

# **Maximising the Value of Surplus Ethane and Cost-effective Design to Handle Rich LNG**

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## **ABSTRACT**

Gas companies have strict requirements for the composition of the natural gas sent out from their LNG receiving terminals in order to satisfy customer requirements in terms of calorific value and quality of the composition. For LNG receiving terminals serving the North American market particularly, the C2+ content and heating value specifications are lower than most natural gases and most existing LNG baseload plants. Cost-effective management of these components enhances the LNG value chain, increasing flexibility in sourcing LNG cargoes. Several alternatives for managing C2+ components in the LNG value chain exist. Generally, extraction can occur at either the receiving terminal or upstream of the liquefaction plant. In many cases it is not cost-effective to reduce the C2 at the baseload plant, as it will reduce the LNG production rate and there may be no local market. Some LNG production plant locations, however, have grown to a size that would meet minimum economic scale for a steam cracker taking C2 and C3 as feedstock for ethylene and propylene production.

To accept cargoes of rich LNG, the LNG receiving terminal has to be able to manage the C2+ components in an economically efficient manner. This paper presents a cost-effective C2+ management process design for send-out gas at the LNG receiving terminal. The process couples Foster Wheeler's C2/C3 terminal extraction process with a pre-reformer that produces a low calorific "substitute natural gas", or SNG, for re-injecting into the LNG terminal send-out gas to control the heating value.

The case study demonstrates that the scheme is able to meet most gas specifications and is expected to be cost-effective relative to alternatives such as nitrogen injection. Key issues such as further reduction of hydrogen content from the pre-reformer are also addressed.

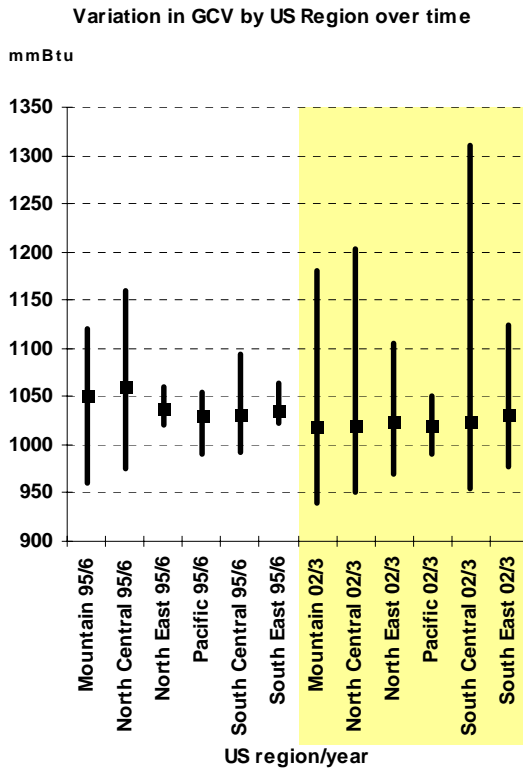
## INTRODUCTION

LNG receiving terminals must comply with strict requirements on send-out gas characteristics, including composition, calorific value, Wobbe Number and quality, in order to satisfy the customers' specific needs.

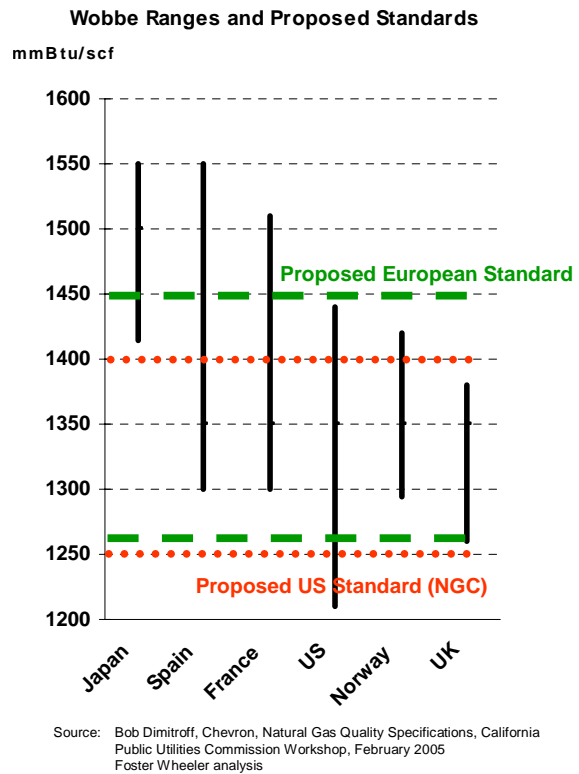
C2/C3 recovery potential depends on customer pipeline gas quality specifications. These are changing but the need for compliant gas qualities remains.

