


# SPOT THE difference



Jon Lewis,  
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complexity and  
contaminants in  
gas processing.

**O**ne often wonders why a new gas plant is designed for every application. Surely a gas plant is a gas plant, and as countless numbers of them have been designed before, one of those must fit any new project. This article explores aspects of gas plant design and configuration and provides insight into why there are so many differences between gas plants.

The functional requirement of a gas processing train is to meet a gas export specification, and that depends on the market, whether it is local power generation, pipeline transfer for industrial or domestic use, reservoir reinjection, or as feedstock for a variety of gas

to product technologies (GTP). Figure 1 shows a range of GTP technologies, but even within any specific area there may not be a universal product specification. LNG specification, for example, varies around the world with gross calorific values (GCV) ranging from approximately 38 - 43 MJ/m<sup>3</sup> in Alaska to Australia respectively.

There are liquid products too. Condensate will invariably be produced, but LPG (mixed propane and butane) or separated propane and butane products may also be required dependent in

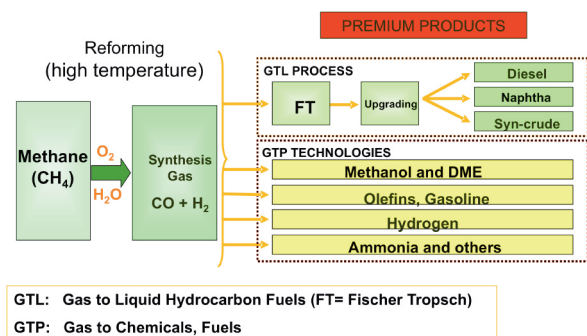


Figure 1. Gas to product technologies.

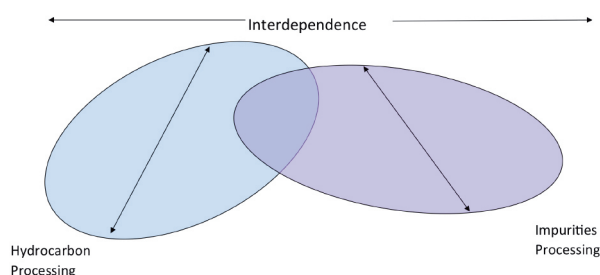


Figure 2. Interdependence of hydrocarbon and impurities processing.

Components	% molar
H <sub>2</sub> O	1.56
N <sub>2</sub>	3.31
CO <sub>2</sub>	1.74
H <sub>2</sub> S	0.66
C1	80.97
C2	5.17
C3	1.92
iC4	0.41
nC4	0.69
iC5	0.28
nC5	0.28
C6 plus	2.48
COS	0.0003
Mercaptans	0.05
MEG	0.44
Total	100.00

part on the local market. In addition a separate ethane product may be required as well.

Typically, a process engineer will be asked during the conceptual phase of a project to meet the gas export specification using the minimum processing steps, i.e. what are the minimum requirements for hydrocarbon liquid removal to meet maybe just a dew point specification and a heating value?

Whether one needs to recover propane to meet the sales gas specification, there may be benefit in doing so for economic reasons, i.e. is the propane worth more as a liquid product rather than just as energy value in the sales gas? Supplemental to this question would be the matter of how much propane should be recovered, and is flexibility required to change the operating mode of the plant to maximise or minimise propane recovery.

So having ascertained the basic process requirement in terms of hydrocarbon processing, the next step is to determine what level of contaminant removal is required. Typical contaminants are H<sub>2</sub>S, CO<sub>2</sub>, mercaptans, carbonyl sulfide, mercury and water, and the level to which their removal is required is determined by the market. For example, a CO<sub>2</sub> specification for sales gas for local or national transmission system could be in the range 1 - 4 vol%, whereas it would be typically 50 ppmv for supply to a liquefaction facility for LNG production. Similarly, the degree of water removal depends on either a sales gas specification or an internal specification within the process scheme to prevent hydrate or ice formation.

Clearly then the processing requirements for hydrocarbon processing and contaminant removal are not independent and plant configuration must be considered holistically to determine the optimum solution. This is illustrated in Figure 2.

Other factors that should be considered include declining reservoir pressure over time, gradual souring of feed gas, uncertainty of feed composition, new product markets and accommodating new reservoir fluids. Arrival pressure and temperature at the plant can also affect design as measures may need to be taken to prevent hydrate formation upstream. Ambient temperature too, and its range, can be significant as gas plants are invariably air cooled.

The remainder of this article considers two case studies that serve to illustrate some of the factors described above.

## Case study 1: impact of ethane recovery on gas sweetening

Table 1 contains a simplified feed gas composition for the gas entering the processing plant. Figure 3 shows a simplified block flow diagram to illustrate the sections of part considered in this case study.

	Base case	Alternative 1	Alternative 2
Solvent	45 wt% MDEA	30 wt% DEA	30 wt% DEA
Solvent rate (m <sup>3</sup> /hr)	254	703	375
Absorber diameter (m)	4.3	5.3	4
Reboiler duty (MW)	21.9	68.7	31.4
CO <sub>2</sub> in treated gas	1 mol%	4 ppmv	100 ppmv
CO <sub>2</sub> in feed to ETU	74 880	30	756

The required products are sales gas, ethane, propane, butane and condensate (C<sub>5+</sub>). The treated gas specification leaving the gas sweetening unit (GSU) for CO<sub>2</sub> and H<sub>2</sub>S was 1 vol% and 3 ppmv respectively.

