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Acid gas removal in geothermal power plant in Iceland

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Abstract

A large part of the energy covering the electricity and heating demands in Iceland is generated in geothermal power plants. The Hellisheiði power plant, designed for 300 MW_e and 133 MW_{th}, is located in close proximity to Reykjavik. The concept of the plant is to co-generate power for energy-intensive industry and hot water for district heating. The steam at Hellisheiði is not pure H₂O, but also contains H₂S, CO₂, H₂, N₂, and CH₄. These gases have, for the most part, been emitted to the atmosphere after separation from the steam. New, emerging environmental regulations in Iceland will limit the emission of H₂S. Additionally, the long-term goal is to decrease CO₂ emissions. Therefore, separation of CO₂ and H₂S from the non-condensable gases in the steam will be necessary, followed by some measure to store these. In this work, four different acid gas capture systems were selected and subsequently modelled and simulated: water absorption, amine absorption with MDEA, amine/low temperature hybrid concept, and stand-alone low-temperature separation. For co-removal of H₂S and CO₂, low-temperature separation seems to be an attractive alternative to the conventional water absorption process due to the low power penalty.

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1. Introduction

A large part of the energy covering the electricity and heating demands in Iceland is generated in geothermal power plants. The Hellisheiði power plant is located in Iceland around 20 km from Reykjavik. The concept of the

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plant is to co-generate power for energy-intensive industry and hot water for district heating from steam extracted from the ground. The plant consists of six 45 MW_e high-pressure turbines (HPT) and one 33 MW_e low-pressure turbine (LPT) for a total of around 300 MW_e. As of 2014 the thermal capacity is 133 MW_{th}, but plans exist for increasing this to 400 MW_{th}. The power plant has 30 production wells located in an 8 km² area around the power plant [1].

The steam at Hellisheiði is not pure H₂O, but also contains H₂S, CO₂, H₂, N₂, and CH₄. As of 2014, these gases are, for the most part (some are captured in a gas separation station for testing), emitted to the atmosphere after separation from the steam. New, emerging environmental regulations in Iceland will limit the emission of H₂S [2]. Additionally, Iceland's long-term goal is to decrease CO₂ emissions. Therefore, separation of CO₂ and H₂S from the non-condensable gases in the steam will be necessary, followed by some measure to store these. Two experimental gas re-injection projects, SulFix and CarbFix, address the storage aspect [2,3]. The current disposal method tested is in-line water dissolution from the wellhead down. Water and gases are mixed slightly below the gas inlet. The residence and dissolution time in the vertical injection pipe is about 5 minutes, as the gases and water are transported down the well.

The focus of this paper is to suggest and evaluate capture technologies for the H₂S and CO₂ contained in the geothermal steam for the Hellisheiði plant.

2. Hellisheiði process description

A detailed schematic of Hellisheiði can be found in [1]. From the production wells, the saturated steam is transported to gas liquid separators with a pressure of 10 bar; the liquid is pumped to the low-pressure liquid–gas separator which operates at 2 bar. The gas from both the high- and low-pressure liquid gas separators is taken to moist separators to prevent liquid from entering the turbines. After the HPT, the steam is condensed. The condensers preheat the fresh water for the district heating system. Because the fresh water is saturated with dissolved oxygen and becomes corrosive when heated, the heated water is deaerated before leaving the plant. The non-condensable gases (NCG) are extracted from the condensers. The NCG from this power plant vary in composition between the different turbines, but all streams contain CO₂, H₂S, CH₄, H₂ and N₂. The mass fraction varies from 59% CO₂ to 83% CO₂. The aggregated gas component mass flows of the incoming geothermal steam are shown in Table 1.

Table 1. Component mass flows of the Hellisheiði geothermal steam. Numbers based on [1,4,5].

H ₂ O	CO ₂	H ₂ S	H ₂	N ₂	CH ₄	Total
[kg/s]	[kg/s]	[kg/s]	[kg/s]	[kg/s]	[kg/s]	[kg/s]
1177.7	1.61	0.67	0.017	0.03	0.0027	1180.0

As aforementioned, there are two projects for acid gas re-injection in Iceland: CarbFix and SulFix, both of which focus on pumping CO₂ and/or H₂S into the ground. Tests done in the CarbFix project, where pure CO₂ dissolved in water is re-injected, have been successful. The CO₂ have been injected at a depth of 400 m. The SulFix project started in January 2013; here CO₂ and H₂S are injected with water to a reservoir below 800 m depth. The power plant is located in an area where much of Reykjavik's drinking water is stored. Because of this, care has to be taken when selecting the type of H₂S and CO₂ capture method.

