Challenges for Processing Oil and Gas from Kazakhstan

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ABSTRACT

In recent years WorleyParsons has been involved in the design of several large (multi billion dollar) oil and gas processing plants in Kazakhstan.

This paper looks at the processing requirements needed for typical sour reservoir fluids in Kazakhstan. These are typified by high hydrogen sulphide levels and modest amounts of carbon dioxide along with organic sulphur. The plants were also at remote locations which along with the difficulty of bringing in large and heavy equipment meant that preference was given to reliable technologies with a proven track record.

A number of options have been considered for designs to meet the oil and gas specifications in terms of acid gas/ mercaptan specification and also for dew point/ vapour pressure. The benefits and drawbacks of each option have been discussed and the reasons for going ahead with a particular design have been explained

INTRODUCTION

The oil and gas streams arriving by pipeline from the wellheads, onshore or offshore, contain sour gas in the order of 14 mol% hydrogen sulphide and 4% carbon dioxide that will tend to concentrate to higher levels in the gas processing area. The inlet oil and gas also contains mercaptans. Mercaptans are undesirable components because of their odour. Methyl and ethyl mercaptan are the most odorous due to their higher volatility.

After crude separation and recompression of flash gases, the feed to the acid gas removal contains about 200 ppmv organic sulphur, which includes mercaptans, carbonyl sulphide and possibly other organic species such as disulphides and carbon disulphide. At this point H_2S concentrations have typically increased to 16 – 19 mol% and CO_2 to 3 – 5 mol%.

The Sales Gas product has a total mercaptan specification of 16 mg/Nm^3 maximum (equivalent to 8 ppmv methyl mercaptan). The Stabilised Crude product specification is in the range 20 to 30 ppmwt maximum of methyl plus ethyl mercaptan. There is a question whether limits should also be set for propyl mercaptan. Propyl mercaptan however, is not as volatile as the lighter mercaptans and consequently does not present a significant odour problem in tankage at the current levels.

Accurate prediction of the mercaptan distribution between the oil and gas phases in the various sections of the plant is important in defining the treating requirements. The Hysys simulation program was used to study the mercaptan distribution in the process scheme using various thermodynamic correlations including Peng-Robinson (PR), Soave-Redlich-Kwong (SRK), Grayson-Streed (GS) and Braun K10 (BK10).

In one of the cases, binary interaction coefficients developed from reservoir fluid analyses were available and were incorporated into the simulations. The results enabled a comparison of the mercaptan distribution to be made with the simulation default properties. In the other, existing plant performance data allowed the adjustment of simulation parameters to match the performance.

It was found that the sections of the plant handling the heavier oil components were best suited to using equilibrium parameters based from the reservoir fluid data or existing operating data. The gas and light ends processing sections were more accurately simulated using the simulation default properties.

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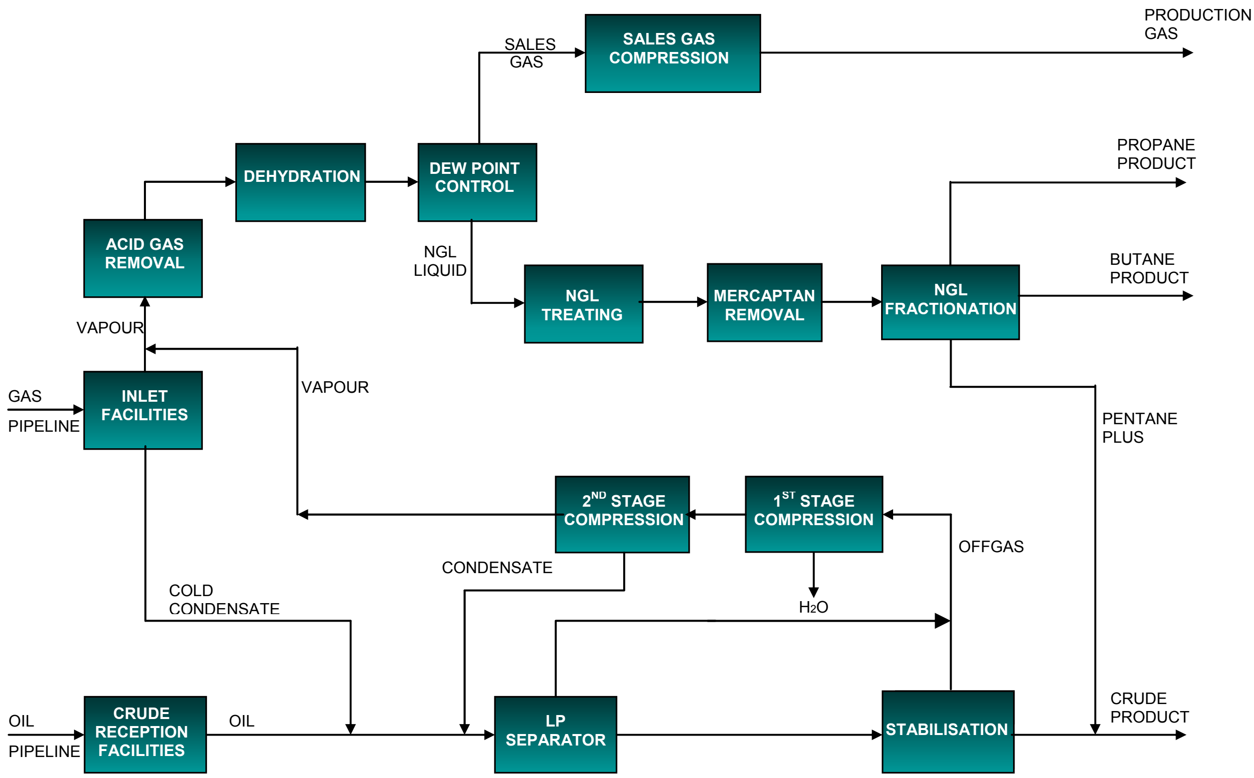
OIL PROCESSING

