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Achieving Product Specifications For Ethane Through To Pentane Plus From NGL Fractionation Plants

Essam Bahnassi (Head of Process Department, GASCO)
Abdul Rahman Khouri (Senior Process Engineer, GASCO)
Peter Alderton (Process Consultant, Foster Wheeler)
James Fleshman (Senior Principal Engineer, Foster Wheeler)

ABSTRACT

This paper examines the techniques that may be applied in evaluating processing options in the production of ethane, propane, butane and pentane plus products from NGL liquids.

The paper reviews the processing routes in the fractionation of NGL liquids in order to produce high quality products including consideration of mercaptans and H_2S in the upstream gas plants.

The paper makes reference to a recent GASCO project to upgrade their fractionation plant at Ruwais in Abu Dhabi. The design of this plant had to consider processing difficulties associated with NGL liquid feeds from sour upstream fields, where the processing route applied to achieve one specification may be detrimental to achieving another. These specifications include sulfur levels, paraffin content and RVP requirements.

In addition, value-engineering techniques were applied in parallel to the conceptual design and FEED to produce resulting Capex and Opex reduction and improved overall operability.

Increasingly, Energy Companies are looking to optimize their developments and this paper highlights some of the key issues that should be considered in the design of NGL fractionation trains and the production of high quality products.

GENERAL CONSIDERATIONS

The processing of an NGL feed stream into saleable products consists essentially of two main steps:

Fractionation followed by Product Treating

The key variables, which must be taken into account when setting the processing scheme, are feed composition (including impurities) and product specifications. These feed and product specifications impact configuration selection. If the wrong selection is made, one may be unwittingly excluded from key markets now and in the future. Also, if a wrong approach is adopted, there will be cost and schedule penalties if corrected later in the project execution phase.

Feed streams for NGL fractionation plants originate from upstream processing plants, which may be remote, and these in turn take their feed directly from gas reservoirs or crude stabilization units. The compositions of these feeds vary depending on their source and also vary over time. Design cases should consider these factors and reflect them if possible and in consultation with the client.

The interaction between upstream/remote processing should be considered carefully since it may be more appropriate to recover impurities in an upstream facility where they can be processed in a more environmentally acceptable way such as, for example, re-injection. Upstream processing is outside the scope of this paper but it is important to note that overall synergies should be investigated to develop the most cost effective approach in the client's interest.

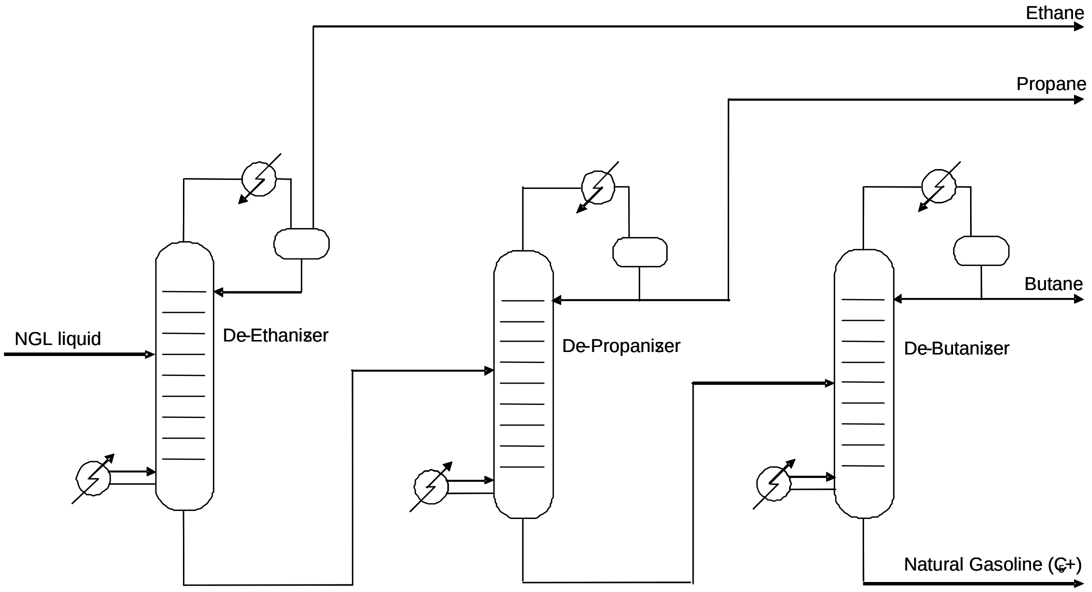
Fractionation

Products from a fractionation process can include:

