

**Screening of H₂S Abatement Options for
Geothermal Power Noncondensable Gas
at Bjarnarflag**

Darryl L. Mamrosh P.E. (darryl.mamrosh@trimeric.com),
Kenneth E. McIntush P.E. (ken.mcintush@trimeric.com), and
Carrie Beitler (carrie.beitler@trimeric.com)
Trimeric Corporation
P.O. Box 826, Buda, TX 78610
512-295-8118

Sigurður H. Markússon (sigurdurm@lv.is), and
Kristján Einarsson (kristjanei@lv.is)
Landsvirkjun
Haaleitisbraut 68
Reykjavik, Iceland 103
354-515-8900

Abstract

Non-condensable gas H₂S abatement technologies were screened – based on capital and operating costs and some other factors – for implementation at new geothermal power production locations at Bjarnarflag, Iceland. Many technologies exist for removing H₂S from gases, but initial screening efforts resulted in narrowing down the available technologies to four likely to be most suitable – water absorption and injection, the modified Claus process, liquid redox sulfur recovery processes, and a burn/scrub process. These four technologies were subject to economic analysis in order to help determine the best technology for Bjarnarflag. This paper presents the results of the evaluation. The analysis showed that water absorption had significantly superior economics for this application; water absorption is, however, associated with the greatest number of unknowns. The potential economic advantage of the water absorption process was considered significant enough to merit recommending further development of the technology.

Keywords: noncondensable gas, NCG, H₂S abatement, hydrogen sulfide removal, sulfur recovery, Iceland

Introduction

Landsvirkjun is developing a new geothermal power production location at Bjarnarflag. The steam at this location contains a substantial amount of hydrogen sulfide (H_2S). Because of H_2S emission regulations under consideration by the Icelandic government, Landsvirkjun is considering the addition of H_2S abatement systems for the non-condensable gas (NCG) from the new power plant(s). Landsvirkjun contracted Trimeric Corporation to screen the available technologies for H_2S removal in order to help Landsvirkjun narrow down the available technologies to those most appropriate for this application.

Criteria that were considered in this analysis to screen the technologies included: capital and operating cost, level of development or commercialization, amount of chemicals to be imported to Iceland, amount and characteristics of waste materials, availability of local resources, level of H_2S removal achieved, emissions of residual H_2S and other species (e.g., methane), reliability, and complexity of operation.

The location of the new facility may result in some of the above criteria being more or less important than they might be in many other areas of the world. Particularly, the need to import large quantities of chemical feed and/or to export products would be considered a negative for this application; also the location has an ample supply of cold ground water available for use, and the disposal of some aqueous liquids by subsurface injection may be possible.

There are a large number of technologies that have been developed for the removal of H_2S from gas streams. However, many were ruled out for technical feasibility issues, environmental concerns, or economic considerations. Only the details of the technologies that were potentially relevant to the Landsvirkjun application are discussed in this paper. The sources for capital and operating cost information for the key technologies are summarized, followed by an economic comparison and description of important characteristics for the different processes.

Design Basis

The design basis for the H_2S removal systems is based on Landsvirkjun's projected steam flow rate, NCG content, and NCG composition. The parameters are given in Table 1. Also included in Table 1 are some economic parameters that were used in comparing the H_2S removal technologies. A 95% H_2S removal requirement was used as the design basis in this work.

The "installation factors" given in the table are the multipliers, based on prior experience at US locations, that were used to escalate the base major equipment cost to an installed plant cost. The "installation factor for base equipment" was used when a capital cost estimate was based on estimated costs for the individually purchased major equipment. The "installation factor for packaged equipment" was used when an estimate was based on costs for modular / packaged equipment. The installation factors are meant to provide a consistent measure of costs to go from purchased major equipment and/or packaged equipment to total installed capital cost. While it is possible that higher factors than those used here may be required due to the remote Icelandic location, the goal of this study was to evaluate all technologies on a basis that was

consistent enough to allow discrimination between technology choices. The use of the factors was not intended to predict actual installed costs with any accuracy.

Table 1. Design Basis Parameters for Screening H₂S Removal Methods

NCG Composition		
	Normal, vol %	Adjusted*, vol %
CO ₂	34	33.4
H ₂ S	23.5	23.1
H ₂	37.4	36.7
CH ₄	1.6	1.6
N ₂	3.6	5.0
O ₂	0	0.3
Steam and H₂S Characteristics		
	Units	Value
Steam Flow Rate	kg/s	84.4
NCG Flow Rate	kg/s	0.5
H ₂ S Flow Rate	kg/s	0.160
Sulfur Equivalent H ₂ S Rate	tonne/day	13
H ₂ S Removal Requirement	%	95
NCG Characteristics		
	Units	Value
Pressure	Bar (absolute)	1.03
Temperature	°C	38
Economic Parameters		
	Units	Value
Assumed plant on-line time	%	96
Capital depreciation		7 yr straight line
Installation Factor for packaged equipment		2
Installation Factor for base equipment		4
Caustic cost	\$/tonne	818
Electricity value	\$/kWh	0.036
LRSR sulfur disposal cost	\$/tonne	115
Claus sulfur disposal cost (value)	\$/tonne	0
Cost for 1 operator (loaded)	\$/year	100,000

* NCG composition was adjusted to account for typical minimum air in-leakage in vacuum systems

Landsvirkjun had tentatively elected, prior to this H₂S abatement technology selection work, to use indirect condensers after the power plant turbines. The choice of condenser can have some impact on the choice of H₂S removal technology, particularly in locations that have ammonia in the NCG. When direct contact condensers are used a higher fraction of the H₂S may be captured in the aqueous condensate; the partitioning of the H₂S to the aqueous condensate phase can become very significant when there is also ammonia in the NCG due to acid-base interactions of the hydrogen sulfide and ammonia. When the partitioning of H₂S to the aqueous condensate phase is significant, the condensate may also have to be treated for H₂S removal, which is referred to as “secondary H₂S abatement”. Because of the Icelandic geothermal steam generally lacks any appreciable ammonia, the partitioning of H₂S to the gas phase in the condensers should

be high regardless of whether direct or indirect condensers are used. Therefore, the choice of condenser type plays a much smaller role in Icelandic H₂S Abatement choices than it may at some other locations.