- Pipeline gas quality specifications have varied not only by country but in the case of the US, as shown in Figure 1, by region.
- The variability of gas input quality has increased over the past five years in every US region (Figure 1)
- The Natural Gas Council in the US is consequently proposing a change in standard, one that would also encompass some LNG imports
- In Europe, the increasing integration of pipeline networks is promoting similar convergence on a European standard (Figure 2)
- The proposed US standard does not overlap with the range of the Japanese standard, so it is likely that LNG will be produced in at least two different qualities (Figure 2)
- Although the European standard overlaps the Japanese standard, some Japanese LNG import terminals have been optimised for higher calorific value LNG than the upper bound of the European standard (Figure 2)
- It is likely that LNG will be produced at one quality for the Atlantic and Mediterranean basin and another for the Pacific basin.

**Figure 1**



**Figure 2**



Pipeline gas C2 and C3 specifications in the US are lower than most LNG baseload plants produce.

As the LNG buyers have multiple LNG specifications, there are several reasons why a reduction in the C2 and C3 content at the baseload LNG plant may not be desirable:

- Reduction in LNG production
- Additional compression equipment will be required
- Desirability to operate all LNG trains at the same conditions

## **PROCESS OPTIONS FOR TERMINAL OWNERS TO CONSIDER**

LNG terminal owners are currently focusing on ways to deal with these specification requirements.

For the reasons outlined, to accept cargoes of rich LNG, the receiving terminals must be able to manage the C2+ components economically. At the terminal there are four main options for meeting the export gas specifications:

1. Nitrogen and air dilution
2. Extraction of C2+ components to the terminal fuel usage
3. Extraction of C2+ components to local markets
4. Extraction of C2+ components and pre-reforming

## **C2+ EXTRACTION VS. NITROGEN INJECTION**

There are many benefits of reducing C2+ at the LNG receiving terminals, including:

- Flexibility to process LNG from multiple suppliers which allows terminal owners to maximize future LNG spot market purchases
- LNG cold may be effectively utilized.
- Flexibility in meeting multiple customer gas specification requirements.

Foster Wheeler presented an answer to this issue at the AIChE 2003 Spring National Meeting entitled “A cost-effective design for reducing C2+ at LNG receiving terminals” [1].

The design described reduces C2+ from the send-out gas in using LNG cold energy while minimizing or eliminating gas compression. The scheme utilizes low-temperature LNG as a cooling medium for the column overhead to achieve an economic separation. This requires no gas compression because LNG is used as a refrigerant in a direct-contact condenser. This condenses enriched C1 vapour from the column overhead to produce a lean LNG for liquid pumping. The design also uses existing vaporizers to vaporize column feed. One or more fractionation columns are used to separate heavy components from the LNG. The extracted C2+ components can either be exported or used as fuel for either power generation, if supplied on site, or for the submerged combustion vaporizers, if they are used.

Studies have shown that the process schemes for extracting C2+ from rich imported LNG are feasible, effective and economic. Building a C2+ or C3+ separation facility ensures that the LNG receiving terminal has the flexibility to receive various LNG feed compositions while meeting stringent calorific value export gas specifications.

Only minor modifications to the typical LNG receiving terminal equipment are required, including modification of the casing of the send-out pumps to a higher pressure rating, but no additional compression is required. This results in both lower capital and operating costs.