- Demethanised Product
- Deethanised Product
- Ethane/Propane mixtures (EP)
- Commercial Propane
- Propane/Butane mixture (LPG)
- Butane(s)
- Butane/Gasoline mixtures
- Natural Gasoline
- Mixtures with a vapor pressure specification

A typical fractionation train is shown below and consists of a series of distillation columns: a de-ethanizer, a de-propanizer and a debutanizer. The mixed butane product can be further split in a de-isobutanizer to produce normal and isobutane. In a case where there is not a market for butane and propane as separate components, a mixed LPG stream can be produced, thus removing one of the columns.

Figure 1: Distillation Train – De-ethanizer, De-propanizer, De-butanizer



The NGL liquid enters the de-ethanizer, which separates the ethane from the mixed NGL liquid. The column operates with a partial condenser and produces a cold ethane gas product. The bottom product from the de-ethanizer feeds the de-propanizer. A propane stream is taken from the de-propanizer reflux drum. The debutanizer removes the butane from the remaining NGL liquid. Pentane plus, or natural gasoline, is taken as a bottom product from the debutanizer.

Before any design calculations can be made on a fractionation train, the column operating pressures must be fixed and these in turn will depend on the cooling medium used for the overhead condensers. Typical cooling media are air, water and refrigerant. Air cooling is the least expensive but process temperatures are limited to between 45 ^{0}C and 50 ^{0}C. With cooling water, process temperatures of between 35 ^{0}C and 40 ^{0}C are possible. Below 35 ^{0}C, refrigeration will be required and this is the most expensive option. Generally, it is best to operate at as low a pressure as possible in order to improve the relative volatility between key components. However, it will be necessary to carry out a life cycle economic analysis between the cost of the cooling system against the cost of adequate separation.

The reboilers of the fractionation columns can be steam or hot oil heated or they can be fired heaters. Steam heating is more commonly used. Increased steam pressure provides a greater temperature driving force, which reduces the surface area of the reboilers. The choice will depend on the availability of utilities and economic considerations.

Another decision to be made is whether products are to be generated as a vapor and subsequently liquefied in the rundown system or whether total condensers are to be used. Again, an economic analysis will dictate which is the best option.

The type of processing will depend on the required state of the products and this will be dictated by market requirements and the export mode. The light products such as ethane can be exported as a pressurized gas if there is a local market such as an ethylene cracker. Alternatively, a refrigerated product can be generated although this is likely to be costly because of the very low temperatures required. Products in the middle of the range such as propane, butane and LPG can be stored and exported in the totally refrigerated state at atmospheric pressure or as a pressurized fluid at ambient temperature. The heavy products such as gasoline are normally stabilized so that they can be stored and exported at ambient temperatures and atmospheric pressure.

The NGL fractionation unit can be modified to improve the heat integration, thus reducing the heating and cooling demands. Some of the most common heat integrations are:

- De-propanizer condenser linked to de-ethanizer reboiler
- Debutanizer condenser linked to de-propanizer reboiler
- De-propanizer and debutanizer feed/bottoms exchangers
- De-ethanizer overheads linked to de-propanizer and debutanizer overhead product coolers

Product Treating

The following table lists contaminants that are typically present in NGL together some reasons for their removal from the products:

Contaminant	Reasons for Removal
• Hydrogen sulfide • Carbon dioxide • Carbonyl sulfide • Carbon disulfide • Mercaptans • Organic sulfides • Nitrogen • Water	• Safety and Environmental • Corrosion control • Product specification • Prevention of freeze-out at low temperatures • Prevention of catalyst poisoning in downstream facilities

Guarantees on product specs need to be established at the outset because off takers need guarantees to meet downstream commitments. The importance of this is that for the project to guarantee the particular product quality, specific features of a proprietary technology may be needed to obtain the correct level of design assurance.