Technologies Evaluated

Numerous technologies were considered for H₂S abatement at Bjarnarflag, including: water absorption and injection, the modified Claus process (and other gas-phase oxidation processes), liquid redox sulfur recovery processes (e.g., Stretford, LO-CAT, others), burn/scrub processes, burn and vent, scrubbing H₂S with caustic, scrubbing with other alkaline earth minerals, wet sulfuric acid process (WSA) and similar processes, direct acid gas injection, Paques/Thiopaq, ThioSolv, Biox, other redox processes, and amines and physical solvents. Additional consideration of the WSA process was also conducted subsequent to the study, but the additional information did not affect the order of preference of the technologies.

Of these, the following four technologies were concluded to have the best fit for the application at Bjarnarflag: water absorption and injection, the modified Claus process, liquid redox sulfur recovery processes, and a burn/scrub process. A schematic and a brief process description is provided as background for each of these four options. For convenience, separate discussions are given for the two main groups of liquid redox sulfur recovery processes: iron based and vanadium based.

Water Absorption and Injection

Water is not typically used as an H₂S absorbent, because there exist a number of other solvents (e.g., chemical solvents, physical solvents, etc.) that have a much higher affinity for H₂S. However, H₂S has some solubility in water, and the solubility at high pressures can be substantial. Further, the Icelandic location may have a unique situation in that an ample amount of cold fresh water is readily available, and it may also be feasible to inject water that is loaded with H₂S into a local injection well that is capable of handling a very high water injection rate. A diagram of a water absorption and injection scheme is shown in Figure 1.

The water absorption process considered in this study was a simple configuration where water is used in a once-through absorption process, absorbing the H₂S and some of the CO₂ from the NCG. The absorption is done under substantial pressure (~48 bar). The water loaded with H₂S and CO₂ will be acidic (pH 3-4) and will have a very high vapor pressure. The sour water would have to be pumped directly to the injection location.

Depressurization of the sour water would result in the release of the majority of the absorbed H₂S and CO₂, which could represent a significant safety issue. A pressure of 48 bar was used since it is lower than the expected pressure at the injection location (~55 bar) and it is a pressure at which an excessive volume of water is not required to absorb H₂S. Process simulations were done (using WinSim DESIGN II) to characterize the amount of water required for the H₂S absorption as a function of absorber pressure (Figure 2). The temperature of the injection location is also important as the required pressure to avoid evolution of H₂S gas in the well bore

is much higher if the injection zone is hot. The capacity of the injection well(s) sets the maximum water rate that can be used for the absorption.

Figure 1. Schematic of the Water Absorption Process for H₂S Removal from NCG

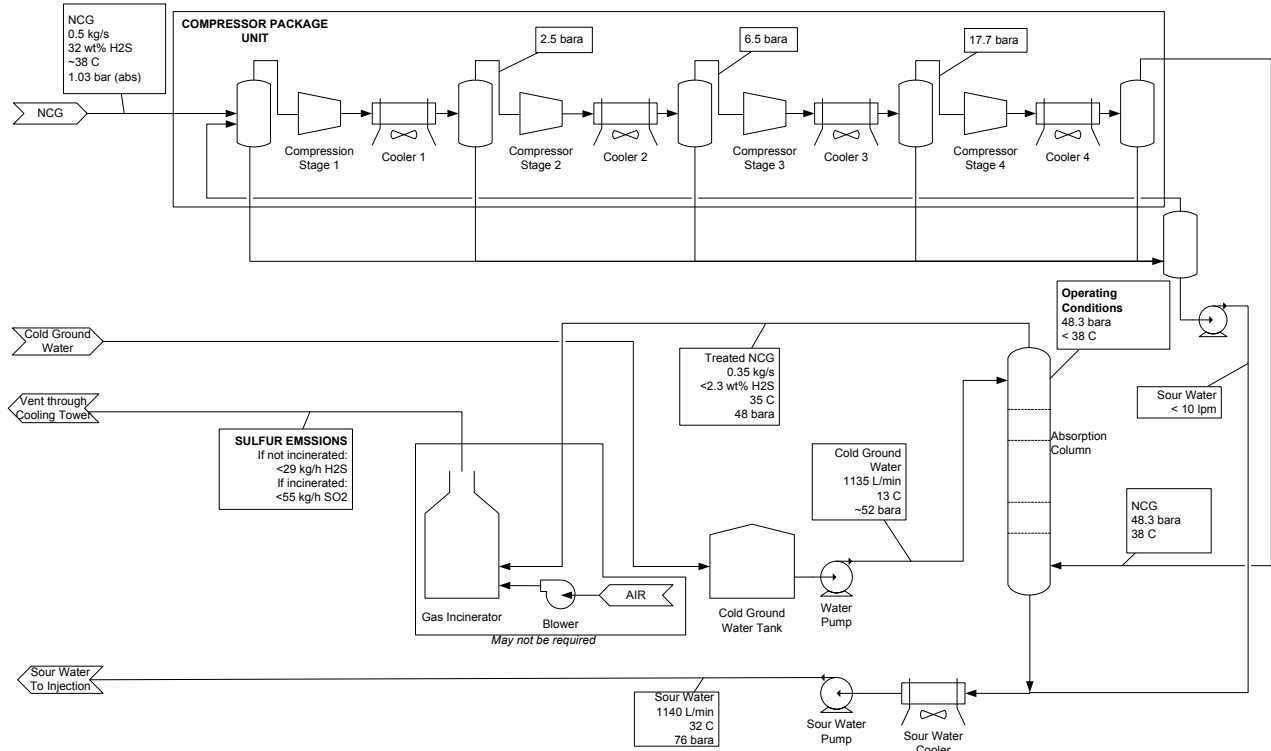
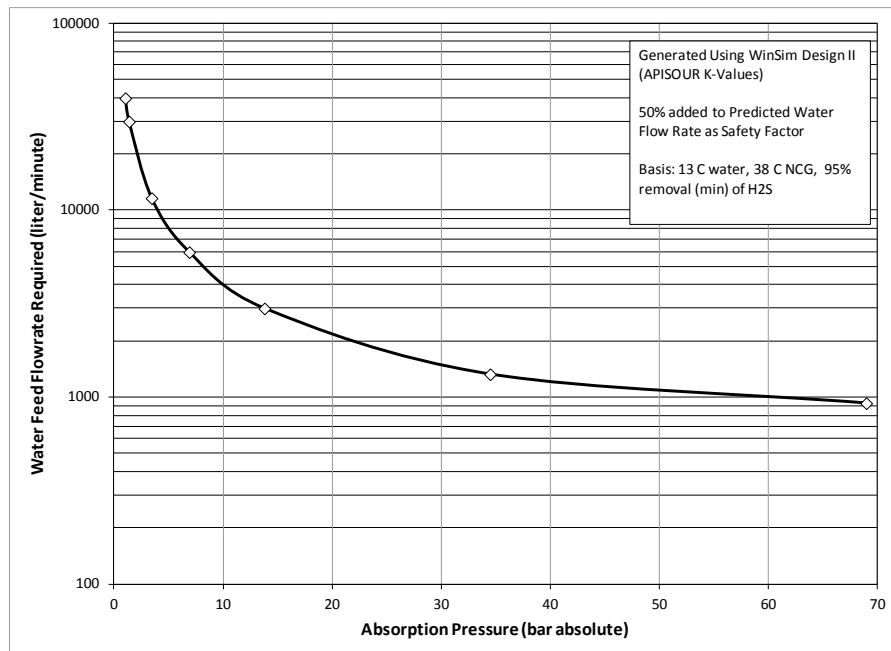


