Sour gas cleaning by condensed rotational separation versus amine treatment, energy and volume use

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Comparing the Volume and Energy Consumption of Sour-Gas Cleaning by Condensed Rotational Separation and Amine Treatments

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Natural gas is expected to be an important energy source for decades to come. About 40% of the known natural gas reserves is known to be sour with the acid gases H_2S and CO_2 . 16% is highly sour with more than 20% acid gas.^[1,2] As the energy costs of conventional amine sweetening increase almost linearly with the contamination level, processes are developed that use cooling and selective condensation of the acid components; that is, SPREX by TOTAL and CFZ by EXXON.^[3] The energy use in these processes is greatly reduced compared to conventional amine treatment but the equipment—large distillation towers under pressure—is still voluminous. The newly developed process of condensed rotational separation (CRS) is a compact alternative to the distillation tower.^[3] In this communication we compare the amine treatment and CRS processes, considering energy and volume use for a binary mixture of CH_4 and CO_2 . Details of the assumptions and models can be found in the supplementary report to this communication.^[4]

Condensed rotational separation is based on two innovative ideas. The first innovation is the fast cooling of the gas mixture in the two-phase region by cooling and expansion (EXP1 in Scheme 1), whereby one component becomes a mist of fine droplets and the other remains in the gaseous phase. The second innovation is the rotational particle separator (RPS, and RPS1 in Scheme 1). It is an efficient and compact demister that is able to remove the fine droplets from the gas.^[5] A representation of such a component is shown in Figure 1.

The pressure and temperature at point A in Figure 2 are chosen such that the methane recovery is maximized. The separated liquid CO_2 still contains a considerable amount of methane, however. The liquid produced in the first condensation step is flashed in a second expansion (EXP2 in Scheme 1) to point B in Figure 2 where the liquid CO_2 recovery is maximized.

The contaminated gas produced is compressed (COM in Scheme 1) and re-fed prior to the first step. In this way optimal gas and liquid purity is achieved. For a binary mixture of CH_4 – CO_2 the methane can be purified to a contamination level of 14%, and the CO_2 leaves the process as a liquid of 98% purity.

The compressor is the main energy consumer in CRS. The compressor work scales with the mass flow, which in turn scales with the amount of CO_2 present in the feed stream. Rescaled to the primary energy with a conversion factor of

Scheme 1. Process scheme of a two-step condensed rotational separation. The gas is cooled and expanded over the first expansion stage (EXP1) where by a mist of small droplets are formed. The cleaned gas is separated in RPS1. The liquid is expanded in EXP2 to produce a clean liquid that is removed by RPS2. The gas in the second stage is re-fed prior to the first expansion stage.

Figure 1. Principle of the rotational particle separator. The gas, which contains droplets, enters in a swirling motion. Large droplets are collected in the pre-separator and the micrometer-sized droplets enter the small channels of the rotating element. They form a film that breaks up in large droplets, which are separated in the post-separator.

0.4, this amounts to to 1 MW per % CO_2 for an incoming gas flow of 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$). The size of the compressor is negligible compared to the total installed volume. As the volume of the RPS scales with the volumetric flow,

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Figure 2. Two-step CRS shown in a pressure–concentration phase diagram. The incoming gas is reduced in pressure and temperature to the point where methane recovery is maximized (point A). The liquid is flashed to a point where CO₂ recovery is maximized (point B).

the size of the RPS is mainly dependent on the incoming gas flow (not on the composition) and contributes to about 5% of the total installed volume. The size of the second separator increases with the flow in the recycle ($\approx 20\%$). The heat exchangers are modeled as the multistream, spiral-wound variety commonly applied in liquefied-natural-gas applications and they account for approximately 75% of the installed CRS volume. For a feed gas flow of 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$) the CRS installation requires approximately 1 m^3 per %CO₂.