Mercaptan removal

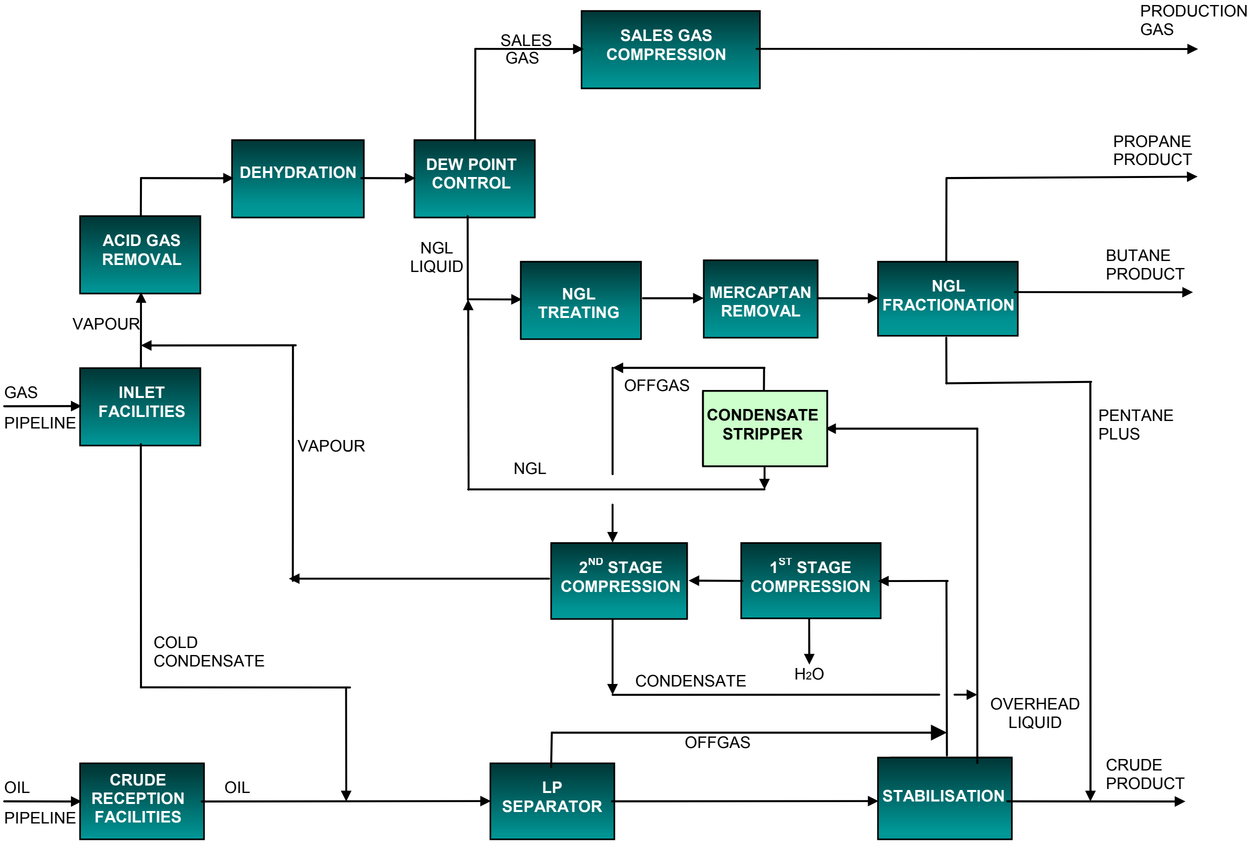
Four main options were considered for reducing the mercaptans in the crude oil processing section. These are illustrated in the following diagrams.