Figure 2. Estimate of Water Scrubbing of H₂S from NCG



Injection of acid gas (i.e., mixtures of H_2S & CO_2) into underground reservoirs is widely practiced in the oil and gas industry (McIntush et al., 2010), and it is known that at some installations the acid gas is dissolved into water and the resulting solution is injected. However, the use of water to absorb the acid gases from other gas components, followed by the injection of the sour water, is not widely practiced; further development work would be required to implement this concept. For this technology it is especially important to understand the characteristics of the injection well, injection zones, and the potential for H_2S migration. Total practical H_2S removal from the NCG is estimated to be limited to about 98% or less. The residual NCG stream would also still contain nearly all of the methane (which may be an environmental concern) and hydrogen that was in the feed NCG. In the case studied here, it was assumed that the treated gas is vented to the cooling tower without an incinerator.

The Claus Process (and other gas-phase oxidation processes)

The Claus process (Figure 3) is the standard technology used for handling large amounts of H_2S at natural gas processing plants and oil refineries. This process converts H_2S to elemental sulfur via simple, gas-phase chemistry. In this technology, the NCG stream would be sent to a special burner in which only enough air would be added to burn one third of the H_2S to SO_2 and to completely burn other combustibles (e.g., H_2 , CH_4). The SO_2 formed in the burner is then catalytically reacted with the remaining H_2S to form elemental sulfur. This chemistry is conducted in the vapor phase and/or via contact with a solid catalyst at elevated temperatures. The elemental sulfur is condensed to molten sulfur.

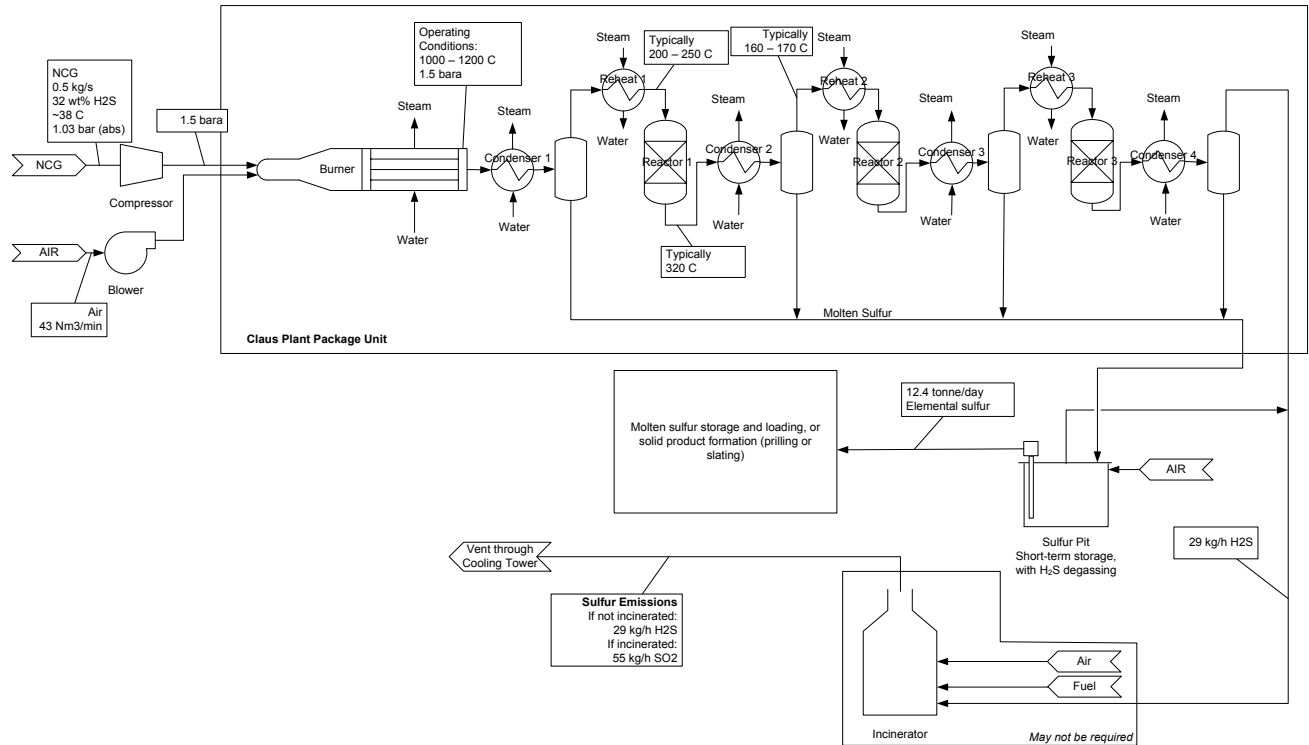
The Bjarnarflag application is on the small side (13 tonne/day sulfur) for a typical Claus application (usually 10-1000 tonne/day of sulfur). The NCG is also on the "lean side", meaning the H_2S concentration is relatively low (23 vol% H_2S vs. typical operation of significantly greater than 20 vol% and most often 70-90%). However, it is known that Claus processes have operated on gases with sulfur of only a few ton/day and also on similarly lean gas streams with significant contents of hydrogen. Although, to the authors' knowledge, the Claus process has not been used on geothermal NCG streams, a similar technology called Selectox has. Selectox has been used at a geothermal site in Japan, at a scale of 15 tonne per day sulfur and 6% H_2S in NCG (Takahashi et al., 2000), thus it seems that using the Claus process at Bjarnarflag would be reasonable.