3. Description of acid gas capture methods

There are many different H₂S and CO₂ capture methods used in the industry [6]. Some of those have been considered for H₂S capture for geothermal power plants [7,8]. Four different capture systems were selected and subsequently modelled and simulated in this work. The selection was deemed a good mixture between available commercial methods (water and amine absorption) with more novel technologies (cryogenic and hybrid concepts).

An important selection criterion was the expected ability of the method to capture essentially all H₂S and a large portion of the CO₂. The four selected methods were:

- 1) Water absorption. This is the reference process and tested on site;
- 2) Amine absorption with MDEA;
- 3) Amine/low-temperature hybrid concept;
- 4) Stand-alone low-temperature ("cryogenic") gas separation.

All process simulations were performed using Aspen HYSYS version 8.3 and the following principal assumptions apply for all cases:

- The gas streams from all seven steam turbines are mixed to form a single, aggregate NCG stream;
- 1 atm NCG feed pressure (i.e. outlet of vacuum pump);
- 5°C fresh water temperature;

The actual disposal of H₂S and CO₂ products were not further considered in the process simulation work, but as described above there are various options for handling of these separation product streams, most likely water dissolution.

3.1. Water absorption

According to the research conducted in the CarbFix and SulFix projects, the two gases CO₂ and H₂S have to be dissolved in water in order to be injected back into the ground [2,3]. An absorption process consists of a contactor in which the specific gases are absorbed by a solvent and a stripper in which the gases are stripped from the solvent. The gases leave the top of the stripper whereas the solvent leaves the bottom and is recirculated to the absorber. However, in the special case where a solution of water and gases is injected directly into the ground in the stripping part of the process is not needed.

The solubility of CO₂ and H₂S are dependent on pressure and temperature. The lower the temperature and the higher the pressure, the better the solubility, which means less water is required to absorb the gases. An absorber pressure in the range of 15–25 bar could lead to a good trade-off between water usage and compression/pumping power. The increase in pressure brings a higher duty for the water pump as well as a higher power for the NCG compressor. A simplified flowsheet of the water absorption method is shown in Fig. 1.

The NCG are cooled before entering the compression train. The water for the absorber is pumped to the targeted pressure before entering the top of the absorber. Packing material with a diameter of 0.09 m was assumed in the absorber. The use of packing material increases the surface area and thus lowers the demand of water. 10 stages were assumed in the absorber model and the Predictive Soave-Redlich-Kwong property method was used for the water absorption simulations.

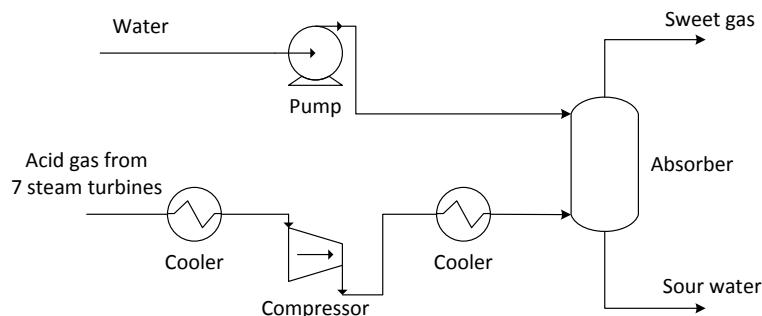


Fig. 1. Simplified water absorption flowsheet.

3.2. Amine absorption (MDEA)

The MDEA unit for acid gas removal (Fig. 2) was first simulated assuming equilibrium conditions. However, as H₂S-selective absorption processes in actuality may operate in non-equilibrium, more CO₂ can pass through the absorber without reaching equilibrium in reaction with the solvent. This mode of operation was investigated in rate-based simulation mode.

As for the water absorption process, the process includes compression of the NCG stream, in this case to 20 bar. The absorber in the equilibrium-based model was simulated with 15 theoretical stages using a 42 wt-% MDEA solution. The rich MDEA loading, for H₂S and CO₂ in aggregate, leaving the bottom stage is 0.55, while the lean MDEA loading is 0.019. The stripper column operates in the range of 1.2–1.5 bar with a reboiler temperature of around 114°C. Preheating of the rich solvent stream is carried out in a liquid–liquid heat exchanger configured to operate with a minimum temperature approach of 15°C. In the non-equilibrium simulations the desorber is operated in the range of 1.5–1.75 bar pressure and the reboiler temperature is around 117°C. The lean and rich loading of the solvent equals 0.044 and 0.53, respectively.