The lighter products from fractionation such as ethane will contain all the CO_2, most of the H_2S and a small amount of the COS from the NGL feed stream. Aqueous alkanolamine processes are the most widely used removal of these impurities. Selectivity between H_2S and CO_2 removal is an important factor based on the ethane product specification and the composition of the acid gas. Ethane product is usually sent to an ethylene cracker and the CO_2 content is critical because affects hydraulic capacity without adding to yield. The acid gas stream from treating needs to have a high H_2S:CO_2 ratio if it is to be fed to a sulfur recovery unit. The degree of selectivity will determine the amine to be used and whether the process needs to be licensed or not. As these processes are aqueous, some form of downstream dehydration such as molecular sieve or TEG will be required.

The propane product will contain most of the COS, about half the methyl mercaptan together with the small amount of remaining H_2S from the NGL feed stream. An amine process similar to that used for ethane product is normally used to remove the H_2S and COS although removal of the latter is somewhat more problematical requiring the use of mixer settlers as well as an extraction column. If the level of methyl mercaptan is significant, caustic wash or a regenerative caustic process would also be used followed by dehydration in molecular sieves. If the concentration of methyl mercaptan is low, the caustic extraction step could be omitted and molecular sieves used for final polishing as well as dehydration.

The butane product will contain the remainder of the methyl mercaptan, most of the ethyl mercaptan together with about half of the CS_2 from the NGL feed stream. As there is no H_2S or COS present, there is no need for amine treating and therefore no need for dehydration. The choice of process for removal of the mercaptans and CS_2 will be either caustic treating or molecular sieves as described above for propane treating.

If an LPG product is generated, the treating process options will be similar to those described above for propane and butane.

The bottom product from fractionation such as gasoline will contain the remainder of the heavy mercaptans and organic sulfur compounds. Processes available are again caustic treating, molecular sieves and hydrotreating.

Economic decisions in order to select which product treating process to use must take account of any licence fee and be made on a life cycle basis.

It should be noted that impurities in the upstream feed gas are concentrated considerably in the fractionation products and it is therefore important for the feed analysis to be as accurate as possible and to include as many components as possible even if their concentration appears to be negligible. It is also important to pay particular attention to the computer simulation model developed in order to ensure that the impurities are correctly distributed between the various products.

RUWAIS FRACTIONATION PLANT

The Abu Dhabi Gas Company (GASCO) is planning to extend the existing gas recovery facilities by installing new phases to the Habshan Onshore and Asab Gas developments. These will provide ethane rich NGL streams to the fractionation plant at Ruwais, which will require a new NGL pipeline to Ruwais and a third train to be installed there. The fractionation plant will separate 24,400 ton/day of NGL feed into ethane, propane, butane and pentane plus products. Ethane will be sent to an adjacent petrochemical complex for conversion to ethylene. The propane, butane and pentane plus products will be exported as liquids by ship.

Foster Wheeler represented GASCO as Project Management Consultant for the FEED phase of this project and was responsible for the engineering design carried out by the FEED contractor, Bechtel.

The map of Abu Dhabi below shows the relative locations of Asab, Habshan and Ruwais