The H_2S removal efficiency for this application would probably be limited to about 94-97% because of the high hydrogen content of the NCG. The hydrogen, while stabilizing the flame in the furnace, hinders the equilibrium conversion in the Claus unit. If 94-97% removal efficiency leaves too much H_2S in the tail gas from the Claus unit, then an incinerator may be required. For the purposes of this study, it was assumed that a tail gas incinerator would be used to increase the overall H_2S destruction efficiency.

The cost to import fuel for the tail gas incinerator to Iceland is considerable. If the Claus process were chosen, then Landsvirkjun could consider the use of electric heaters to replace the fuel or to route the Claus tail gas to the cooling tower instead. Some complications from routing to the cooling tower include: precipitation of solids, low pH water from absorption of residual H_2S and

SO₂ and the need for caustic to neutralize these species, and hazards of venting a gas stream with significant H₂S and SO₂ at a relatively low height above ground level.

Figure 3. Schematic of Claus Process Applied to NCG for H₂S Removal.



Liquid Redox Sulfur Recovery: Vanadium-Based Process

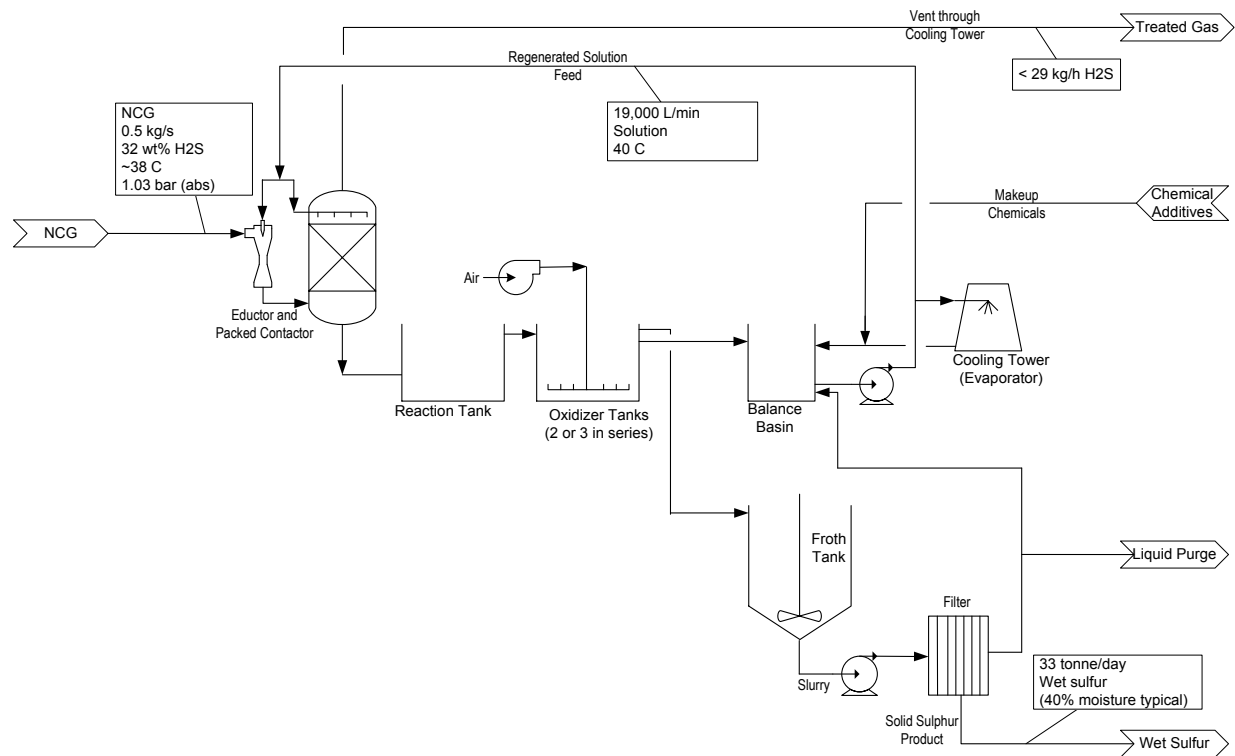
The Stretford process is a vanadium-based aqueous liquid redox process used for primary H₂S abatement in some other geothermal power plants. Stretford is most often used in conjunction with indirect contact condensers. Stretford is an older technology, and most plants were built before 1990. Iron redox technologies have displaced Stretford in most new applications because of concerns associated with vanadium content in the waste sulfur and solution. Figure 4 shows a schematic of the process.

H₂S in the NCG is removed by contacting the NCG with the Stretford solution in a venturi scrubber and packed tower. The Stretford solution passing through the Venturi scrubber imparts a slight pressure increase to the NCG; thus, the Stretford process can operate on low pressure gases with minimal pressure drop, usually without a blower. In a reaction tank, alkali vanadates convert the hydrosulfide ions (HS⁻) to elemental sulfur, and V⁺⁵ changes to V⁺⁴.

To regenerate the V⁺⁴ ions back to V⁺⁵, anthraquinone disulfonic acid (ADA) is used as an oxygen carrier / catalyst for the oxidation reaction; the V⁺⁴ oxidation via ADA/oxygen occurs in the oxidation tanks, which are sparged with air.

The oxidation tanks also serve to separate the sulfur from the solution by floatation. The sulfur froth that floats / forms on the surface of the liquid in the oxidizers overflows to the froth tank. Sulfur solids are separated from the slurry in the froth tank using different separation techniques (i.e., filter press, centrifuge, melter, etc.). Water used to wash sulfur cake in the filtration process is commonly sent to waste (“blowdown”). The Stretford process typically also contains a surge/pump tank, and an evaporator to remove water from the system (the reactions produce water). The main chemicals added to the Stretford system are vanadium (as sodium vanadate), ADA, and caustic; some other chemicals are known to be added at some plants.