Option 1. **Strip the light mercaptans in the Stabiliser and recycle the overheads through the flash gas compression. Remove the mercaptans in the NGL treater and re-inject heavier NGL into the stabilised crude after NGL fractionation.**



Increasing the Stabiliser boil up to meet the mercaptan specification in the oil raises the amount of heavier components in the overheads to the flash gas compression train. This creates large recycles because the heavier components (propane and butanes) drop out at the inter-stage cooler and are resent to the stabiliser. While this design minimises oil mercaptan treating by sending all of the lighter mercaptans to the NGL treating unit for removal, the reboiler and compression duties entailed are excessive, making the scheme unattractive.

Option 2. **As option 1 but a Condensate Stripper tower is added to take condensates derived from the Stabiliser and the flash gas compressor train and route the resulting Stripper bottoms product directly to the NGL treating. The Stripper overhead gas is sent to amine treating.**

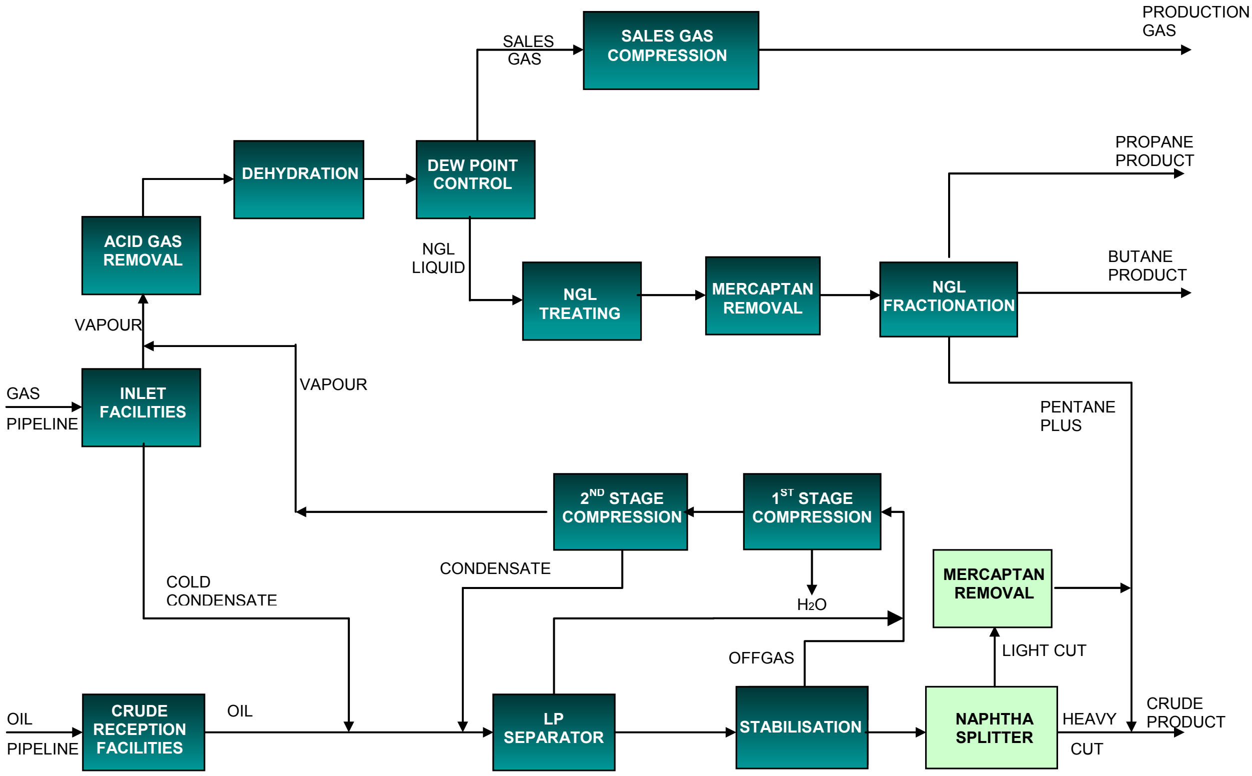


The addition a Condensate Stripper eliminates recycle build up, as the heavier components are then routed directly to the NGL treating area along with the De-ethaniser bottoms. The H2S is stripped from the condensate so the condensate can bypass the AGR. This scheme has the advantage of reducing the dew point of the flash gas. It may be the only workable option if there is insufficient pipeline gas to mix with the flash gas to reduce the dew point of the gas feed to the AGR to prevent hydrocarbon condensation in the amine

The prime drawback of this scheme is that removal of more product overhead at the Stabiliser, as with the previous scheme, raises the Stabiliser bottoms temperature and reboiler duty.