In order to minimise the capital and operating cost of the facility, the Foster Wheeler design processes the minimum amount of LNG from the LNG terminal to meet the send-out gas

specifications. The design also provides an option to by-pass the facility when the purchased LNG has met the send-out gas specifications. The highlights of the designs are:

- Using fractionation column(s) for C2+ reduction
- Using a gas direct contact condenser
- Using LNG as a cooling medium for condensing
- Pumping condensed LNG to gas send-out pressure for regasification
- No further compression requirements
- A cost-effective design to reduce C2 and heavier components

Nitrogen gas injection to the LNG terminal send-out gas is an option for Wobbe index control. However, it can prove costly (Table 1) in initial investment as well in operating cost. Most pipeline gas specifications have a maximum inert gas content of 3 mol% along with other limitations on the heavy components. As an example, California’s compressed natural gas (CNG) specification requires methane to be above 88-mol%, C2 to be less than 6 mol%, and C3 and higher to be less than 3 mol%. Consequently, injecting nitrogen gas alone is often not able to meet the pipeline gas specifications when presented with rich LNG. Injecting a small amount of air is an option; however, the oxygen content should be less than two-tenths of one percent (0.2%) by volume in the send-out gas after the injection. Consequently, in meeting the pipeline specifications, up to 1% of air will also be utilized to optimize the LNG calorific value reduction process.

<b>Table 1 - Case Study for Adjusting LNG Properties for a 1,000 MMSCFD LNG Receiving Terminal</b>				
	Scheme Case	Conditioned Gas Quality	Estimated Installed Cost (\$MM)	Annual OPEX (Delta)
I	FW Proposed C2 Reduction Scheme I	<ul style="list-style-type: none"> <li>• Reduce C2 in LNG from 9 mol% to 6 mol%</li> <li>• Heating value from 1103 to 1058 Btu/SCF</li> </ul>	35	\$0.0 MM (Reboiler heat duties are part of LNG vaporizer duty, 29 MW)
II	FW Proposed C2 Reduction Scheme II	<ul style="list-style-type: none"> <li>• Reduce C2 in LNG from 6.06 mol% to 5.20 mol%</li> <li>• Heating value from 1059 to 1050 Btu/SCF</li> </ul>	10	\$0.0 MM (Reference Case) (Reboiler heat duty is part of LNG vaporizer duty 3.9 MW)
III	N <sub>2</sub> Injection Scheme By Inject N <sub>2</sub> 0.8 mol% in LNG gas	<ul style="list-style-type: none"> <li>• Reduce C2 in LNG from 6.06 mol% to 6.00 mol%</li> <li>• Heating value from 1059 to 1050 Btu/SCF</li> </ul>	34	\$1.75 MM (Electric power 5 MW @ 4¢/kW.hr)

## MARKETING AND HANDLING OF EXTRACTED C<sub>2</sub> AND HEAVIER COMPONENTS

After part of the extracted C<sub>2</sub> and heavier components is consumed as fuel in the LNG terminal, the remainder may have a potential market.

In some locations there is an economic market for LPG (C<sub>3</sub> separated from C<sub>2</sub>), but a combined C<sub>2</sub>/C<sub>3</sub> feed to a cracker may be more attractive.

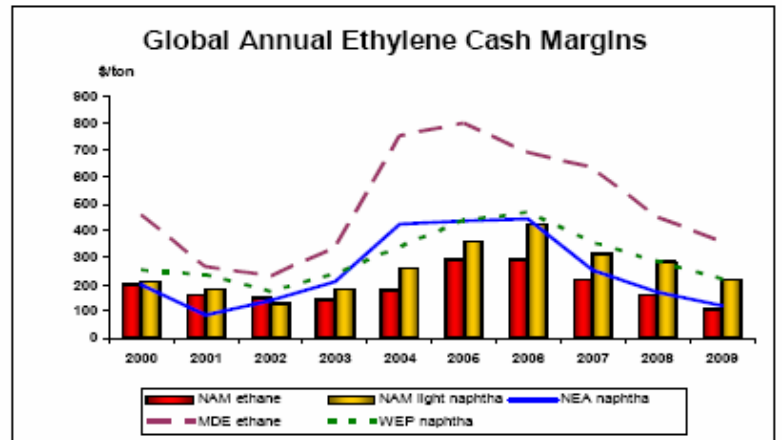
- Market values for LPG are given in Figure 3
- New ethylene production will tend to be based on steam cracking of ethane rather than naphtha cracking (Figure 4). This reduces the relative amount of propylene produced (Fig 5).
- A combined C<sub>2</sub>/C<sub>3</sub> feed would be economically more attractive as long as there are few indications of a balancing increase in on-purpose propylene production
- Figure 6 shows yields of the alternative on-purpose propylene processes
- A key indicator will be a significant increase in the number of propane dehydrogenation plants beyond those being built in Saudi Arabia (two in Jubail and one at Yanbu, due onstream 2007/8)

