product and a low HHV product. For a liquefaction plant with a lean feed gas this might be accomplished by importing LPG and injecting the LPG on a cargo by cargo basis. The plant operator runs the LPG injection equipment for a cargo destined for a high HHV market, but the equipment remains idle when loading a ship headed for a low HHV market. In theory it is possible to chill the LPG prior to injection using liquefaction refrigeration as mentioned above, but every loading would significantly change operations within the train. For this reason it is more likely that actual designs would have a stand-alone refrigeration unit, or would rely on greater boil off gas compressor capacity.

An important technical feature of the design is to avoid sudden vaporization of the LNG as the LPG is injected. This can be achieved by pumping the LNG to higher pressure before using the LNG to chill the LPG.

If the opposite situation exists where a plant with significant LPG in the feed is considered for both low and high HHV markets, the LPG can be extracted and stored or exported until a high HHV cargo is needed. At that time the stored LPG is injected during loading, but this concept has the same disadvantages mentioned earlier – expensive LPG transport and diluted LPG product value.

Perhaps the most promising way to accommodate multiple products is based on a large facility where two separate products can be produced in different trains with the two products stored separately. This becomes more convenient if two berths are also present, each with their own loading lines. The disadvantage of this method is the extra storage capacity that would have to be installed compared to a single product facility.

A plant producing multiple products will cost more than a plant marketing a single product. The multiple product plant may still be competitive if it is closer to the consumer or has economy of scale advantages over single product plants in the same market.

RECEIVING END SOLUTIONS

As mentioned in the previous discussions, there are certain situations which make adjustment on the receiving end of the LNG chain attractive. As with the liquefaction end, the selected solutions can depend on whether the facility is new or a modification to an existing plant. For example, in existing receiving terminals "grandfathering" is an issue, and minor modifications to lines containing LNG raise the possibility that the entire facility will have to be brought up to current code revisions. Such factors obviously have a large bearing on the process selection.

Reduce HHV

HHV can be reduced either by adding inert nitrogen or by extracting LPG. Typical US pipeline specifications limit the amount of inert material to 3 mol %, thus adding inert material is limited to 0.9 – 1.2 MJ/Sm^3 (25-35 Btu/SCF) heating value reduction depending on the amount of nitrogen in the LNG as it is received. However, if this is the only adjustment needed the process is relatively simple and there are no other products to deal with besides LNG send out.

There are also consumers on the US east coast and in the UK who can take up to 5 mol % nitrogen which makes greater HHV adjustment by adding inert material possible.

Add Nitrogen. Adding nitrogen to the natural gas product would seem to be a straight forward proposition, and yet there are a myriad of options. The first option is the method for generating the nitrogen, and the following are possibilities:

- Membranes
- Pressure Swing Adsorption (PSA)
- De-oxygenation of Submerged Combustion Vaporizer flue gas using hydrogen
- Inert Gas Production by Combustion
- Cryogenic Fractionation

The suitability of these processes depends to a great extent on the capacity requirement and pipeline gas specification. Some pipeline specifications limit oxygen concentration to only 1 ppm which effectively eliminates all but the last option from consideration. Furthermore, if the design send out rate is one billion standard cubic feet per day (1 BCFD) and the nitrogen addition is 3% of that, the 1,250,000 SCF per hour nitrogen plant capacity makes cryogenic air separation the only practical choice (though inert gas by combustion can compete at this capacity if oxygen content of up to about 50 ppm is allowed in the send out).

Once the nitrogen is generated, injection into the natural gas stream can be done in a straight forward manner by compressing to pipeline pressure and then mixing the nitrogen and natural gas streams together. This simple approach has the advantage of not requiring modification to lines containing cryogenic material which may be important in a grandfathering situation. The disadvantage is that there may be another process option requiring less power.

LNG receiving terminals typically take advantage of the LNG cold temperature sink as a way to reduce boil off gas compression. The LNG is pumped out of the storage tank to an intermediate pressure, which puts the LNG in subcooled state. It is then possible to add heat by "recondensing" boil off gas where the boil off gas is mixed with the LNG and the combined mixture is a single liquid phase; the higher pressure allows the boil off gas to be recondensed back into the LNG and the LNG remains in the liquid state (though at a higher temperature).