While a number of different variations are possible for this option, the scheme, while readily workable, needs greater Stabiliser duty and an additional tower. One such variation is to use steam stripping for the Condensate Stripper rather than a reboiler. However, it was found that the liquid recycle streams increased which in turn increased the duties of and the Stabiliser reboiler, condenser and flash gas compression

Option 3. **Reduce the stabiliser duty to meet the vapour pressure specification but allow light mercaptans into the stabilised crude. Add a separate Naphtha Splitter downstream to separate a light naphtha overhead product from the crude, which contains the light mercaptans. Treat the light naphtha and blend back into the crude.**

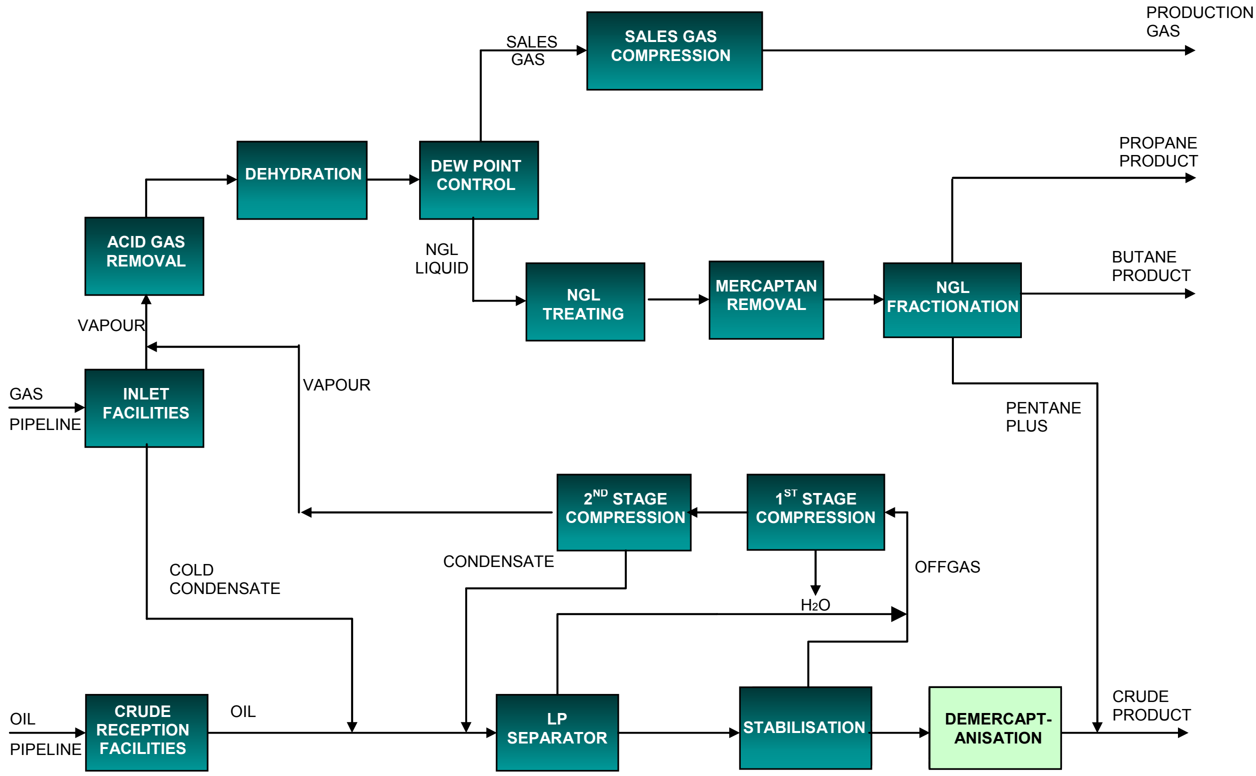


The Stabiliser is run to meet the oil vapour pressure specification and also to minimise the H2S level (to below 1 ppmwt) as H2S will consume the caustic in the downstream mercaptan treating unit. As the stabiliser duty was reduced some light mercaptans were allowed to remain in the oil.

A Naphtha Splitter, downstream of the Stabiliser, is added to cut out a light naphtha stream that will contain the light mercaptans. The naphtha overheads are condensed, treated for mercaptans and blended into the crude product.

This option minimises the overall reboiler duty (stabiliser, condensate splitter and naphtha splitter)

Option 4. Stabilise the crude and treat the whole crude for mercaptan removal.



Similarly to option 3 the stabiliser is used only to stabilise the crude and not to strip off all the light mercaptans

Treating the whole crude for mercaptan removal means that a much larger flow has to be treated and involves treating the heavy oil components simultaneously with the light material. Treating rates will be higher to achieve adequate contact and treating the heavier liquid will make separation and recovery of the oil and caustic phases more difficult. In addition, the crude may well contain components such as naphthenic acids, which will complicate treating, and increase caustic consumption.

Recommended Scheme

Scheme 2 (the Condensate Stripper) was found to be the most attractive scheme for one application and scheme 3 (the Naphtha Splitter) for another. The selections were influenced by differences in feed composition and upstream arrangements i.e. one plant had offshore wells with some crude separation offshore and the other had all separation consolidated at the same location.

Treating the whole crude with a caustic process was considered too cumbersome in terms of caustic recovery, operation and effluent handling/disposal.