**Figure 3 – LPG Prices (\$/Tonne)**

	C3	C4
India	600*	
Japan	530	530
China	490-500	510-520
Mont Belvieu, USA	494	491
NW Europe	482	490
Arabian Gulf (Export)	450	
KSA (internal use)	※	※

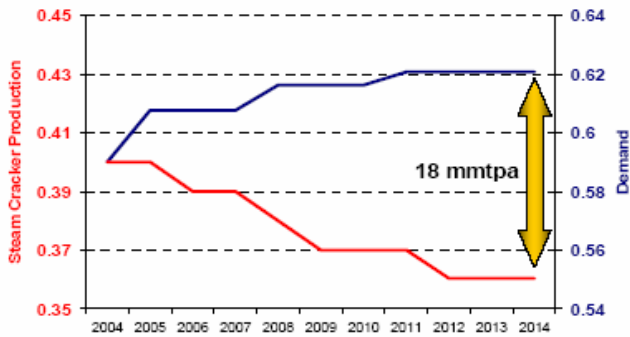
Prices 6 November 2006, cif, \*FWEL Study ※ ~30% reduction vs. export price due to lower infrastructure requirement

**Figure 4**



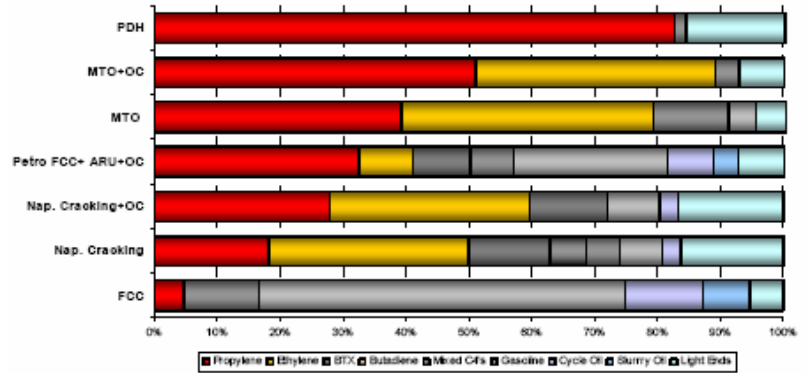
Source: Global ethylene market... by Mark Eramo. CNAI Petrochemical Focus Conference, May 2005

**Figure 5 – C3=/C2= Ratio**



Source: "On purpose" propylene.. By Jim Anderson, UOP, Global Petrochemicals Conference, Cologne, April 2005. CMAI 2005 data.

**Figure 6 – On purpose Propylene Production**



Source: "On purpose" propylene.. By Jim Anderson, UOP, Global Petrochemicals Conference, Cologne, April 2005.

In the Middle East, there is significant potential in Iran and some options in Qatar, UAE and Oman. KSA could expand ethylene production by adopting LNG to monetise methane.