At many terminals the amount of LNG cold is more than enough to recondense the boil off gas, and in such cases there is cold available for recondensing at least some of the nitrogen to be added. The recondensed nitrogen is then pumped to high pressure with the LNG rather than being compressed and thus requires less power.

Separating air requires power. Based on an ambient temperature of 22 degrees C and adding 3% nitrogen to a pure methane send out, air separation and compressing the nitrogen to pipeline pressure requires roughly 0.25-0.3 % of the natural gas send out as fuel. This amount of fuel is relatively minor, given that a rule of thumb for terminals with submerged combustion vaporizers is 1.8 % of the send out required as fuel.

The LNG cold can also be used to assist the air separation process, and a typical concept is to cool air using a non-flammable intermediate fluid. This concept can reduce the energy consumption somewhat and reduce cost. One thing to remember, though, when exploring gas quality modifications on the receiving end is that receiving terminals are usually expected to operate 365 days per year and send out flow rates can also vary widely. As a result, instead of one large air separation unit it may be necessary to have two 50% units running and one identical unit as a spare. Adding a spare unit and losing some economy of scale increases cost.

Extract LPG. Like adding nitrogen, LPG extraction also offers many options, and selecting between them depends on the amount of extraction desired, whether there is a higher than ambient temperature heat source available, and the economics of compression vs. pumping. With LPG extraction LPG product is likely to have a high enough value to support the additional equipment and operations cost.

Once the decision is made to extract LPG the question is then how much to extract. It is possible to limit the extraction to only what is needed to meet the natural gas pipeline specs, or it is possible to extract more to increase LPG product sales. Generally speaking, if any extraction at all is required the tendency is to extract as much as possible.

As with the liquefaction end, ethane may also present challenges if heating value or Wobbe Index is too high after extracting all the propane and butane. Hopefully there is a consumer in the vicinity to take the ethane product via pipeline. This scenario is currently more likely on the US Gulf coast than on the east and west coasts, thus no general solution are available. If no consumer is available then the owner will have to use other measures such as burning the ethane as gas turbine generator fuel to produce electric power.

If ethane extraction is not required, a process such as the one shown in Figure 7 will extract propane and butane to low concentration levels in the LNG. This process is also capable of extracting up to 80-85% of the ethane. It is possible to extract over 95% of the ethane by adding a condenser to the recovery column.

There are many commercial offerings for receiving terminal LPG extraction by companies such as Black and Veatch, Foster Wheeler and Ortloff. The different technologies available are discussed in more detail in "Selecting the Optimal Process for C_2+ Extraction from LNG" by Huang, et al [6]. Selecting between the options usually involves studies and a selection process [7].