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GAS PROCESSING

Acid Gas Removal / Mercaptan treating

There are a number of processes for removing mercaptans from gas streams; however, the gas specification of other contaminants, such as H_2S, CO_2 and other dew point specification, will influence the selection of the most appropriate mercaptan removal method.

1. Solvents
   There are a variety of wet solvents that can partially remove mercaptans but are primarily for H_2S and CO_2 removal.

   1. Chemical solvents e.g. Non-proprietary amines, hindered amines, activated amines. Usually only partial removal
   2. Physical solvents e.g. Selexol, Purisol, Fluor solvent, Ifpex-2. High level of removal.
   3. Mixed solvents (chemical and physical), usually proprietary e.g. Flexsorb, Sulfinol, Ucarsols. Depending on the type, the removal of mercaptan may be partial or total.

2. Solid bed
   1. Molecular sieves – also dries the gas – regenerable - well established.
   2. Chemical adsorbent e.g. Puraspec type – non regenerable - used for trace amounts

3. Scavengers/Inhibitors
   - Used more typically for low levels of H_2S and mercaptan.
   - Limited data available for mercaptans
   - CO_2 consumes the chemical in some formulations.

4. Membranes
   - Well established for bulk CO_2
   - Not usually used for other acid gases.

5. Caustic Wash
   - More typically used for treating hydrocarbon liquids.
   - H_2S and CO_2 will consume caustic.
   - Used downstream of bulk acid gas removal process.

Acid gas removal

As indicated above, there are a number of processes capable of acid gas removal. Typically for the major fields in Kazakhstan the problem is that of bulk H_2S removal down to low levels in conjunction with more modest CO_2 removal. In addition, there is the requirement to remove mercaptans

Bulk removal of acid gas, with H_2S as the dominant component, eliminates all but the solvent processes from the above selection. In addition, the dominance of H_2S also reduces the advantage that selective solvents would provide, as there is modest amount of CO_2 to selectively leave in the treated gas

While there are strong advantages to removing the organic sulphur with the acid gases, this provides a dilemma. Removal of organic sulphur will be favoured by mixed or physical solvents. These will increase the capture/solubility of hydrocarbons in the rich solvent which creates design problems for the Sulphur recovery unit. The extra hydrocarbons in the acid gas will take more air to combust them and increase the size of these already huge units. The rejected effluent will also increase

As the acid gas removal units will be very large, expensive and relatively remotely located, the following key factors are important:

- Removal of organic sulphur components would be an advantage to limit their downstream recovery.
- Excessive pick up of hydrocarbons by the treating solvent will increase the size/cost of the sulphur recovery and tail gas treating units and increase stack effluent discharges.
- Equipment size could be a factor in transportation and supply.
- The process needs to have a proven record, as failure to properly function will have serious economic consequences and quick repairs will be difficult due to the remote location
- Lower heat/power input would reduce utility requirements and ultimately their size and flue gas discharges.
- Established and secure supply train for the solvent/chemicals into Kazakhstan.

The above points lead to a conflict of priorities between straightforward aqueous solvents and their mixed or physical solvent alternatives. In addition, robust and relevant experience is essential.

Large train sizes have been deliberately chosen to reduce capital costs, the resulting size of the sulphur recovery and tail gas treating units is a critical design factor. This counts against those processes that pick up more hydrocarbons into the solvent. Combining this with past experience in Kazakhstan has lead to the selection of DEA as the preferred solvent.

In order to limit the potential disadvantages of DEA in terms of circulation rate and boil up requirements, higher than conventional solvent concentrations and acid gas loadings have been utilised. This means special attention to metallurgical selection and operation within given limits.

The use of DEA means that only partial removal of organic sulphur is feasible. Since this partial removal of mercaptans does not achieve the mercaptan specification, additional processing is required.

Further treating for Mercaptans

Two processing schemes were considered for downstream processing to reduce the amount of mercaptans in the gas.

1. Molecular Sieve

The gas, downstream of the AGR, also requires drying; therefore a potentially viable option is to select a molecular sieve that is also capable of removing mercaptans.