Figure 4. Schematic of the Stretford Process Applied to NCG for H₂S Removal



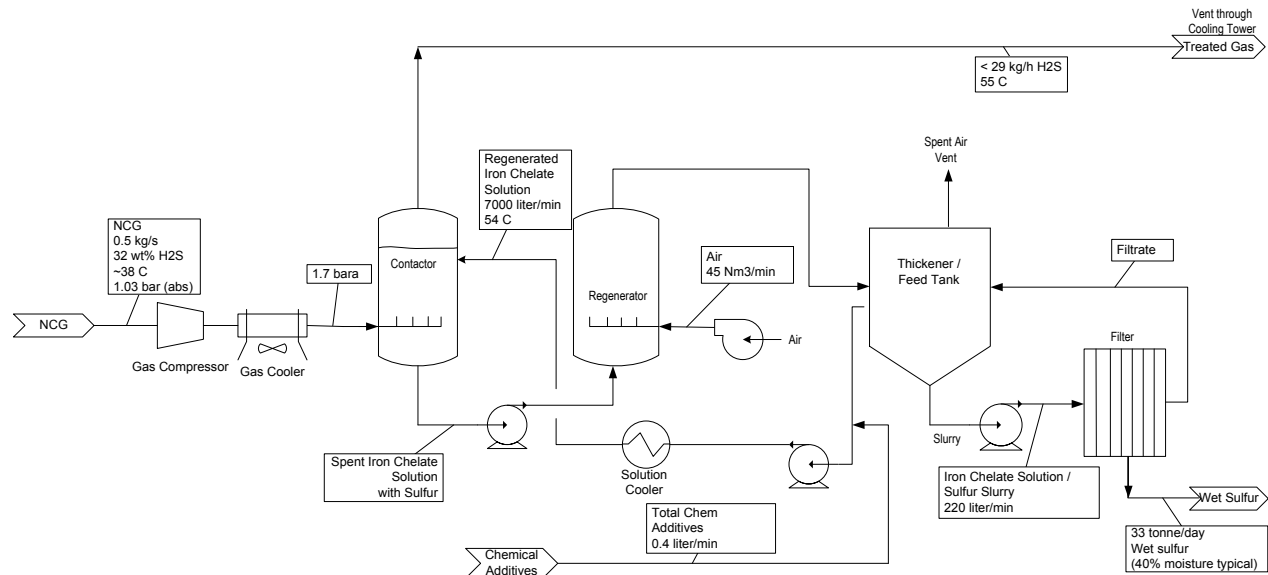
Liquid Redox Sulfur Recovery: Iron-Based Processes

The use of chelated iron to remove H₂S from gases via redox chemistry is well known. The chelant is required when using iron because the chelant allows substantial concentrations of soluble iron in the aqueous solution. The technology is widely practiced for gases that contain moderate volumes of H₂S. LO-CAT and SulFerox are the two most commonly licensed iron-based technologies although others exist. A schematic of a typical iron chelate H₂S removal process is shown in Figure 5 (the flow rates shown in the figure are based on SulFerox technology).

The H₂S reacts with the iron chelate in the gas contactor, quickly forming solid elemental sulfur particles. Very high removal efficiency of H₂S can be designed for, if desired. The sulfur particles are carried with the solution to the regenerator, where oxygen from air regenerates the iron chelate. Sulfur particle removal is usually accomplished by settling followed by filtration.

Several chemical additives are required, such as iron chelate to replace that lost from the system physically, excess chelant to replace chelant loss due to side reactions, sulfur conditioning additives (surfactants), caustic, and sometimes chelant degradation inhibitors and biocides.

Figure 5. Schematic of Iron Chelate Process Applied to NCG for H₂S Removal



Chelated iron and vanadium based H₂S removal processes share many of the same operating problems, which are most typically: solution foaming, plugging of vessels and pipe with sulfur, and higher than desired chemical makeup rates.

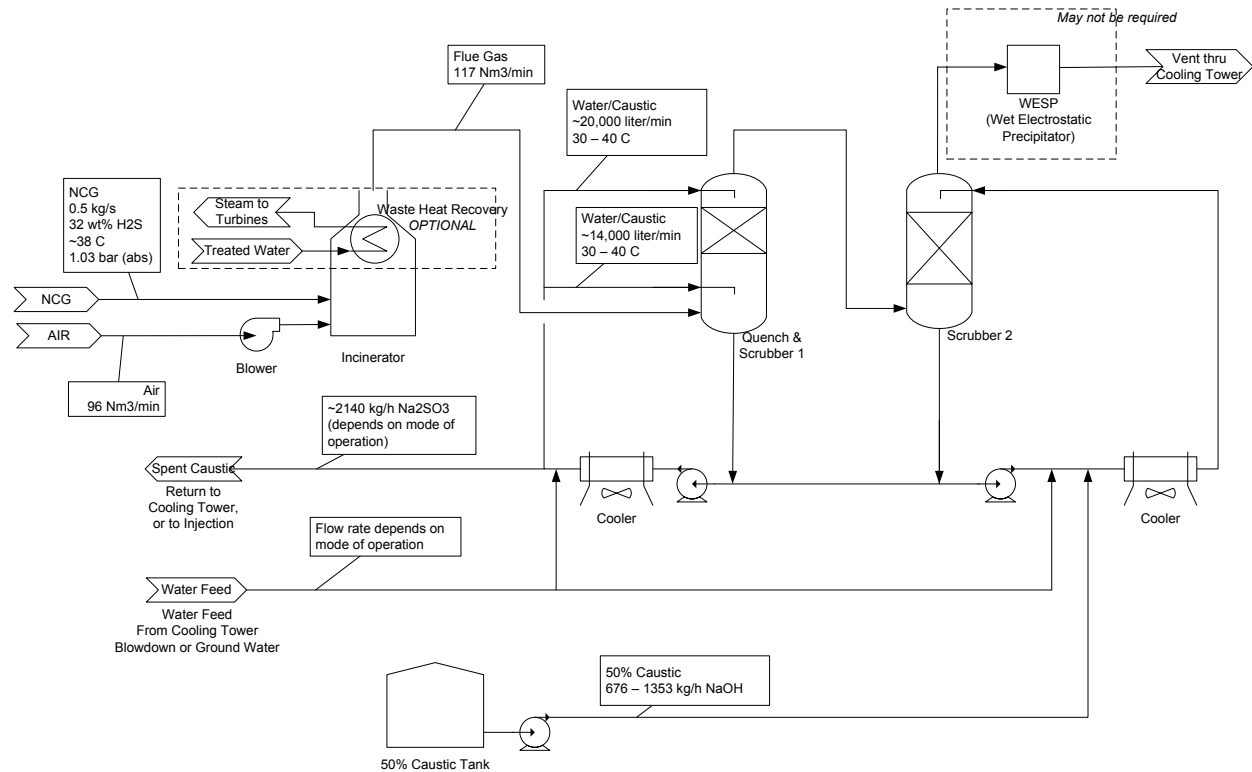
It should be noted that the treated gas will contain nearly all of the methane and hydrogen from the feed NCG. An incinerator could be used to convert any residual H₂S to SO₂ and burn the methane and hydrogen. However, the economic estimates assume that the clean NCG is simply vented via the cooling towers.