Basically the processes can be divided into three different types based on the way the residue gas (the demethanizer overhead product stream) is handled: Residue Gas Compression, Residue Gas Compression and Condensing, and Residue Gas Condensing. Representative examples of each type are shown in Figures 7, 8, and 9 below.
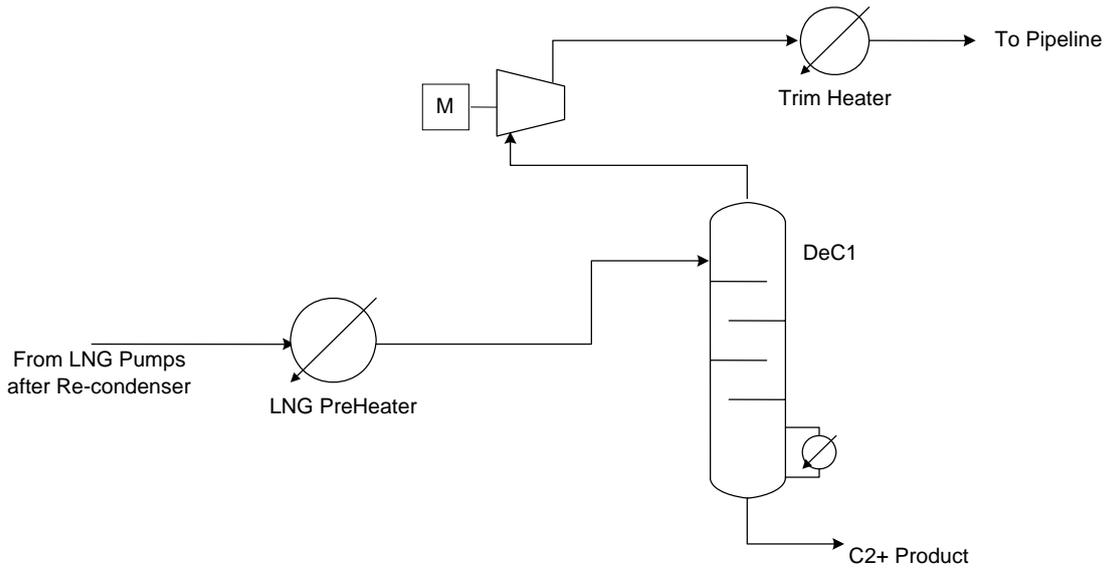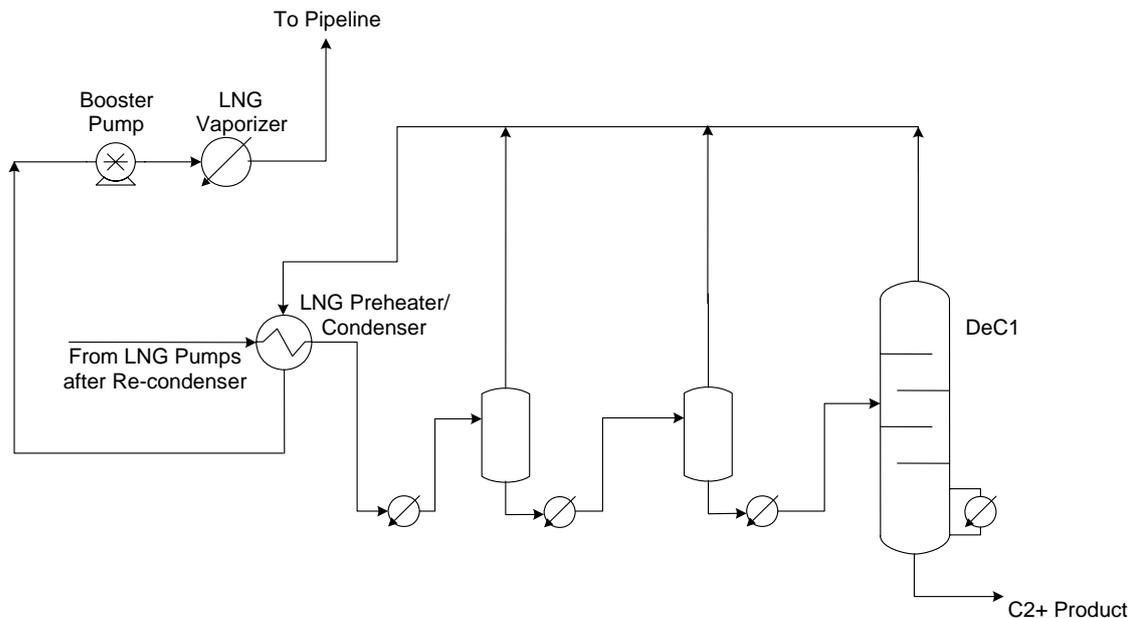
Residue Compression [Rambo et al., 1992]