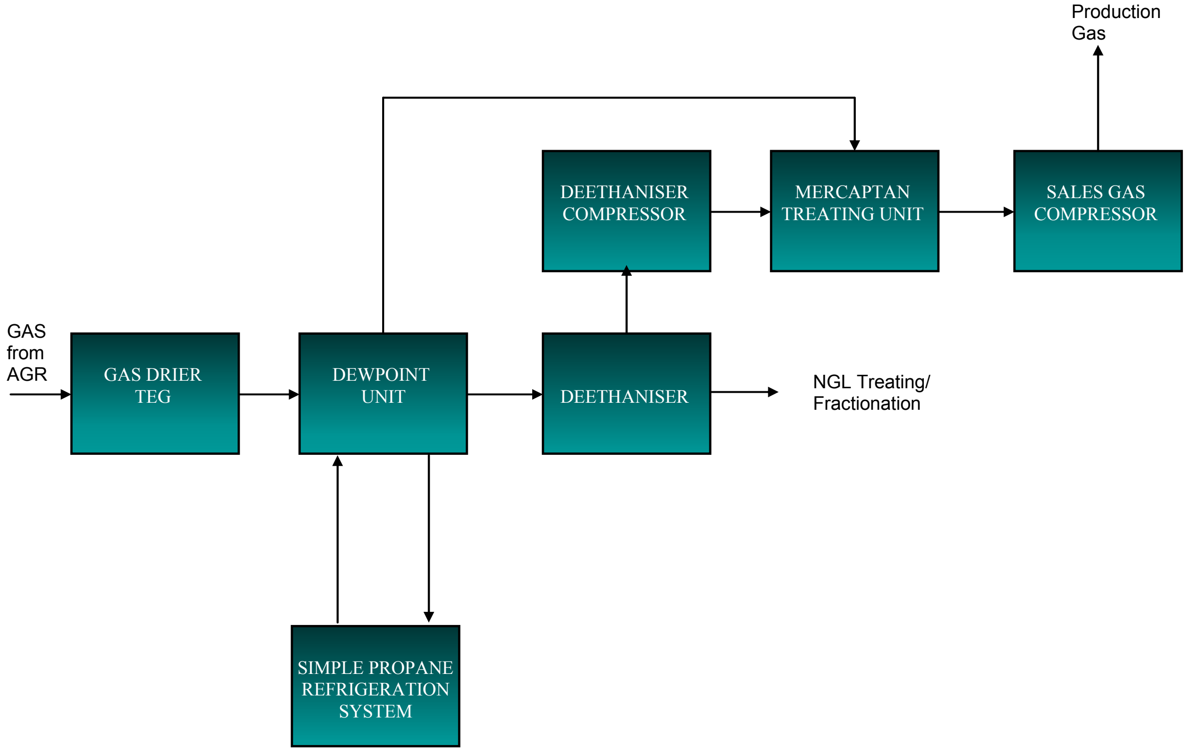
Equipment sizes and duties were obtained from a number of suppliers. A feature of the scheme is that during the regeneration cycle, the regeneration gas can contain over 2,000 ppmv of mercaptans. Consideration was given for a suitable disposal route including use as fuel gas and/or incineration. It was found that further treatment with a mixed or physical solvent would be necessary to capture the mercaptans from the resulting regeneration gas. The disposal route would then be via the solvent regenerator acid gas to the sulphur plant.

The complexity of adding another acid gas treating solvent facility did not favour this option, which when, coupled with the large size of sieve beds needed, was too expensive.