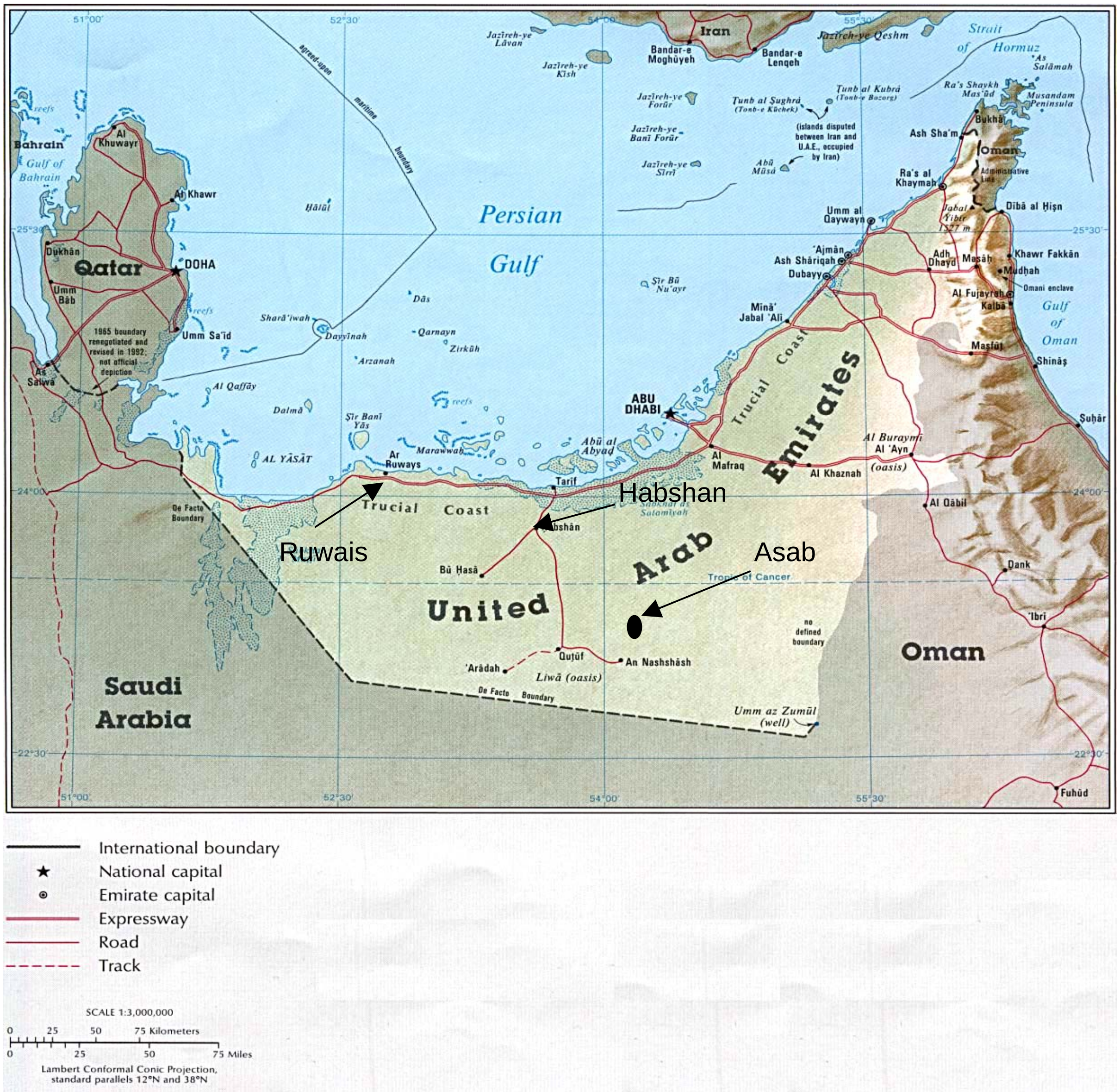
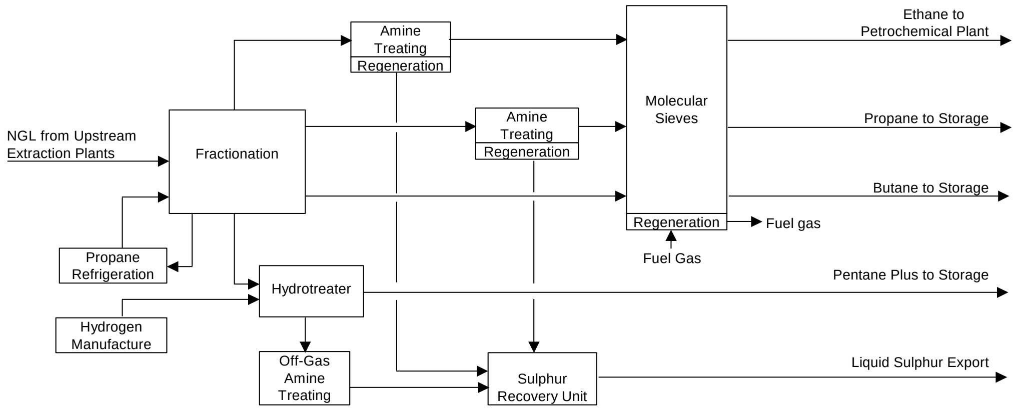


Figure 2 – Abu Dhabi

Figure 3 - Ruwais Fractionation Plant



During the early stages of FEED, several studies were carried out for fractionation, treating and refrigeration in order establish the optimum configuration for the plant. These included:

- Fractionation sequences and process options
- Configuration selection and reuse of existing units
- NGL fractionation with blended feedstock
- Alternative refrigeration system design
- Sulfur recovery
- Technology upgrade of product treating
- Reboiler heating medium
- Refrigeration package drivers
- Cooling system
- Molecular sieve regeneration and reduction of sulfur impurities

Hydrocarbon Processing

The hydrocarbon processing train is designed to take the feed NGL stream and separate it into four products, ethane, propane, butane and pentane plus.