Figure 7 – Residue Gas Compression
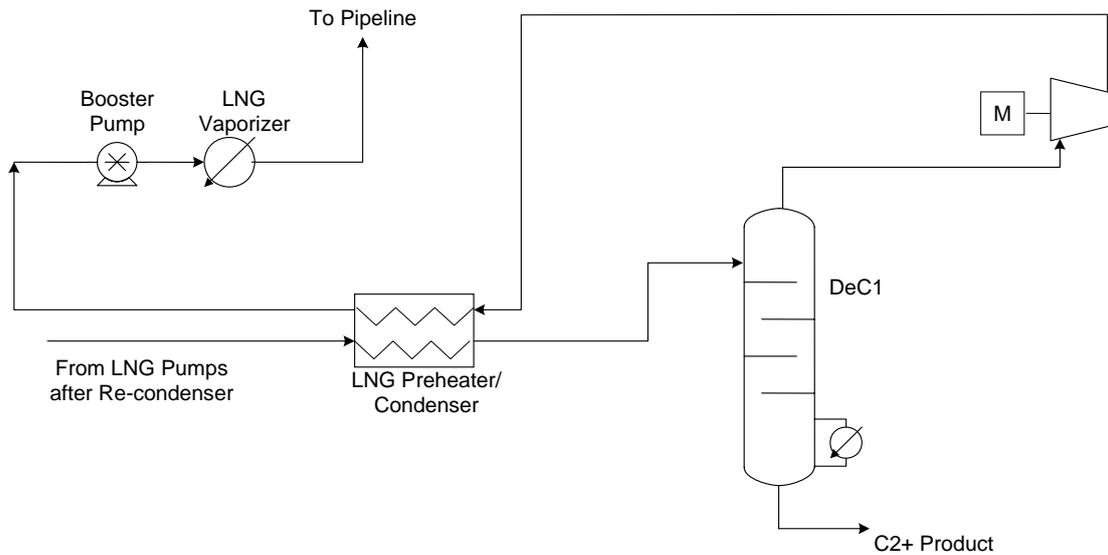
Residue Compression and Condensing,
Configuration 1 [McCartney, 2003]

Figure 8 – Residue Gas Compression and Condensing

Residue Condensing - Exchanger, Configuration 1
[KBR Design for Cove Point Receiving Terminal, 1972]

Figure 9 – Residue Gas Compression and Condensing

The optimal process for a given site is a trade off between cost and process complexity according to the following summary:

- Residue Compression is simple and easy to operate but requires expensive vapor compression.

- Residue Condensing followed by liquid LNG pumping and vaporization requires more complex process controls, however the expensive compression is eliminated.

- Several processes are available which are combinations of the two previously described. Some compression is required but not to the extent of the Residue Compression type. Some compression savings are achieved while still maintaining relatively simple operation and flexibility.

At some locations (such as the US Gulf Coast) it may be possible to tie into existing LPG pipelines to transport the extracted LPG products. If the LPG is transported by truck or ship then LPG storage will be required, and the storage inventory must be factored into the safety assessment. It may be necessary to acquire more land depending on the size of the storage and LPG spill criteria.

Increase HHV. Adding LPG is the only practical method increasing HHV, outside of blending with gas from another source. Adding LPG is relatively simple from a process standpoint since it can be pumped to pipeline pressure as a liquid. At pipeline pressure propane and butane are dense phase (as is the methane) and mixing the LPG with the natural gas is not difficult.

LPG injection is common in Japan where the heating value specification for natural gas is high to maximize utilization of the infrastructure. One source of LNG to Japan is Kenai, Alaska where the LNG is nearly pure methane. This LNG must have LPG injection for interchangeability purposes.

If received LNG has a high heating value of 1020 Btu/SCF that must be increased to 1070 Btu/SCF, it takes about five propane trucks (each truck carrying 3,000 gallons or 11.4 m^3) per billion SCF to achieve the required heating value. The resulting concentration of propane in this case is just over 3 mol %.

It is likely that some amount of LPG storage is needed to accommodate the LPG delivery package size and to handle possible transportation delays. As with the case where LPG is extracted and stored on site, the storage capacity and spill criteria may influence the plot space required.

COMMERCIAL SOLUTIONS TO TECHNICAL PROBLEMS

Imagine if a rich LNG (that is, one with a high HHV) is produced in Australia for a market in California where the specification requires a low HHV. At the same time a very lean LNG produced in Alaska is being shipped to Japan where LPG is injected to raise the heating value. Technically, it would be better for the Australian LNG to go to Japan and the Alaskan LNG to go to California. There would be alignment on LNG quality as well as shorter shipping distances, but with the usual long term contracts between buyer and seller it would be difficult to make this happen in spite of the incentives.