Burn/Scrub Process

A typical burn/scrub process that could be applied to the Bjarnarflag unit is shown in Figure 6, although many variations exist. The basic concept in the burn/scrub process is to incinerate the NCG stream, oxidizing the H₂S to SO₂ (a small amount will be oxidized to SO₃); the hydrogen and methane in the NCG will also be combusted. The incinerator is often provided with a heat recovery system in which water is converted to steam; for geothermal systems, the generated steam would be routed back to the power generators. The incinerated NCG is then routed through a quench vessel to cool the gas and scrubbers to remove the SO₂. The quench and scrubbers are sometimes combined. The scrubbing of the SO₂ is usually done using dilute caustic solution, forming sodium salts of sulfite. Wet electrostatic precipitators (WESP) are sometimes placed at the outlet of the system to capture sulfuric acid mist prior to the stream

being emitted to the atmosphere. For the Bjarnarflag application, the burn/scrub process could be integrated with the cooling tower system or operate standalone.

Figure 6. Schematic of Burn/Scrub Applied to H₂S Removal from NCG



Most geothermal burn/scrub plants operate the SO₂ scrubbers at a moderate (pH 6-8) pH and are integrated with the cooling tower system. Operation at this pH range results in the requirement of approximately two equivalents of caustic per equivalent of SO₂ to form Na₂SO₃ and water. Operating at roughly neutral pH allows for the use of water from the cooling tower, with some caustic added, to be used as the scrubbing solution. The scrubbing solution would be “once through”, meaning it would be sent through a scrubbing tower and returned directly to the cooling tower. This strategy allows for the removal of heat from the scrubbers by the cooling tower, and the use of the scrubbing solution in secondary H₂S abatement.

Because it is not planned to abate H₂S from the condensate at Bjarnarflag, it is also not necessary to integrate the SO₂ scrubbing system with the cooling tower. If designed as a standalone unit, the bulk of the scrubbing solution would be recirculated, with a portion removed as blowdown, and fresh caustic solution fed to the recirculating solution. Some equipment for the removal of heat from the system would also be required. As a result of this more equipment (coolers) are required to operate the scrubbers as standalone, but operation in the standalone mode provides an opportunity to minimize caustic usage: with the spent caustic solution from the SO₂ scrubbers not returned to the cooling towers, the scrubbers can be designed to operate at a lower pH.

Sulfur dioxide ($pK_{a1}=1.9$) can be absorbed into mildly acidic solution, and aqueous solution at pH as low as 4 may be capable of absorbing the majority of the SO_2 from the incinerated NCG. The reason for operating the scrubber at low pH would be to minimize caustic use. At pH 4, the main product of the absorption of SO_2 into caustic solution would be sodium bisulfate ($NaHSO_3$).

While at moderate pH the scrubbing of SO_2 requires about 1.5 to 2 equivalents of NaOH (depending upon pH and extent of side reactions), at low pH as little as one equivalent is required, thus resulting in lower theoretical NaOH requirements. Other side chemistry is possible, and will probably occur, which alters this theoretical requirement. An example of a side reaction that increases the NaOH requirement is the oxidation of sulfite to sulfate.

Operating in the standalone mode, due to the lower pH, the scrubber blowdown (spent caustic) would have to be disposed of separately (could not be added to the cooling tower). The only economic option for disposal of this stream would probably be injection. Therefore, the operation of SO_2 scrubbers as a standalone system would only be considered if Landsvirkjun could handle and inject the scrubber blowdown at a low pH (pH 4 – 5).

Although operation of the SO_2 in the standalone mode at low pH should be feasible, and should allow for lower caustic usage, it is unusual to operate an SO_2 scrubber in this mode. Any potential savings in caustic usage by design in this mode should be considered as preliminary and may require some development work to better characterize.

Another version of the burn/scrub process would be to burn the NCG and then scrub with cold ground water instead of caustic. This would need to be done at near atmospheric conditions because the increase in gas volume due to incineration makes compression of this stream uneconomic. At atmospheric conditions, the water required to scrub the SO_2 would be higher than the amount needed by the ‘water absorption and injection’ process that was previously discussed. The high water rate and larger processing equipment / power consumption would make this process less desirable than the water absorption of H_2S directly, so an economic evaluation was not performed.

Capital and Operating Costs

This section presents the basis for the capital and operating costs. The approximate total capital costs given here are on a 2010 basis. Although the Icelandic location would typically be associated with higher capital costs compared to many other locations, no special ‘location factor’ was included in this analysis. This is because the costs generated here are meant for the purpose of discriminating between technologies, and all of the technologies would have roughly proportional additional costs due to the Icelandic location and its climate extremes. A seven year, straight line capital depreciation was assumed (interest charges for capital were not included). Major operating costs include expenses such as chemicals, waste disposal, operating labor, and power; maintenance and overheads costs were not included. Chemical and utility unit costs were based on Landsvirkjun estimates, vendor estimates, or historic data; no adjustments were made for importation costs to Iceland.

Water Absorption

A conceptual design was developed for this process based on data from the simulation work. Major costs in the capital cost of the process include the NCG compressors and the absorption column. The material of construction for the majority of the equipment was assumed to be 316L stainless steel. The capital cost of the compression system was based on literature data (Carroll, 2010) while costs for all other major equipment were estimated using PDQ\$ cost estimating software. Costs were also included for supplying the cold ground water to the H₂S removal facility, for the pipeline to transfer the “sour water” from the H₂S removal facility to the injection well, and for the injection well (Carroll, 2010). It should be noted that the process design and cost estimate are conceptual and have the greatest amount of uncertainty of the technologies considered in this study.