The NGL feed stream hydrocarbon composition in mol% is as follows:

Methane	2.11
Ethane	37.26
Propane	25.84
i-Butane	6.55
n-Butane	13.03
i-Pentane	3.42
n-Pentane	3.61
n-Hexane	2.78
C6+	3.34

The balance of the NGL feed is made up of CO_2 and sulfur impurities. The levels of these impurities are given later in the document in the section headed “Treating”.

The following product specifications relating to hydrocarbon composition are to be achieved by using three fractionation columns and varying the operating pressures, operating temperatures and number of trays:

- Ethane
  4.46 wt% max C1
  88.7 wt% min C2
  1.13 wt% max C3+

- Propane
  2 liq vol % max C2
  96 liq vol% min C3
  2 liq vol % max C4s
  Vapor pressure 200 psig max at 100 ^{0}F

- Butane
  2 liq vol % max C3
  96 liq vol% min C4s
  2 liq vol % max C5s
  Vapor pressure 70 psig at 100 ^{0}F

- Pentane +
  RVP 12.5psi at 100 ^{0}F

Five fractionation sequences were identified for consideration:

- De-ethanizer/De-propanizer/De-butanizer (Sequence of existing trains at Ruwais)
- De-ethanizer/De-butanizer/Propane-Butane Splitter

- De-propanizer/De-butanizer/Ethane-Propane Splitter
- De-butanizer/De-propanizer/Ethane-Propane Splitter
- De-butanizer/De-ethanizer/Propane-Butane Splitter

The last two sequences were not considered further as they had the highest operating costs. The first three sequences were evaluated against the following criteria:

- Operating Costs
- Equipment Costs
- Complexity and Operability
- Interchangeability with the existing two trains at Ruwais

The first sequence had the lowest NPV. The second sequence had the lowest fractionation plant TIC of all the cases and would permit combined propane and butane LPG treating. However, feedback from two molecular sieve vendors showed that a combined LPG treating plant would require a larger total molecular sieve volume than for separate LPG treating and this would negate any potential advantage.

Based on the above criteria, the first sequence was selected for development into the design for the new Ruwais Train 3.

The de-ethanizer is reboiled against LP steam and the overheads are condensed against propane refrigerant. The overhead vapor product containing 90% ethane is then sent to treating.

The de-ethanizer bottom product is fractionated in the de-propanizer into a propane overhead liquid product and a C4+ bottom product. The de-propanizer is reboiled against LP steam and the overheads are condensed against cooling water. Propane liquid product is then sweetened in the propane treating section of the plant before chilling with propane refrigerant and rundown to storage as a fully refrigerated product.

The de-propanizer bottom product is fractionated in the debutanizer into a butane overhead liquid product and a pentane plus bottom product. The debutanizer is reboiled against LP steam and the overheads are condensed against cooling water. Butane liquid product is then chilled with propane refrigerant and sweetened in the butane treating section of the plant before further chilling with propane refrigerant and rundown to storage as a fully refrigerated product. Pentane plus bottom product is cooled and chilled before rundown to storage as a stabilised liquid.

Utilities

Cooling water was selected rather than air-cooling as there is an abundant source of seawater at Ruwais, which is situated on the coast. Further, direct seawater cooling was used instead of an indirect system for reasons of simplicity and compatibility with the cooling system on the existing trains

Options for the column reboiler heating are steam, hot oil or fired heaters. Steam was selected on economic grounds, safety, reliability, and plot area.

The refrigeration compressor driver was selected as an electric motor instead of gas or steam turbine since there is a cheap source of electric power from an adjacent General Utilities Plant.