A simplified diagram for the amine gas-sweetening process is given in Scheme 2. The sour gas is contacted gas in counter flow with an amine solution in an absorption tower. The CO₂ dissolves in the solution and the amines in solution react with the CO₂, creating a CO₂-rich solution. The enriched solution in the absorber flows into a flash tank that is operated at lower pressure, which permits the venting of any entrained light hydrocarbons from the solution. The rich solution is pumped through a product-feed exchanger and then heated in a stripper tower by a reboiler at the bottom of the stripper

Scheme 2. Simplified process scheme for amine treatment.

tower in order to break the chemical bond between amine and CO₂. The CO₂ flows out the top of the stripper and the CO₂-lean amine solution is pumped to the absorption tower.

The primary energy consumption for the amine process originates from the steam furnaces providing steam to the reboiler and reclaimer. The energy use of amine treatment is (almost) linear with the amount of CO₂ in the feed gas stream at 8 MW per %mole for 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$).

The absorber diameter is either scaled with the superficial gas velocity or the diameter is calculated from a flooding constraint, which scales with the liquid flow rate. The height scales with the contamination level and the flow of inert species. However, the bulk of the CO₂ removal contributes to a relatively small portion of the total height of packing. Typically it is only the last few % of CO₂ concentration to be removed from the flow that contribute the most to the total required height of the packing. The reason for this is that the absorption rate is strongly linked to the concentration difference between the gas and the liquid. The height of packing scales linearly with the molar gas flow of inert species. For a fixed inlet flow of 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$) both effects cancel each other and the absorber size is approximately 60 m^3 independent of the contamination. The flow of the amine solution, however, scales linearly with the molar flow of CO₂ in the feed-gas stream. As a consequence the same holds for the other main components: the stripper, flash tank, reboiler, Reclaimer and the reflux drum. Together they require an installed volume of 35 m^3 per %mole for 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$).

For a first comparison of the amine treatment with CRS the goal is separation of CO₂ to 0.7%, which is the US pipeline specification for natural gas. As the CRS process can only go down to a concentration of 13.6%, the amine process is only compared to a combination of CRS + Amine. The CRS process is considered at a feed gas pressure of 70 bar and a temperature of 293 K. The amine process is considered at a feed gas pressure of 40 bar and a temperature of 293 K (equal to the outlet pressure and temperature of CRS). Choosing the operating pressure of the amine absorber is a trade-off between the volume of the absorber and the energy cost of pumping the lean amine from the stripper towards the absorber. Therefore, the operating pressure of the amine absorber is chosen to be 40 bar, regardless of the presence of CRS equipment.

In Figure 3 a plot is shown of the absolute value of the available energy and the energy cost of the processes together with the higher heating value of the incoming gas. Amine behaves as expected; its energy cost rises linearly with the CO₂ content of the gas field, even consuming more than 100% of the available energy in the case of CO₂ concentrations over 65%. As the energy penalty of CRS is a factor of 8 lower than for the amine process, using CRS as a bulk separator above concentration levels of 15% lowers the total energy consumption, as the flow to the amine plant is reduced. The reduction in volume due to this effect is even larger, see Figure 4.

Figure 3. A plot of the available energy content in the incoming gas stream and the energy cost of the amine process as well as the CRS + Amine process. Valid for a flow of 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$) and binary composition of CO_2 and CH_4 .

The absolute values of the energy and volume consumption of both processes are dependent on the assumptions made for process conditions, heat integration, required recovery, etc. Nevertheless we draw the conclusion that CRS and amine absorption complement each other very well. Amine absorption has the capability to bring the gas purity up to pipeline specification, whereas CRS has a resounding advantage over amine in terms of energy cost and equipment size. The combination of both technologies allows the exploitation of previously uneconomical gas fields.

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Figure 4. A plot of the installed volume of the amine process and the CRS + Amine process. Valid for a flow of 125 MMscf per day ($40 \text{ Nm}^3 \text{ s}^{-1}$) and a binary composition of CO_2 and CH_4 .

Keywords: absorption • carbon dioxide • distillation • hydrocarbons • natural gas

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