Claus Process

The cost of the Claus process was based on an initial estimate by the authors (in-house) and a budgetary estimate from a Claus process vendor. The in-house capital and operating costs were estimated based on literature data (Tannehill, 2000), seminars given at the Gas Processors Association Convention, and data provided by an independent consultant. The estimate provided by the vendor for the Claus process did not include the full consideration of operating costs, so the in-house estimated operating cost was also applied to the vendor estimates to calculate the total treating cost. Likewise, there was some equipment (e.g., tankage) that was necessary for the overall process that the vendor did not include in the proposal so the authors added these to the overall capital cost.

It was also assumed that the sulfur produced in the Claus process would have a zero value. The zero value is based on an assumption that the storage and shipping costs would cancel out the sales value of the sulfur. Because of the need to incinerate the remaining NCG after this process, the cost of propane for the tail gas incinerator represents about 60% of the total operating costs. Alternatives to propane may exist (e.g., electric heat), but were not evaluated.

Liquid Redox Sulfur Recovery (LRSR)

This section summarizes the costs for the two LRSR processes: Vanadium-based (Stretford) and iron-based (LO-CAT and SulFerox). Budgetary cost estimates were obtained from vendors of the LO-CAT and SulFerox processes. The capital costs provided by the vendors were for the total cost of the equipment modules as shipped from their shops. Additional costs would be accrued for shipping, site preparation installation of the skids, and provision of the utilities that the processes require. A factor of 40% was applied to the vendor supplied skid cost to account for these costs. Additionally, both the SulFerox and LO-CAT designs called for delivering the NCG feed at a higher pressure than will probably be available; the cost of a compressor was therefore also added to the capital costs.

Capital costs for the Stretford unit were obtained by adjusting the economics for similar applications – both in-house and literature (Trofe et al., 1987) data – based on sulfur tonnage. A 0.6 exponent was used to scale the capital cost data. The Stretford process did not require the

compression of the feed gas, unlike the LO-CAT and SulFerox processes. The Stretford liquid blow-down (contains vanadium) was assumed to be injected down hole.

LRSR processes produce low-quality sulfur that is filtered from the aqueous operating solution. In the economic estimates in this study, a disposal cost of \$115/ton is assumed for LRSR sulfur. In the economic estimates, it was assumed that a tail gas incinerator is not required. Other operating costs for the LRSR processes were obtained from in-house and literature data (Hardison et al., 1990; Trofe et al., 1993).

Burn/Scrub

The capital cost for the burn/scrub option was obtained from individual equipment costs, and the installation factors shown in Table 1. The cost for the major pieces of equipment were obtained from costing software (PDQ\$), literature data (EPA, 1984; EPA, 2006), and scaled from vendor quotes for similar applications. A high capital cost estimate assumed that a waste heat boiler and WESP were installed and a low capital cost estimate did not include these items.

The operating costs are dominated by the cost of caustic (NaOH). A cost of \$818/tonne (dry) was used in this study, which is based on expected cost in Iceland; this is a rather high cost for caustic when compared with USA costs (typically roughly \$400/tonne). A high and low estimate for operating costs was also generated. The low value uses a 1:1 molar ratio of NaOH to SO₂ (standalone operation with injection of acidic spent caustic) and the high value uses a 2:1 ratio (integration with cooling tower system).

Economic and Key Characteristics Comparison

Table 2 shows a comparison of the capital and operating costs for the four main options evaluated. The capital cost estimates for the technologies vary substantially. Burn/scrub has the lowest capital cost, and Claus has the highest. Given the fact that the operating costs also vary greatly, the capital and operating costs must be considered together by the calculation of "Total Treating Cost," which is the total of the annual operating costs and the annual capital depreciation cost.

The Total Treating Cost clearly favors the use of the Water Absorption process, followed by the Claus process, and then LRSR. Given the high cost of caustic used in the study, burn/scrub would be dropped from consideration for this particular application.

Table 3 shows a comparison of some of the key characteristics of the different processes. Because of the unique location of the application, some of these parameters can have a big impact on the selection of the optimal technology.

Table 2. Summary of H₂S Removal Options

	Capital Cost (million \$)	Operating Cost* (million \$/y)	Total Treating Cost ** (million \$/y)	Comments on Technology
Water Absorption	10.95	0.37	1.94	Depends on ability to inject ~1140 lpm of low pH water with high H ₂ S vapor pressure. Technology needs to be developed.
Claus	19.2 – 21.9	0.72	3.47 – 3.85	~4600 ton/y pure sulfur product. Limited to ~94-97% H ₂ S removal without additional equipment. Fuel for incinerator (if installed).
LRSR	11.1 – 14.6	2.21 – 3.71	4.12 – 5.29	Produces ~7600 MT/y moist impure sulfur product (landfill). Must import a number of chemicals. High H ₂ S removal possible.
Burn and Scrub	4.55 – 6.20	4.89 – 9.54	5.54 – 10.16	Most economical if can be operated at low pH, and inject low pH water product. Have to import 11,000 – 22,000 MT/y of 50% caustic.

* Excludes maintenance and overhead

** Sum of Operating Cost and Capital Cost straight line depreciation over 7 years (no interest charges)

Table 3. Comparison of Key Characteristics for Different H₂S Removal Processes

Parameter	Water Absorption	Claus	LRSR	Burn and Scrub
Level of development / commercialization	No commercial applications. Development required.	Many Claus units in other industries, but not geothermal.	Well proven in geothermal; risks include occasional poor operability.	Well proven in geothermal; some process upsets, corrosion can occur
Technical risk	Some risk: technology not proven	Some risk: not proven in geothermal	Little technical risk	Little technical risk
H ₂ S abatement	95% practical, greater than 98% difficult	95% probable; higher possible	>95% easy; can go to 99.9% if required	>95% easy; can go to 99.9% if required
Turndown	Unlimited, with loss of efficiency	Turndown to 50% of design sulfur	Unlimited turndown	Moderate
CH ₄ abatement (not required)	Does not remove methane	Burns NCG	Does not remove methane	Burns NCG
Chemicals to be imported	None	Propane (~1.1 million liter/yr)	1000-2000 ton/year of total chemicals	5700 to 11,000 ton/yr of NaOH
Waste materials (or side products)	Sour water stream (~950 liter/minute) to inject	13 tonne/day of high purity sulfur	Dispose of wet sulfur, purge liquid	Na ₂ SO ₃ disposed of in cooling tower blowdown
Local laboratory requirements	Minimal local analysis required	Not likely needed	On-site laboratory for routine analysis	Minimal local analysis required
Estimate of plot space relative to cooling towers	1/3 the area of cooling towers	2/3 area of cooling towers because of need to store sulfur	2/3 area of cooling towers because of need to store sulfur	1/3 the area of cooling towers

Summary

This analysis shows that for this application in Iceland, the water absorption technology is potentially the most economical process. If the time and money can be invested in further developing this technology, the risk associated with implementing this somewhat unproven technology is accepted, and it is determined that the product sour water stream can be injected locally, then this technology would be favored for further development towards the goal of commercialization.