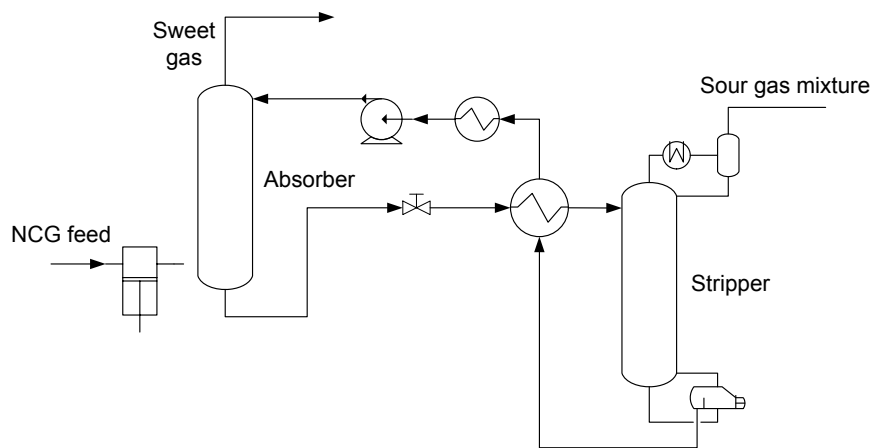


Fig. 2. Amine absorption acid gas separation scheme.

3.3. Amine/low temperature hybrid concept

In addition to a stand-alone amine scrubbing process, a two-stage hybrid gas separation concept (Fig. 3) was also investigated for the NCG. The first stage is made up of an MDEA chemical absorption process, similar to the stand-alone process, to selectively remove H₂S from the feed gas. The sweetened CO₂-rich gas from the MDEA absorber is passed on to a CO₂ condensation unit in which the main portion of CO₂ is separated from the more volatile components H₂, N₂ and CH₄. The gaseous product from this separation stage has high hydrogen concentration, which may be further processed to produce high-purity hydrogen for use in the immediate area, for instance as energy carrier for fuel cell vehicles.

For high initial CO₂ concentrations such as those obtained for the sweet gas (57 mol-% and 71 mol-%, for equilibrium and non-equilibrium scrubber simulations, respectively), condensation and phase separation is an effective bulk removal process for CO₂. Prior to entering the low-temperature separation unit, complete dehydration is required, typically by molecular sieve adsorption. Subsequently, the dry and sweet gas is assumed to be compressed to around 60 bar and cooled to a separation temperature between -56°C and -55°C. Depending on the

exact CO₂ concentration, pressure and separation temperature, 90–95% of the CO₂ is condensed and separated in the liquid phase, while a hydrogen-rich gaseous product is extracted from the top of the phase separator.

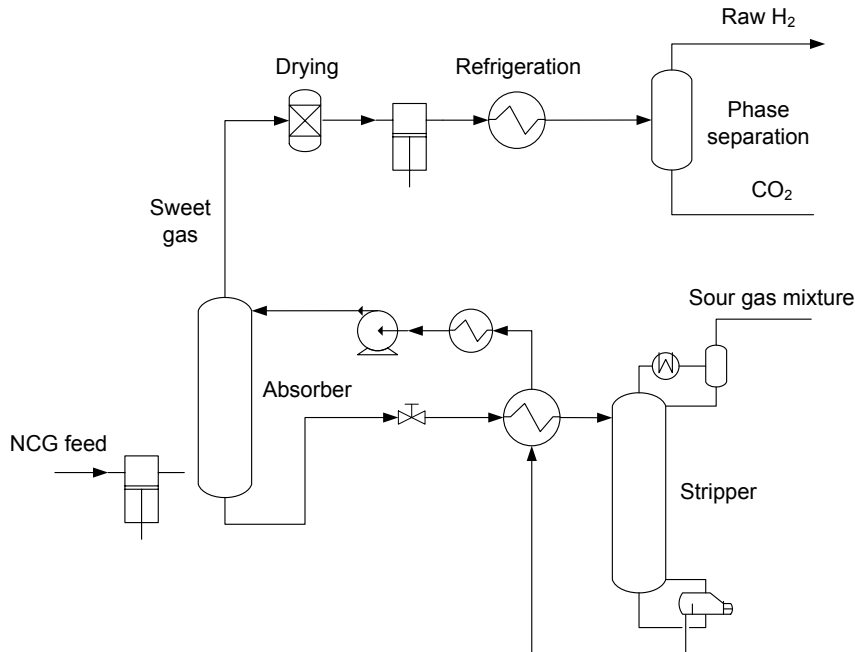


Fig. 3. Simplified hybrid two-stage acid gas separation scheme.