Key factors in the market are:

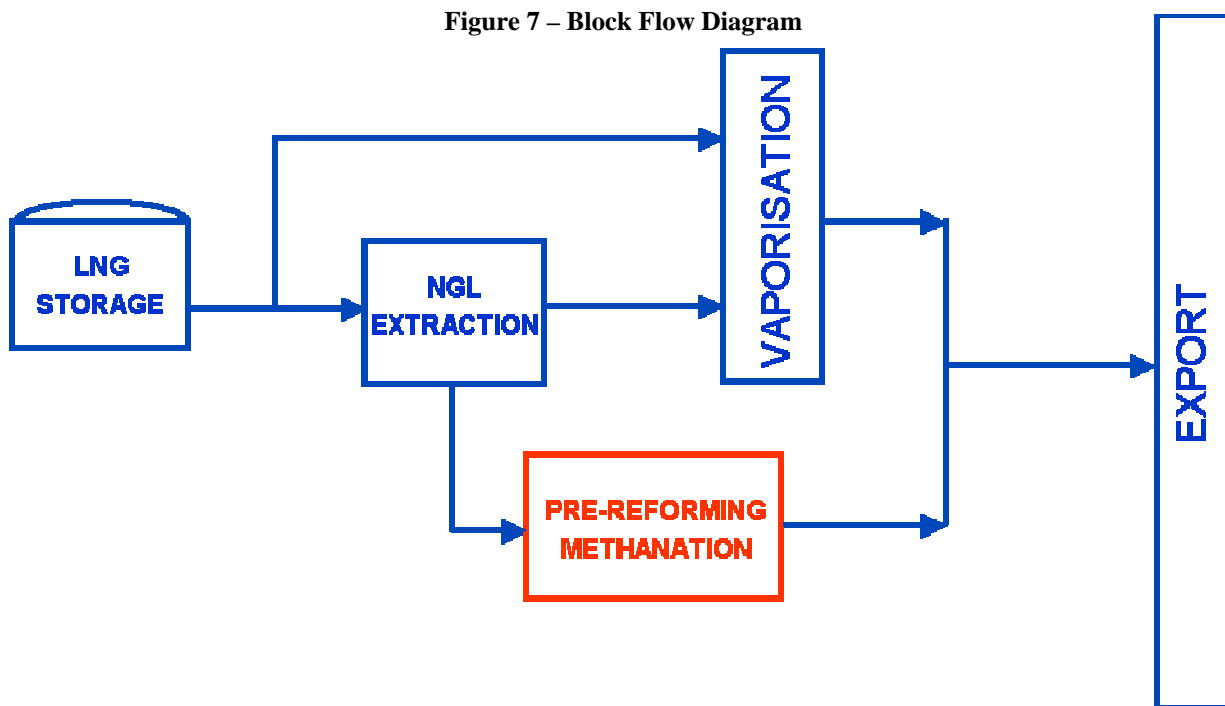
- Iran has the motivation to develop an LNG business as well as a sequence of ethane crackers in order to monetise its portion of the North Dome field (shared with Qatar), BUT, at present many companies with access to LNG are not able to do business in Iran
- Qatar exploiting the synergy with LNG
- Borouge, Abu Dhabi, separated by 60 nautical miles from LNG plant at Das Island
- No new LNG train planned in Oman before 2010, although Qalhat LNG successfully completed in 2005
- Kingdom of Saudi Arabia's ethane production constrained by capacity to process crude oil and associated gas
- LNG as an additional path for monetising surplus methane to processes currently used (methanol, etc.)

In North America, however, the extracted C2 plus heaviers are difficult to sell or handle economically in most regions, while the consumption of the fuel for the LNG terminals is often much less than the extracted C2 and heaviers from the imported rich LNG. In this paper, Option 4 - Extraction of C2+ components and pre-reforming, is the favoured route. In this case, the extracted heavies are processed using pre-reforming technology to producing low calorific Substitute Natural Gas for reinjecting into the LNG terminal send-out gas.

### CONVERION OF C<sub>2+</sub> COMPONENTS TO LOW BTU GAS FOR BLENDING TO SEND-OUT GAS

The pipeline gas specification is a major consideration, in particular, the maximum allowable concentrations for hydrogen, carbon dioxide, carbon monoxide. In processing the extracted C2+ care is required when selecting catalyst and conversion conditions. Milder conversion for reducing C3+ should also be considered. This will result in requiring a minimum amount LNG to be processed in the C2+ extraction facility.