The Claus process is the technology that shows the second best economics. In addition to being an economically reasonable technology, it also does not produce a product that would be classified as a waste, and does not require the import of large volumes of chemicals. However, assuming that a tail gas incinerator is installed, it may be required to import a substantial amount of fuel (assumed to be propane) to operate the incinerator (alternately, it may be possible to design a tail gas incinerator that uses electric resistance heating, rather than propane). The need for tail gas incineration can also be reconsidered: injection of the tail gas into the cooling tower may also be possible. It appears that the Yanaizu-Nishiyama geothermal power station (Takahashi, 2000) may inject their Selectox process (similar to Claus) tail gas stream directly into the cooling tower. Although the similar Selectox process has been demonstrated on geothermal NCG, the Claus technology itself has not been demonstrated on geothermal NCG and is typically limited to 94 – 97% H₂S removal for a gas stream of this sort.

The LRSR technologies have somewhat higher total treating costs than the Claus process. Since the LRSR technologies have been proven in many geothermal NCG applications, the LRSR technologies are considered to be essentially free of unknown risks: the technology is known to work (although can be troublesome to operate) and the volumes of chemicals used and waste produced are not excessive.

The burn/scrub technologies are most favorable in areas of the world where bulk caustic solution is available at favorable prices and/or there is some ammonia in the steam; neither of these is the case at Bjarnarflag. Burn/scrub was not recommended for further consideration unless i) much lower costs for caustic than those assumed here are found, and ii) it is found that spent caustic from the SO₂ scrubber could be injected locally.

References

McIntush, K.E., C.A.M. Beitler, C. Wallace, D.T. Miller, D. Wentworth, P.F. Ellis, 13-17 September 2010, "Overview of the Design and Troubleshooting Efforts at Regency's Waha and Tilden Acid Gas Injection Facilities", Presented at the 2010 Vail Sulfur Recovery Symposium, Vail, CO.

Takahashi, Kazuo and M. Kuragaki, "Yanaizu-Nishiyama Geothermal Power Station H₂S Abatement System", May 28-June 10 2000, Proceedings World Geothermal Congress 2000, Kyushu – Tohoku Japan, page 719-724.

John J. Carroll, 2010, Acid Gas Injection and Carbon Dioxide Sequestration.

Mamrosh, McIntush, Beitler, Markússon, & Einarsson

Tannehill, C.C., 13-15 March 2000, "Budget Estimate Capital Cost Curves for Gas Conditioning and Processing", Proceedings of the 79th Annual GPA Meeting, Atlanta, Georgia, USA.

Trofe, T.W., D.A. Dalrymple, F.A. Scheffel, February 1987, "Stretford Process Status and R&D Needs", Topical Report, Gas Research Institute, Report No. GRI-87/0021.

Hardison, L.C., D.E. Ramshaw, 19-22 August 1990, "Economics of Redox Process Selection for Natural Gas H₂S Removal", AIChE 1990 Summer National Meeting, San Diego, California, p. 5.

Trofe, T.W., K.E. McIntush, M.C. Murff, September 1993, "Stretford Process Operations and Chemistry Report", Final Report, Gas Research Institute, Report No. GRI-93/0121.

EPA, January 1984, "Retrofit Cost Relationship for Hazardous Waste Incineration", EPA-600/2-84-008.

EPA, October 2006, "Air Pollution Technology Fact Sheet Wet Electrostatic Precipitator, Wire-Pipe Type", EPA -452/F-03-029.

McIntush, K.E., C. Beitler, M. Swadener, C. Wallace, "Screening Processes for Removal of H₂S from Enhanced Oil Recovery CO₂ Streams", 60th Laurance Reid Gas Conditioning Conference, 21 February to 24 February 2010, Norman, Oklahoma, USA, pp. 365-389.

Benn, B., K. McIntush, C.M. Beitler, D.L. Mamrosh, and O.E. Hileman, 24-27 October 2010, "Unit 14 / Sulphur Springs H₂S Abatement Process Screening and Stretford Improvements Study – Part 1", Geothermal Resources Council Transactions, v. 34, pp 1209 – 1212, 2010 Annual Meeting.

McIntush, K.E., N.A. Hatcher, J.A. Stavros, T.A. Andreatta, B. Christlieb, 24-27 February 2008, "Revised from 2007: Sharing Best Practices for Optimization of Beavon Stretford TGUs", 58th Laurance Reid Gas Conditioning Conference, Norman, Oklahoma, USA, pp. 429-460.

Mamrosh, D.L., C.M. Beitler, K.E. McIntush, K.S. Fisher, and S. Stem, 22-25 February 2009, "Use of Caustic in a Short Contact Time Approach to Selectively Scrub H₂S from CO₂-Contaminated Gas Streams", 59th Laurance Reid Gas Conditioning Conference, Norman, Oklahoma, USA, pp. 445-459.

Mamrosh, D., C. Beitler, K. Fisher, S. Stem, January 2008, "Consider improved scrubbing designs for acid gases", Hydrocarbon Processing, pp. 69-74.

Skinner, F.D., K.E. McIntush, M.C. Murff, December 1995, "Amine Based Gas Sweetening and Claus Sulfur Recovery Process Chemistry and Waste Stream Survey," Topical Report, GRI-95/0187, Gas Research Institute, Chicago, IL.

McIntush, K.E. and B.J. Petrinc, April 1995, "GRI Testing of SulFerox® for the Direct Treatment of High-Pressure Natural Gas at NGPL's Kermit, Texas Site," Final Report, GRI-94/0432, Gas Research Institute, Chicago, IL.