The separation processes have not been optimised to any specific criteria in this work. The choice of process and process sequencing will be subject to preferred separation product specifications (purity/composition, pressure, phase, etc.). This applies in particular to the H₂S and CO₂ products. There seem to be two typical options for the hydrogen stream – combustion/flaring/purging or purification and sale – and the option of hydrogen purification and sale will be dependent on the processing cost in relation to commercial value. Fuel-grade hydrogen is likely to be of no particular value at Hellisheiði due to the very large amounts of available geothermal heat.

3.4. Stand-alone low-temperature ("cryogenic") separation of H₂S and CO₂

Direct cooling and phase separation of the NCG can be an interesting alternative if re-injection of a mixed H₂S/CO₂ product is acceptable. The liquid separation process will be more or less identical to the low-temperature process in the hybrid concept, and the liquid product will consist of mainly CO₂ and H₂S while the volatile components will remain in the gaseous separation product. A possible process configuration is shown in Fig. 4. Since CO₂ and H₂S are captured in liquid phase, pressurisation of this stream is obtained by liquid pumping, which is very energy efficient.

Table 2. Stream compositions for the two simulated cases of the two-stage hybrid gas separation scheme (MDEA + low-temperature).

Case		Amine/low-temperature hybrid case process streams									
		Amine absorption case process streams									
Unit		NCG feed gas		Sweet gas		Sour gas mix		CO ₂		Raw H ₂	
Flowrate	kmol/h	206.1	206.1	72.5	106.6	143.0	105	37.1	71.2	35.3	35.2
Temperature	°C	25.0	25.0		26.12		40				
Pressure	bar	20.5	20.5	20.0	19.5	1.2	1.5	7–100+	7–100+	57	57
Composition											
H ₂ O	mol-%	0.0	0.0	0.21	0.19	6.75	5.0	0.0	0.0	0.0	0.0
CO ₂	mol-%	55.18	55.18	57.10	70.74	50.71	36.51	99.81	99.80	12.48	12.47
H ₂ S	mol-%/ppm	29.74	29.74	0.01	0.02	42.43	58.44	179 ppm	285 ppm	19 ppm	30 ppm
CH ₄	mol-%	0.26	0.25	0.72	0.48	0.0	0.0	0.07	0.07	1.41	1.31
N ₂	mol-%	1.62	1.62	4.59	3.12	0.01	0.0	0.08	0.08	9.36	9.27
H ₂	mol-%	13.21	13.21	37.36	25.45	0.09	0.04	0.02	0.02	76.75	76.95

Equilibrium absorber simulations	Non-equilibrium absorber simulations
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The energy requirement for the amine absorption cases is shown in Table 3. The main driver for power requirement is that for the front-end sour gas compression from 1 atm to 20 bar, which amounts to around 0.7 MW. The remainder of the power requirement is caused by the solvent circulation pump. In addition to power, 4.6 and 3.0 MW of regeneration heat was required for the equilibrium and non-equilibrium case, respectively.

4.3. Amine/low temperature hybrid concept

The product streams and energy requirement for the amine scrubbing stage of the hybrid process are identical to those for the non-equilibrium case given in Table 2 and Table 3, respectively. The additional power requirement in form of compression of the sweet gas to 60 bar and for generating the cooling duty in the low-temperature process is modest, and amounts to roughly 0.2 MW. The total energy requirement is shown in Table 3 and includes pressurisation of the CO₂ product to around 110 bar by liquid pumping. It should also be noted that thermal energy required for gas drying before the cooling process is not included in any of the low-temperature results.

4.4. Stand-alone low-temperature/cryogenic processing

In this process configuration both H₂S and CO₂ will be captured in the same liquid-phase product stream while the volatile components remain in the gaseous phase.

Hydrogen is a potentially valuable byproduct which is available at 70–80% purity in the gaseous phase. At this purity it is likely of no particular value, but if a pressure-swing adsorption (PSA) purification unit is added to the process, hydrogen with very high purity can be supplied at a rate of roughly 1 ton per day. This product could for instance be supplied to filling stations for fuel cell vehicles. The H₂S+CO₂ separation ratio depends on the separation pressure. The impact separation pressure is having on purities of the liquid phase (H₂S+CO₂) and gaseous components (H₂, N₂, CH₄), as well as H₂S/CO₂ separation ratio, is shown in Fig. 5. For two selected example cases, 17.5 and 37.5 bar separation pressure, overall results are summarised in Table 3. The estimated power penalty is around 1.0 MW and 1.15 MW, respectively. It is important to emphasise that these power figures include compression of H₂S and CO₂ to 110 bar. In addition, the raw hydrogen is available at high pressure and can be further purified, flared or purged.