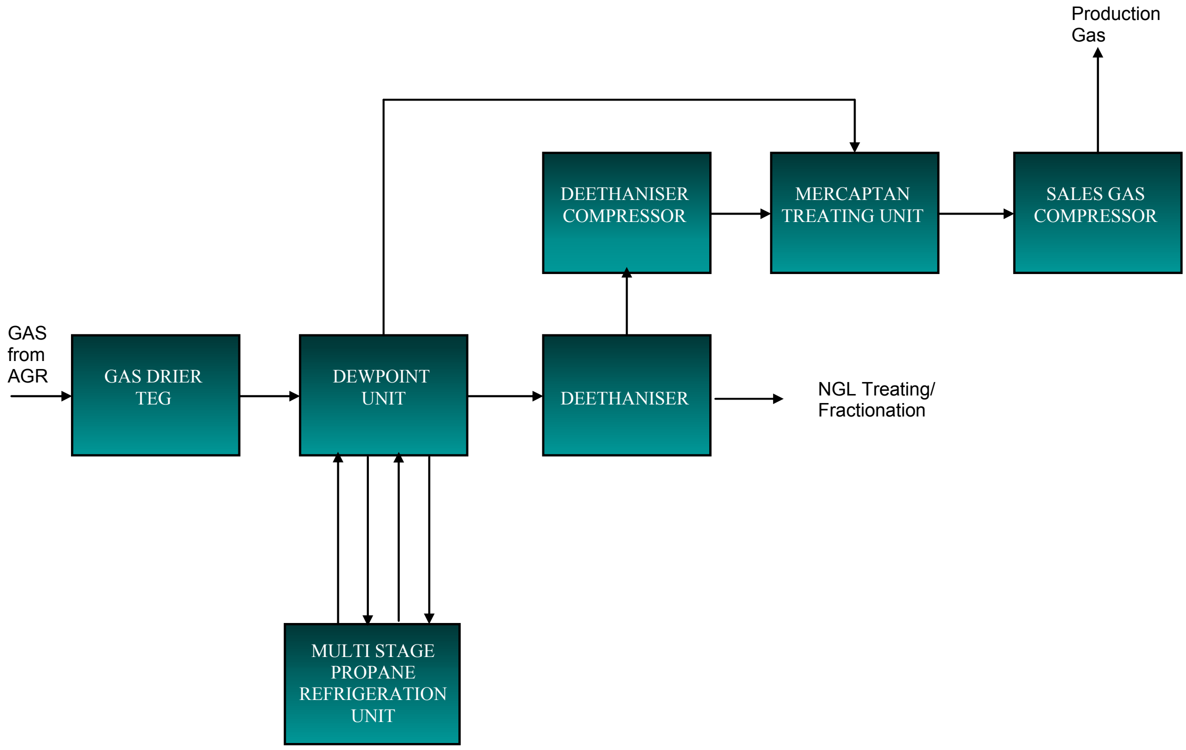
2. Dew Point Control/Liquid Treating

The gas has to meet hydrocarbon dew point specifications for sales gas, seal gas for compressors and fuel for turbines. A study was made for three main options to meet the dew point and mercaptan specifications.

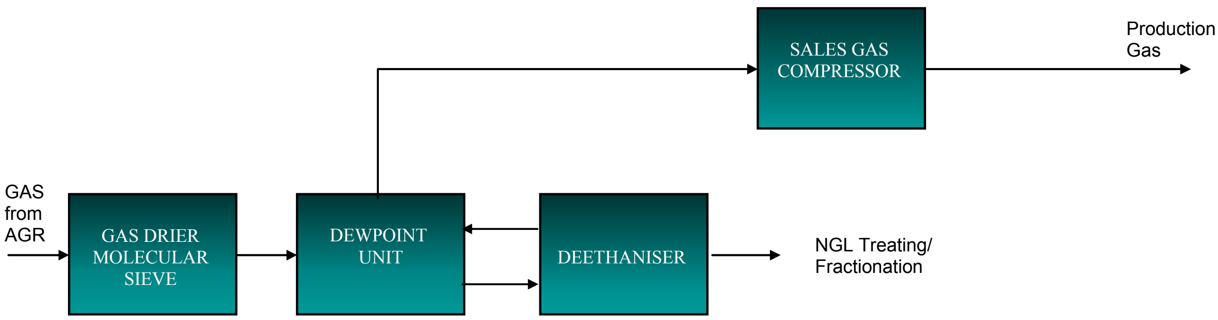
Option 1 - LOW PROPANE RECOVERY OPTION



Option 2 - INTERMEDIATE PROPANE RECOVERY OPTION



Option 3 - HIGH PROPANE RECOVERY OPTION



Option

1. A sales gas dew point specification case that achieves 34% propane recovery using modest mechanical refrigeration via propane.
2. A propane recovery case of 64% utilising mechanical propane refrigeration at the practical limit of application.
3. A 90% propane recovery case as set by an efficient expander configuration.

The first two options use propane refrigeration for chilling the gas. Option 1 utilises a simple single stage concept whilst Option 2 requires a multi stage design with low-level propane chilling the process to – 35°C. Both refrigeration schemes permit retention of gas pressure through the system and hence minimise sales gas compressor power but need a separate booster compressor to raise the pressure of the Deethaniser off gas to production gas levels. Option 3, the high recovery expander scheme, eliminates the need for the intermediate compression for the de-ethaniser offgas but greatly increases sales gas compression needs, as the whole gas is reduced in pressure through the expander. The expander option however, eliminates the need for the mechanical refrigeration system with its extra compression and equipment items.

Increasing propane recovery reduces the level of mercaptans in the sales gas. It can be seen from the following table that only the high recovery case meets the mercaptan specification directly.

Option	Propane recovery	Mercaptans in production gas
1. Low recovery	34%	105 ppmv
2. Intermediate recovery	64%	68 ppmv
3. High recovery	90%	< 5 ppmv

Option 3, therefore, leads to the possibility of eliminating the molecular sieve polishing unit required for controlling mercaptans in the sales gas for the lower recovery cases. It also avoids having to treat the molecular sieve regeneration gas for the peak mercaptan rate.

Option 3 also increases propane recovery. The extra propane recovered could be sold or usefully used as fuel. Of course the whole NGL liquid treating unit will now be larger as a result of the addition propane recovered. The recovered NGL liquids from the dew point unit are next treated for COS and mercaptan removal, the former via hydrolysis and the latter by a caustic based process, before further downstream processing/fractionation.

The following diagrams illustrate the three processing schemes:

Recommended Scheme

The high propane recovery case is the more favourable scheme because it achieves a lower mercaptan content specification without the need for molecular sieve polishing and regeneration gas treatment.

The scheme also offers the potential for high NGL/propane recovery

OTHER ISSUES

Gas Re-injection

Due to the relative isolation of the Kazakhstan facilities and distance from markets the production gas has relatively low value. There is also the problem of the production and storage of thousands of tons of sulphur per day. In the long term, there is the issue of reservoir pressure maintenance. This leads to the possibility of re-injecting sour residual gases rather than processing and exporting them. This raises questions of reservoir acceptability to re-injection, its cost and technical viability.