The different arrangements are shown in Figure 10 where the red lines indicate the first scenario and the blue dotted lines show the second case where the destinations are swapped, with the resulting shorter shipping distances. Given the growing spot market, perhaps this arrangement could be carried out occasionally based on cargoes that are available beyond the long term contract requirements. Eventually, if LNG really does become a commodity, this sort of deal could be done on a regular basis by selling cargoes FOB to an arbitrageur who would then find a suitable market and price for the delivery.

The Dubai Techno Park and Dubai Multi Commodities Centre (DMCC) announced in August of 2006 plans to build a massive storage facility. This facility may be a solution for some interchangeability challenges. Dubai has access to Pacific and Atlantic markets and the storage hub would provide some flexibility in quality blending and shipping that does not now exist.
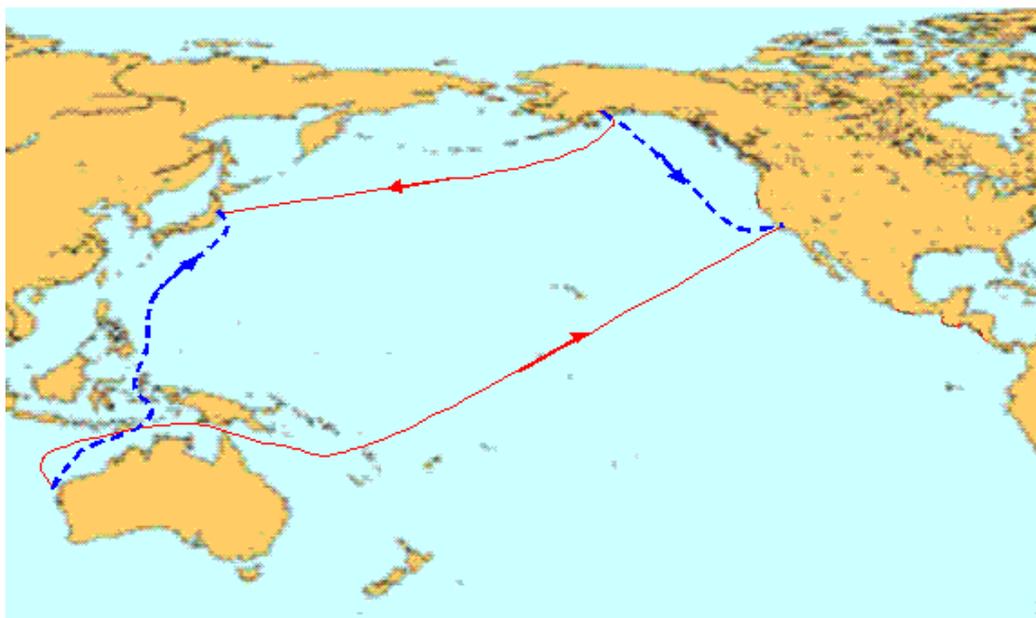
Figure 10 - Exploring Commercial Possibilities

CONCLUSIONS

Gas interchangeability is becoming a more important issue as the LNG market becomes more global. At the same time, it is very unlikely that the players in different markets could ever reach agreement on a uniform standard on a global basis. Indeed, the US Federal Energy and Regulatory Commission has decided, for the time being at least, that implementing a universal standard within the US would be disruptive to existing trade. This reality will continue to be a factor in plant designs as LNG buyers and sellers alike try to keep their options open.

- LPG extraction is already common and will continue to be a preferred method for adjusting LNG HHV downward both on the receiving and liquefaction ends.
- Nitrogen injection is also employed in existing terminals to reduce HHV and will probably continue to be implemented as a low cost method on the receiving end, especially when the controlling specification is Wobbe Index.
- LPG injection to increase HHV is more likely to be seen on the receiving end instead of liquefaction because of transportation costs (more expensive to ship with the cryogenic LNG).
- Ethane disposal on the liquefaction end will be a real challenge at some locations. If the ethane cannot be burned as fuel within the plant proven technical solutions, while expensive, do exist.

Finally, where there is enough economic incentive, it may be possible to swap cargoes and save on plant modifications and shipping. This involves more parties each with their own risk and reward needs, but after the nightmares of negotiations all parties may awaken to a bright day of enhanced returns on their investments.

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