Treating

The NGL feed contains impurities, which are distributed between the products as follows:

Impurity	Concentration	Ethane	Propane	Butane	Pentane Plus
CO_2	1.98 mol %	100%			
H_2S	642 ppm mol	97%	3%		
COS	154 ppm mol	7%	93%		
CS_2	0.5 ppm mol			50%	50%
M-Mercaptan	15 ppm mol		47%	53%	
E-Mercaptan	2.7 ppm mol			71%	29%
2C3Mercaptan	14 ppm mol				100%
nPMercaptan	5.5 ppm mol				100%
nBMercaptan	20 ppm mol				100%
1Pentanthiol	26 ppm mol				100%
1-C9-Thiol	1.4 ppm mol				100%
diM-Sulfide	16 ppm mol			8%	92%
M-E-Sulfide	1.4 ppm mol				100%
diMdiSulfide	3.6 ppm mol				100%
iP-Sulfide	5.9 ppm mol				100%

These impurities are removed by treatment of the individual product streams

Ethane Treating

The ethane feed to treating contains 6.84 %wt CO_2, 1523 ppmv H_2S and 22 ppmw COS.

The main specifications for the treated ethane product are as follows. A full specification is given in Attachment 1.

CO_2	5.60 wt% max
H_2S	20 ppmv max
COS	7 ppmw max
H_2O	24 ppmw max
Mercaptans	9 ppmw max

The high CO_2 to H_2S ratio (30:1) in the Train 3 feed means that for ethane treatment, a highly selective hydrogen sulfide removal process, such as one using methyl diethanolamine (MDEA), is necessary. A two-stage absorption process was originally thought to be needed to generate an acceptable sulfur recovery unit (SRU) feed and separate surplus carbon dioxide to meet the product specification for the petrochemical plant. For a single stage MDEA absorber a selectivity ratio of 4:1 is typically achieved. A two-stage absorption process in which the regenerator overhead is reabsorbed in MDEA leads to an overall 16:1 selectivity ratio.

Alternatively, a hindered amine process can be considered for selective absorption of H_2S. Physical solvents or mixed solvents should be avoided, as these will lower selectivity for H_2S, even if they offer improvements in COS removal.

For economic reasons, it was finally decided to use a single stage licensed process from Shell Global Solutions International (SGSI) with an MDEA based solvent for ethane treating. However, a further acid gas enrichment absorber was included as part of the sulfur recovery unit in order to achieve a suitable acid gas feed to the Claus reactor.

The required dryness of 24ppmw water content can be met by activated alumina, which has a higher absorption capacity than molecular sieves. Further, there is a grade of alumina available specifically for COS removal (Selexsorb®), so that activated alumina would be an acceptable alternative for molecular sieves. However the final selection remained as molecular sieves for reasons of similarity with the existing trains.

The selected treating scheme for ethane is therefore SGSI MDEA followed by molecular sieves.

Propane Treating

The propane feed to treating contains 52 ppmv H_2S, 762 ppmw COS and 31 ppmw methyl mercaptan.

The main specifications for the treated propane are as follows. A full specification is given in Attachment 1.

Moisture Dryness (ASTM D2713 test)	Pass
Copper Strip	No. 1
H_2S (ASTM D2420)	Negative
Total sulfur	15 ppmw max

Amine treatment of the propane product is required to remove H_2S and COS. In view of the high levels of COS expected in this stream, both diglycolamine (DGA) and diethanolamine (DEA) can be considered as alternatives to diisopropanolamine (DIPA) which is used in the existing trains 1 and 2. Maximizing COS removal at the amine stage is preferred, as the COS can then be converted to CO_2 and sulfur in the SRU thereby reducing overall plant emissions. Failure to remove COS would lead to an inability to pass the copper strip test.

For economic reasons and compatibility with the existing trains 1 and 2 treating, it was finally decided to use the SGSI licensed ADIP process with DIPA as the solvent

Caustic treating is normally used to remove mercaptans from a propane product following the amine removal of H_2S and COS. After chilling and coalescing to remove free water from both streams, molecular sieves can then be used for dehydration and sulfur polishing to meet the required product specifications.

Removal of mercaptan sulfur using only molecular sieves was originally rejected because the intermittent peaks of sulfur compounds released during regeneration cannot be absorbed and satisfactorily processed in the SRU. If the regeneration gas is left untreated, then short term, high atmospheric sulfur emissions will result from the fuel gas.