Each receiving terminal importing LNG from a source where the composition does not meet the end-market specification will need a strategy to manage the C2 and heavier components of the gas. The following block flow diagram (Figure 7) shows an overall schematic diagram to process part of the LNG using pre-reforming technology to convert the C2 and heavier components to a low-BTU gas before recombining with the by-passed send-out gas.



The basic scheme consists of feed purification followed by pre-forming and then methanation steps. A number of reaction stages may be adopted depending upon the final gas specification required.

Pre-reforming is a mature technology established in the late 1960s. Methanation is used to reduce the hydrogen content and enable the send-out gas specification to be met.

In this case a process has been developed to convert the C2+ stream, extracted from the LNG, to a methane-rich stream (SNG) which will satisfy the export pipeline specification when recombined with send-out gas. The designer may pursue a number of technical areas in order to achieve maximum optimization and integration of the process scheme. These are:

1. The requirement for feed gas pre-treatment
2. Pre-reformer inlet pre-heat
3. Hydrogen source (for start-up and normal operation)
4. Steam generation
5. CO reduction

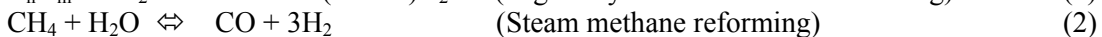
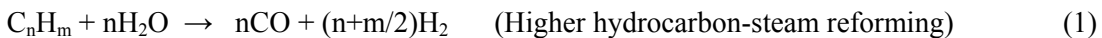
### **PROCESS OVERVIEW** (Figure 8)

The pre-reformer is the central part of this process scheme. Key features of the pre-reformer include:

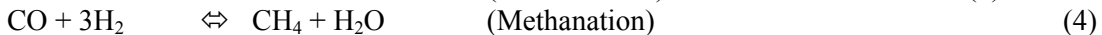
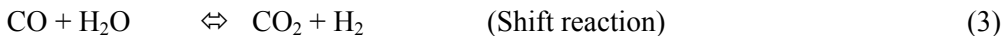
- Conversion of C2+ components to methane-rich stream by steam reforming reactions
- Nickel-based catalytic reactor
- Hydrogen required
- High inlet temperature (hence heat requirement)

The reforming reactions are shown below:

Endothermic steam reforming reactions



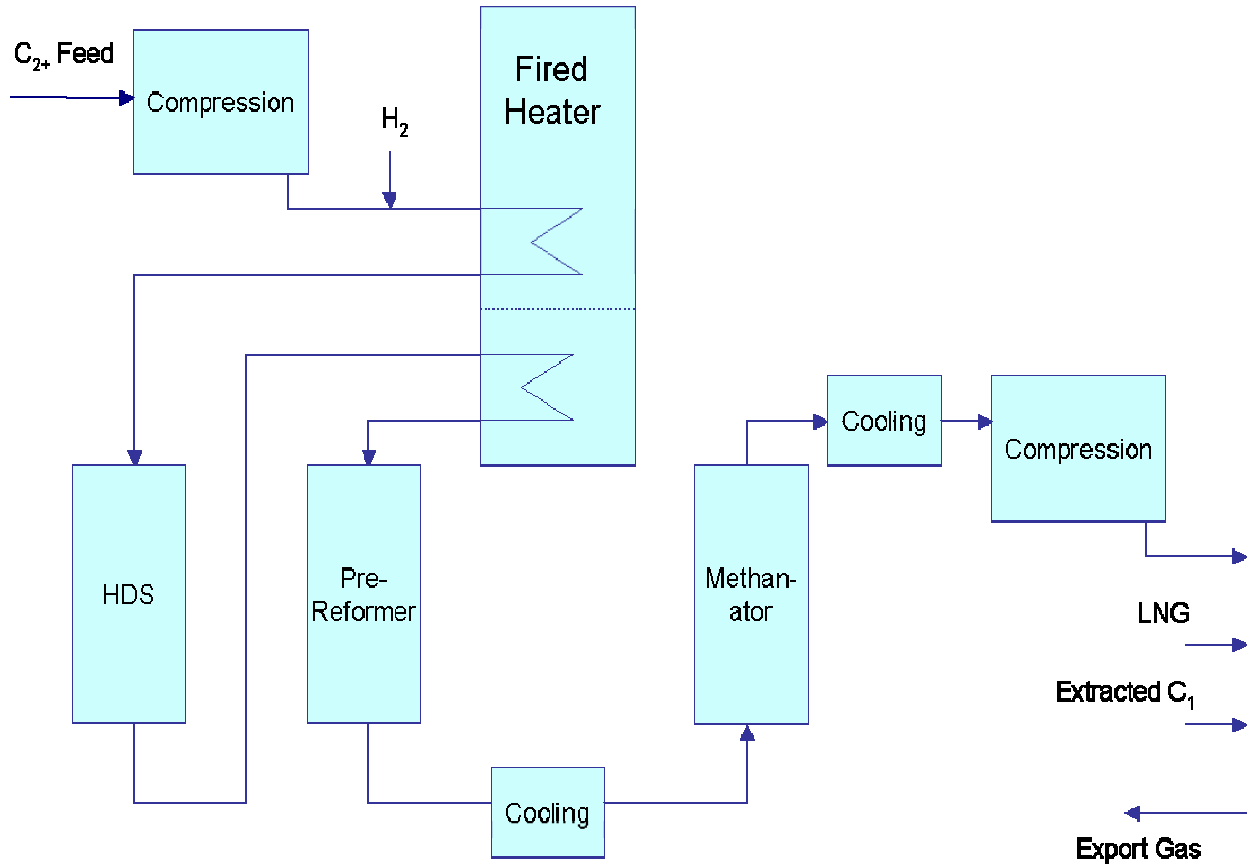
Exothermic methanation and shift reactions