The required re-injection pressures have typically varied between 550 and 800 bar for the larger developments in Kazakhstan. This has led to a development of re-injection compressor design capability. In addition, the climate in Kazakhstan has high extremes of ambient temperature (+40°C to -40°C) which affect driver performance, i.e. lower power output in summer than winter. While this would only appear to limit re-injection gas flow, more seriously it reduces crude oil production as less off-gas can be compressed.

The key design issues are the overall design and integrity of the injection system, which comprises trunklines, flowlines, injection compressors, and the implied safety and metallurgical consideration. Interfaces with the compressor vendor and layout considerations require specific focus.

The facilities design must have flexibility for the variation of the associated gas production rates and composition with time.

Sulphur Recovery Units

Due to the high sulphur content of the crude oil, the potential sulphur production could be around 6-12 tons of sulphur per 1000 barrels of crude oil produced, depending on case/operation

The number of very large sulphur recovery units (>1500 tons per day of sulphur) installed worldwide is very small. The Kaybob plant in Canada, designed by WorleyParsons, is thought to be the largest unit with a single reaction furnace at 1650 tons per day. This unit was a four reactor Claus with a maximum recovery of around 98%. Examples of larger sulphur trains at Astrakhan, Russia and Shell Carolina, Canada used two parallel reaction furnaces. These units used processes to achieve higher sulphur recoveries; Sulfreen at Astrakhan for around 99% recovery and Claus + SCOT at Caroline for >99.8% recovery.

The new sulphur units for Kazakhstan are large units and have been designed to meet the new industrial environmental standards of >99.9%.

Sulphur recovery units process large volumes of gas at low pressure, resulting at these capacities, in very large equipment. The maximum single sulphur train size is limited initially by transportation constraints, and, if these are removed by manufacturing limitations. Some of the limitations evaluated in the design of these units are discussed below.

Acid Gas Burner/Reaction Furnace/Waste Heat Boiler

This piece of equipment is at the heart of the Claus process and each section has its own limit.

Modern high intensity burners, such as those supplied by HEC and Duiker, have been built for capacities equivalent to about 1200-1300 tons per day of sulphur, increased burner capacities up to 1700 tons per day are not considered a significant risk.

Reaction Furnaces are large refractory lined vessels and care has to be taken in the design of the refractory not to exceed crushing limits at operating temperature. This may limit the diameter but can be compensated for by increasing the length. However the furnace diameter cannot be greatly different from the Waste Heat Boiler tube sheet.

Waste Heat Boiler design is limited by the diameter of tube sheet that can be manufactured. The size limit is based upon the maximum achievable with a single tube sheet weld. Larger tube sheets would require two or more welds which was judged to be potentially a problem to produce a flat tube sheet.

When the combination of factors related to the Reaction furnace and Waste Heat Boiler exceed these engineering limits, a solution is to split into parallel trains.

Sulphur Condensers

Size limitations are again based on tube sheet diameters. The condensers are designed with separate steam drums. Multiple sulphur rundown lines are recommended.

Catalytic Reactors

Multiple inlets and outlets for the catalytic reactors will be required. This implies that careful consideration must be given to the piping design to ensure uniform flow of process gas through the catalyst beds.

TGTU Columns

The Quench Column and Amine absorber are large columns. However, they operate at low pressure and the design is relatively straightforward compared to the Acid Gas Removal Absorber, which is of a similar size but a much higher design pressure.

Piping and Valves

For these size of units the main process lines are very large in diameter. Valves, where required, are very important for trouble free operation; tight shut off butterfly valves will be specified in every case. Specifications for the valves should be well stated and, preferably, only specific vendors allowed. The reason for this is that butterfly valves can be very cheap, and the use of ill specified valves can lead to continuous operating problems occurring.

Actuators for the large control valves should be robust and powerful. The opening and closing of the valves, particularly the emergency shut down valves, must be as quick as possible in order that a large Sulphur Recovery unit will operate as well and as safely as a small unit.

WorleyParsons' investigations into the design of large sulphur recovery units suggest that it would be possible to design and construct a unit with a capacity of up to 3,000 tons per day of sulphur providing there are no transport limitations.

Logistics

Kazakhstan is essentially a land locked country and hence access for large equipment items can be a serious consideration. Clearly there would be a motive for making large processing trains, as this would minimise requirements for plot space, complexity of operation and capital cost.

The routes into the country are effectively limited to road/rail transport and barge transport through the Volga and connecting canals to the Black Sea.

If the plant location is near the Caspian, barges will probably have reasonably close access despite the relative shallow nature of the Caspian at its shoreline. Getting equipment from the barges to the construction site has been studied and appears to be feasible by using large hovercraft

If the plant is significantly inland then greater restriction will be imposed by limitations from road or rail access.

Not withstanding the above, very large items can be transported and the most practical is to use barge transport as far as possible to bring the larger items through the above canal system. The canal systems are limited to summer operation but do permit transport of very large vessels (up to 10 m diameter)