Subsequently it was decided to remove more of the light mercaptans in the upstream NGL extraction plants at Habshan and Asab. This makes it possible to use only molecular sieves for polishing and dehydration at Ruwais without exceeding sulfur emission limits when the regeneration gas from the molecular sieves is burnt as fuel gas.

The selected treating scheme for propane is therefore SGSI ADIP followed by molecular sieves

Butane Treating

The butane feed to treating contains 33 ppmw methyl mercaptan, 13 ppmw ethyl mercaptan and 8 ppmw dimethyl sulfide.

The main specifications for the treated butane are as follows. A full specification is given in Attachment 1.

Moisture Dryness (ASTM D2713 test)	Pass
Copper strip	No. 1
H_2S (ASTM D2420)	Negative
Total sulfur	15 ppmw max

As there is no H_2S or COS present, amine treating is not required. The preferred method of treating for removal of heavy sulfur compounds is again normally caustic wash. However, with the removal of more of the mercaptans upstream, it is possible to use molecular sieves for final polishing without exceeding sulfur emission limits when the regeneration gas from the molecular sieves is burnt as fuel gas.

The selected treating scheme for butane is therefore only molecular sieves.

Pentane Plus Treating

The pentane plus product contains CS_2, ethyl mercaptan, dimethyl sulfide as well as all the heavier mercaptans and disulfides.

The main specifications for the pentane plus product were initially as follows. A full specification is given in Attachment 1.

Total Sulfur	100 ppmw max
Total Mercaptan Sulfur	100 ppmw max

Originally no treating was envisaged for the pentane plus product. However, following two extensive sampling campaigns, it was concluded that the total sulfur content of the pentane plus product could not be guaranteed to be less than 100 ppmw simultaneously with an RVP of 12.5 psi. The RVP could be controlled in the upstream plant but this would be at the expense of allowing more heavy sulfur compounds to enter the NGL product. It was therefore decided to include a hydrotreater at Ruwais as a future option in the design and relax the sulfur specification to 240 ppm wt. The hydrotreater would allow removal of sulfur compounds from the pentane plus product down to a level of 1ppmw. Also, the debutanizer design was modified by using additional trays to reduce the butane content in the pentane plus product. Another contingency measure incorporated at Ruwais was to leave space for a de-isopentanizer column. This would produce a separate isopentane product and have the effect of reducing the RVP of the pentane plus product.

During the course of the project, the paraffin content of the pentane plus became an issue. Although not a contractual requirement, a minimum concentration of 85% paraffins was desired. It was clear that this specification was going to be exceeded and no additional processing was necessary at Ruwais to increase the paraffin content.

Value Engineering

A Value Engineering exercise was completed and the following table summarizes the five ideas related to hydrocarbon processing and product treating which were selected for further consideration

Proposal	Capex Saving	Opex Saving	Schedule	Risk	Ease of Implementation
Employ a common regeneration system for propane and ethane treating.	>$5M	$500K to $1M	Schedule Increase	No Change	Easy
Replace ethane mol sieve dryers with water dew point control to meet the ethane water specification, on a fit-for-purpose design basis.	>$5M	>$5M	No Saving	No Change	Easy
Use direct air coolers instead of either indirect water cooling or seawater cooling, where possible.	<$500K	OPEX Increase	No Saving	No Change	Easy
Install sponge loss recovery system in mol sieves in order to recover valuable hydrocarbons, which would be otherwise lost to the fuel gas (i.e. achievement of an improved cash flow).	CAPEX Increase	<$500K	No Saving	No Change	Easy
Replace steam reboilers with reboilers running on water/glycol.	>$5M	OPEX Increase	No Saving	No Change	Easy

In the final event, only the sponge oil recovery proposal was recommended for implementation.