This section will outline the key process steps.

The process units in converting the C<sub>2+</sub> to a low-BTU gas include a number of the processes:

Figure 8



### PRE-TREATMENT

The feed stream from the C<sub>2+</sub> extraction process is pumped, vaporised and then mixed with recycle gas from the pre-reforming reactor, to provide hydrogen for hydrodesulphurization (HDS). The feed stream is then heated in a process fired heater to raise the temperature before entering the HDS.

The HDS is comprised of three separate catalytic layers designed to treat the feed gas before entering the pre-reformer. In the first layer, any organic sulphur and chloride compounds present are converted to hydrogen sulphide and hydrogen chloride and any unsaturated hydrocarbons are saturated. Since the feed is extracted hydrocarbons from LNG, it is unlikely that there will be any unsaturated hydrocarbons present.

The treated gas then passes over a layer of zinc oxide catalyst, where the hydrogen sulphide and hydrogen chloride is absorbed by reaction to produce zinc sulphide and zinc chloride.

The third layer of catalyst protects the pre-former by ensuring an ultra-low sulphur slippage. Catalyst life in the HDS ranges from two to six years. The changeover will take up to two days.

During changeover the HDS may be bypassed so as not to disrupt LNG send-out. Alternatively, the HDS shutdown may be scheduled by exploiting opportunities when the terminal is receiving a lower gross heating value LNG.

## **PRE-REFORMING**

The desulphurised feed is then mixed with superheated high-pressure steam and preheated in the radiant section of the process-fired heater.

The mixed feed enters the pre-reformer where it flows over a bed of pre-reforming catalyst. Here the hydrocarbon feed is reformed adiabatically to produce a gas containing methane with very low C<sub>2</sub>+ hydrocarbon content and equilibrium levels of hydrogen, carbon oxides and steam. The reactions taking place over the reactor may be considered as the hydrocarbon reforming reactions (1) & (2) and the shift reaction (3).

Reactions (1) and (2) are endothermic as shown from left to right and reactions (3) and (4) are exothermic. Reactions (2), (3) and (4) are reversible and once the heavier hydrocarbons are reformed (reaction (1)), the carbon monoxide and hydrogen produced will then react together through the reverse of reaction (2). This is the methanation reaction (4) and in this direction is highly exothermic. (Reaction (1) is also reversible but under the reaction conditions adopted the equilibrium value lies very heavily to the right and hence virtually all heavier hydrocarbons are removed). For this case the overall reactions taking place in the pre-reforming reactor are exothermic, with the final gas analysis leaving the reactor being determined by an approach to the thermodynamic equilibrium of the above reactions.