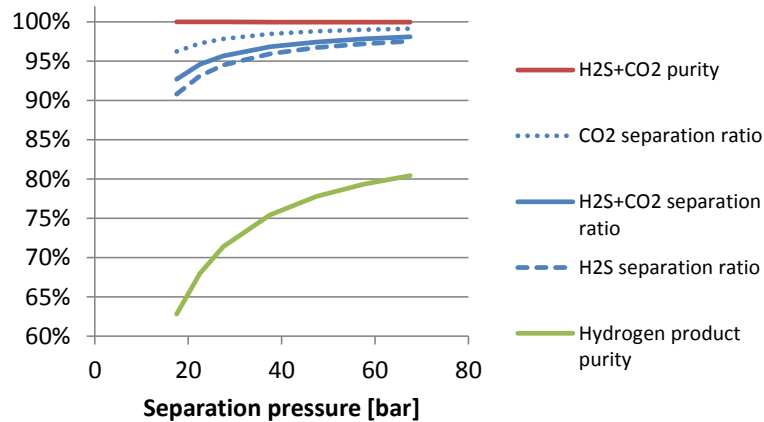


Fig. 5. Results for low-temperature separation of H₂S and CO₂ from non-condensable components (H₂, N₂, CH₄).

4.5. Summary of results

For the four different separation processes alternatives, overall results are summarised in Table 3. As already indicated, the CO₂ and H₂S delivery state differs between the process alternatives, but the disposal method is not further considered in this work. Hence, power figures are not directly comparable without defining the delivery state and modifying the processes in accordance. This matter is further complicated when including steam consumption for the amine process and hydrogen output from the low-temperature process.

Table 3. Key results for the different capture methods. The non-equilibrium MDEA simulations were used as basis for the low-temperature part in the hybrid method.

Capture method	Pressure [bar]		Separation ratio ^a [%]		Power penalty [MW]	Thermal energy [MW]
	Absorber	Low-temp. separator	H ₂ S	CO ₂	Total	Reboiler
Water absorption	15	-	99.0	78.5	0.75	-
Amine absorption Eq.	20	-	98.9	63.7	0.8	4.6 (114°C)
Amine absorption Non-eq.	20	-	99.9	33.6	0.7	3.0 (117°C)
Amine/low-temp hybrid	20	57.5	99.9	96.0 ^b	0.9	3.0 (117°C)
Low-temperature	-	17.5	90.8	96.2	1.0	-
Low-temperature	-	37.5	95.9	98.5	1.15	-

^a Relative to feed

^b 33.6% (entrained in H₂S product) + 62.4% (99.8% pure CO₂ product stream at 100+ bar pressure)

5. Conclusions

Different technologies can be used to remove H₂S and CO₂ from non-condensable gases in geothermal power generation. Of these technologies, the following options have been simulated in order to estimate energy requirement and separation capabilities: water absorption (reference process currently being tested), MDEA absorption, MDEA in combination with low-temperature ("cryogenic") separation, and stand-alone low-temperature separation.

Simulation results for water absorption correspond well with Hellisheiði test data. An absorber pressure of 15 bar was considered a reasonable trade-off between compressor and pumping power requirements. A water/gas massflow ratio of 26.5 resulted in 99% H₂S removal, 78.5% CO₂ removal and inflicted a power penalty of about 0.75 MW. For MDEA absorption, 20 bar was selected as absorber pressure. 99.9% H₂S removal was obtained along with 33.6% CO₂ removal with an energy penalty of 0.7 MW power and 3.0 MW heat for 117°C reboiler temperature. This penalty figure does not include compression of the acid gas from the desorber pressure (around 1.5 bar). In the amine/low-temperature hybrid process the energy requirement was equal to the MDEA case, with an additional 0.2 MW power penalty from the low-temperature unit. In addition to the H₂S and CO₂ removal from the MDEA process, an additional 62.4% CO₂ was removed in a separate stream at high pressure. A hydrogen stream of 77% purity and 57 bar pressure is a potentially valuable byproduct provided that further purification is carried out, for instance by PSA. In the stand-alone low-temperature case, 95.5% H₂S removal and 98.5% CO₂ removal were obtained with a power penalty of 1.15 MW, including compression of the H₂S/CO₂ stream to 110 bar. A hydrogen stream of 75% purity and 37 bar pressure was also produced, which can be upgraded by PSA.

For high co-removal of H₂S and CO₂, low-temperature separation seems to be a possible alternative to the conventional water absorption process due to the relatively low power penalty. Ultimately, the energy estimates provided in this work must be complemented by cost estimations in order to get a more complete basis for comparing the different technologies.

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