CONCLUSIONS

The key issues, which should be addressed when designing NGL fractionation plants, are:

- **Market requirement for products**
  Pure products or combined
- **Required product specifications and prices**
  Economic considerations to evaluate the higher CAPEX and OPEX vs. improved specifications and thus higher prices
- **Product export options**
  Refrigerated or pressurized
  Pipeline or carrier
- **Existing plant synergies**
  Using similar processes to those existing for economy of spares and operator familiarity
- **Possibilities for upstream removal of impurities**
  Removal of lighter mercaptans in molecular sieve treating feed gas to NGL recovery unit
- **Availability and cost of utilities**
  Location of fractionation plant and use of local general utility plants
- **Environmental considerations**
  Atmospheric emissions of sulfur compounds removed from products
- **Value Engineering**
  To highlight potential cost savings at an early stage in the design

ATTACHMENT 1: FULL PRODUCT SPECIFICATIONS

Ethane Product

Component	Value
C1	4.46 wt% (max)
C2	88.70 wt%.(min)
C3+	1.13 wt% (max)
CO_2	5.60 wt% (max)
H_2S	20 ppmv (max)
COS	7 ppmw (max)
H_2O	24 ppmw (max.)
Mercaptans	9 ppmw (max)

Commercial Propane Product

Property	Test Method	Value
Ethane (Liq vol %)	ASTM D2163	2 % (max)
Propane (Liq vol %)	ASTM D2163	96 % (min)
Butanes (Liq vol %)	ASTM D2163	2 % (max)
Vapor Pressure calc @100°F	ASTM D2598	200 psig (max).
Moisture (Dryness test)	ASTM D2713	Pass
Volatile Residue (95% evaporation) °F	ASTM D1837	-37 (max)
Residual Matter (R)	ASTM D2158	20 (max)
Oil Stain observation (ml)	ASTM D2158	0.1 (max)
Corrosion Strip 1hr @ 100°F	ASTM D1838	1 (max)
Hydrogen Sulfide	ASTM D2420	Negative
Total sulfur, ppmw	ASTM D3246	15 (max)

Commercial Butane Product

Property	Test Method	Value
Propane (Liq Vol %)	ASTM D2163	2% (max)
Butane (Liq Vol %)	ASTM D2163	96% (min)
Pentane (Liq vol %)	ASTM D2163	2% (max)
Vapor Pressure (calc) @ 100°F	ASTM D2598	70 psig (max)
Moisture (Dryness Test)	ASTM D2713	Pass
Volatile Residue (95% evap) °F	ASTM D1837	+36 (max)
Residual Matter (R)	ASTM D2158	20 (max)
Oil stain Observation (ml)	ASTM D2158	0.1 (max)
Corrosion Cu strip 1 hr @ 100°F	ASTM D1838	1 (max)
Hydrogen Sulfide	ASTM D2420	Negative
Total sulfur, ppmw	ASTM D3246	15 (max)

Pentane Plus Product

Property	Test Method	Value
RVP @ 100°F	ASTM D323	12.5 psi (max)
Total sulfur. ppmw	ASTM D3120/ASTM D 2622/ UOP357/IP 336	100 (max)**
Total Mercaptan Sulfur, ppmw	ASTM D 3227	100 (max)**

*Colour Saybolt	ASTM D156	+30 (min)
*Density at 15^{o}C, kg/l	ASTM D1298 / ASTM 4052	0.64 (min)/0.68(max)
*Distillation	ASTM D86	
IBP, ^{o}C		25 (min)
10%vol, ^{o}C		Report
50%vol, ^{o}C		40 (min) / 85 (max)
90%vol, ^{o}C		115 (max)
EP, ^{o}C		170 (max)
*PONA	Gas Chromatograph	
Paraffins, vol%		85 (min) (report iso-/n-)
Olefins, vol%		1 (max)
Naphthenes, vol%		Report
Aromatics, vol%		7 (max)
*Hydrogen Sulfide	ASTM D235 / UOP 41	Negative
*Corrosion Cu strip, 3hr @ 40^{o}C	ASTM D130	Report
*Lead, ppbw	IP 224 / ICP-MS	50 (max)
*Arsenic, ppbw	ICP-MS /Cold Vap AAS /UOP 387	20 (max)
*Mercury, ppbw	ICP-MS / Cold Vap AAS	1 (max)
*Total Chlorine, ppmw	ICP-MS / Total Coulometric Titration / ASTM D 4929B	1 (max)
*Oxygenated Products, ppmw	IP-BG 91	50 (max)

* These product specifications are shown as expected values, and are assumed to be met by the production from the existing fields. No facilities are provided to control these parameters.

** If the hydrotreater is installed these specifications will be 1ppmw (max).