## **METHANATION**

In order to meet the export gas specification for hydrogen and carbon oxides content, the pre-reformed gas must pass through a methanation process. Hydrogen, carbon monoxide and carbon dioxide react to form methane in accordance with the methanation reaction (4). Again the exothermic reaction takes place adiabatically and, due to the reduced inlet temperature of operation, the methane level in the gas is increased over the reactor. It should be noted that the removal of water is favourable so that reaction (4) is driven in the forward direction.

## **EXPORT GAS**

Heat contained in the hot gases leaving the methanator is recovered and the gas is cooled to a lower temperature to ensure the water moisture content, after combining with the unprocessed send-out gas, complies with the pipeline gas specification, typically 7 lbs/MMSCF gas. Process condensate from the knockout drums are collected and sent to a deaerator. The dry gas is compressed to export pipeline pressure and combined with the vaporised LNG and methane from the extraction unit.

## **STEAM AND CONDENSATE**

Condensate that has been produced in the process is reused to produce steam to satisfy the steam to carbon ratio at the inlet to the pre-reformer. The process condensate is of good quality as there is no nitrogen present in the feed, hence no ammonia is produced. There is also no shift reactor; hence no methanol will be formed. The condensate will be sent to the deaerator to produce high quality boiler feed water (BFW).

However, the condensate does not produce enough steam to fulfil the process requirements; hence additional make-up water is needed. The make-up water is combined with the condensate and sent to the deaerator where the CO<sub>2</sub> is stripped. A BFW dosing package, oxygen scavenger, boiler chemicals and pH control chemicals are injected into the BFW system.

BFW water is taken from the sump of the deaerator and pumped via BFW pumps to a steam generation package. Steam is generated from waste heat from the outlet of the pre-reformer. The remaining heat is provided by LNG fuel. High pressure superheated steam will be provided upstream of the pre-reformer.

## **DESIGN CONSIDERATIONS**

In applying the pre-reforming technology for converting the C<sub>2+</sub> components to a lower heating value gas, the following design strategies are to be taken into consideration:

1. The pipeline gas specification, in particular, the maximum allowable concentrations for hydrogen, carbon dioxide, and carbon monoxide. Producing the lowest BTU gas is preferred in choosing the process conditions. Milder conversion to reducing C<sub>3+</sub> should be considered. This will minimise the amount of LNG to be processed in the C<sub>2+</sub> extraction facility. Some pipeline gas has a higher maximum carbon dioxide specification of up to 4-mol% and a maximum inert gases specification of up to 7-mol%. In some cases pipeline gas has a higher hydrogen specification.
2. Minimizing hydrogen yield is an important issue in applying the pre-reforming technology. CO is likely not an issue, as it will be in compliance with the pipeline gas specifications after the converted gas blended with the send-out gas.
3. There is potential to reduce the requirement for methanation reactors. This centres on the feasibility of extracting hydrogen that could be used in the process. Another alternative would be to use one wet and one dry methanator as opposed to two wet methanators.
4. Other technologies for separating hydrogen from the syngas such as using membrane or adsorption processes may be considered.

## **CONCLUSIONS**

Cost-effective management of the C<sub>2+</sub> content and heating value specifications components can enhance the LNG value chain, increasing flexibility in sourcing LNG cargoes. Several alternatives for managing C<sub>2+</sub> components in the LNG value chain exist. We have seen that extraction can occur at either the receiving terminal or upstream of the liquefaction plant. In cases where it may not be cost-effective to reduce the C<sub>2</sub> and C<sub>3</sub> at the baseload plant, we have demonstrated that a proposed C<sub>2+</sub> management process design for send-out gas at the LNG receiving terminal could be economically viable under certain circumstances.

## **ACKNOWLEDGMENTS**

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*[1] Yang, CC, Kaplan, Al, and Huang, Zupeng, "A Cost Effective Design for **Reducing C2 and C3 At LNG Receiving Terminals,**" AIChE Spring National Meeting, March 30-April 3, 2003, New Orleans, Louisiana.*