A base case design was developed using generic MDEA as the solvent in the GSU as significant CO<sub>2</sub> slippage was deemed appropriate, which would not be provided using DEA. This approach was identical to that adopted in similar projects but the difference in this case was the requirement for ethane recovery.

The scheme resulted in a CO<sub>2</sub> content to the ethane treating unit (ETU) of 7.5 wt% whereas the product ethane specification from this unit was 50 ppmwt. The main components of the ETU were amine treating with dedicated regeneration plant, and molecular sieve for drying and removal of COS.

Alternative cases using generic DEA were considered to investigate removal of more CO<sub>2</sub> in the GSU in order to reduce the treating requirements in the ETU. In order to obviate the need for any ethane treatment the CO<sub>2</sub> concentration in the ethane stream would need to be 50 ppmwt and this required removal of CO<sub>2</sub> in the GSU to a level of 4 ppmv, i.e. a level which is theoretically possible, but not practicable due to concerns around liquid distribution on trays with high residence times and questionable reaction kinetics. A more practicable GSU outlet concentration for CO<sub>2</sub> using generic DEA is 100 ppmv and this gave a CO<sub>2</sub> concentration in the ethane steam to the ETU of 756 ppmwt (0.05 mol%).

Two options for the ETU were considered: one using DEA to remove the reduced level of CO<sub>2</sub>, with molecular sieve for dehydration and COS removal; and a second just using molecular sieve for removal of both CO<sub>2</sub> and COS. Dehydration is not required in the absence of DEA treating. This second option, although theoretically possible with the base case design, is not economic for such high levels of CO<sub>2</sub> in the ethane feed.

Tables 2 and 3 give summary parameters for the options for both the GSU and the ETU.


The base case design for the GSU using MDEA was discarded in favour of DEA (Alternative 2) despite the disadvantages within that unit, and the ETU was revised to use molecular sieves only. There was an additional disadvantage concerning the design of the sulfur recovery unit (SRU) as the revised acid gas feed stream was considerably leaner in H<sub>2</sub>S content but this too was considered acceptable. CO<sub>2</sub> emissions were also increased as a result of reduced slip in the GSU.

## Case study 2: impact of uncertainty of mercaptan levels

In this study a hybrid solvent, containing amine for H<sub>2</sub>S removal and concomitant CO<sub>2</sub> reduction, and a physical component for removal of organic sulfur species, primarily mercaptans, was used in the GSU to treat a natural gas stream. The use of a physical component resulted in some absorption of hydrocarbon components, which were released in the high pressure flash drum and the solvent regeneration facilities and subsequently lost, thus contributing to the acid gas stream routed to the SRU.

During plant operation the concentration of mercaptans in the feed gas was much lower than that used in the design basis for the facility resulting in a suboptimal design. Opportunity was seen to reformulate the solvent to address this by reducing the physical component concentration to reduce the physical absorption of hydrocarbon components, leaving them in the treated gas stream, thereby reducing the flash gas rates and improving the quality of the acid gas stream to the SRU. Other advantages achieved were higher CO<sub>2</sub> slip in the H<sub>2</sub>S absorber, lower emissions, reduced power consumption, reduced entrainment of gas bubbles in the rich solution and improved performance of the flash drums.

In order to prove that the installed plant could operate with the reformulated solvent, checks were made on each component. Table 4 contains a summary of solvent related parameter changes. There was essentially no change in the reboiler and lean/rich solvent exchanger duties.

These case studies are just two examples serving to illustrate that the removal of contaminants in gas streams should not be considered independently from the requirement to process the hydrocarbon components to meet the product specifications. The degree of interdependence for an optimum overall processing scheme will differ for each application and must be carefully considered during the design phase. 

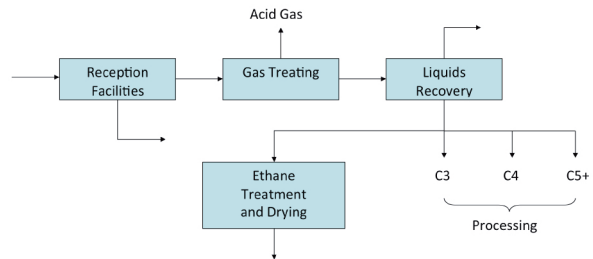


Figure 3. Simplified block diagram

Table 3. Ethane treating unit parameters

GSU	Base case	Alternative 2	
	With amine unit and molecular sieves	Amine unit and molecular sieves	With molecular sieves
Solvent	30 wt% DEA	30 wt% DEA	-
CO <sub>2</sub> in feed (mol%)	5.2	0.05	0.05
Solvent rate (m <sup>3</sup> /hr)	132	1.5	-
Absorber diameter (m)	1.8	1.3	-
Reboiler duty (MW)	9.7	0.14	-
Molecular sieve	Dehydration	Dehydration/COS	CO <sub>2</sub> /COS
Diameter (m)	2.3	2.3	2.1
Bed height (m)	6.5	6.5	6.4
CO <sub>2</sub> in ethane	50	50	50

Table 4. Summary of solvent parameters

Parameter	Original solvent	Reformulated solvent
Circulation rate (m <sup>3</sup> /hr)	135	115
Absorber residence time (mins)	2.9	3.4
Gas entrainment in rich solvent (calculated bubble micron size)	500	320
Solvent viscosity (cP)	20	9