

MIT Open Access Articles

Techno-economic assessment of sour gas oxy-combustion water cycles for CO₂ capture

The MIT Faculty has made this article openly available. **Please share** how this access benefits you. Your story matters.

Citation: Chakroun, N. W., and A. F. Ghoniem. "Techno-Economic Assessment of Sour Gas Oxy-Combustion Water Cycles for CO₂ Capture." *International Journal of Greenhouse Gas Control* 36 (May 2015): 1–12.

As Published: <http://dx.doi.org/10.1016/j.ijggc.2015.02.004>

Publisher: Elsevier

Persistent URL: <http://hdl.handle.net/1721.1/105415>

Version: Author's final manuscript: final author's manuscript post peer review, without publisher's formatting or copy editing

Terms of use: Creative Commons Attribution-NonCommercial-NoDerivs License



Techno-Economic Assessment of Sour Gas Oxy-Combustion Water Cycles for CO₂ Capture

N.W. Chakroun*, A.F. Ghoniem

*Massachusetts Institute of Technology, Department of Mechanical Engineering
77 Massachusetts Ave., Cambridge, MA 02139, United States*

Abstract

Growing energy demand coupled with the threat of global warming call for investigating alternative and unconventional energy sources while reducing CO₂ emissions. One of these unconventional fuels is sour gas, which consists of methane, hydrogen sulfide and carbon dioxide. Using this fuel poses many challenges because of the toxic and corrosive nature of its combustion products. A promising technology for utilizing it is oxy-fuel combustion with carbon capture and storage, including the potential of enhanced oil recovery for added economic benefits. Although methane oxy-fuel cycles have been studied in the literature, using sour gas as the fuel has not been investigated or considered. In this paper, water is used as the diluent to control the flame temperature in the combustion process, and the associated cycle type is modeled to examine its performance. As the working fluid condenses, sulfuric acid forms which causes corrosion. Therefore, either expensive acid resistant materials should be used, or a redesign of the cycle is required. These different options are explored. A cost analysis of the proposed systems is also conducted to provide preliminary estimates for the levelized cost of electricity (LCOE). The results show the acid resistance cycle with a 4.5% points increase in net efficiency over the cycle with SO_x removal. However there is nearly a 9% decrease in the cycle's LCOE for the latter case.

Keywords: Oxy-fuel combustion, Power cycle analysis, CO₂ capture and sequestration, Sour gas

1. Introduction

Greenhouse gas (GHG) emissions are the leading contributors to climate change. Combustion of fossil fuels results in the formation of significant amounts of carbon dioxide (CO₂), the primary GHG released into the atmosphere. At the global scale, CO₂ emissions accounted for approximately 77% of all GHG emissions in 2007 [1]. According to the Environmental Protection Agency (EPA), about 26% of all global GHG emissions were produced by the electricity generation sector. It is also estimated that the world CO₂ emissions from electricity production will increase by approximately 43% by 2035, from 30.2 billion metric tons in 2008 to

*Corresponding author
Email address: nwc@mit.edu, Telephone number: +1 (765) 409-7402

8 43.2 billion metric tons in 2035 [2]. Much of this growth in emissions is attributed to the developing non-
9 OECD countries which continue to rely on fossil fuels to meet their growing energy demand. By 2040, these
10 non-OECD countries are expected to contribute as much as 69% of the world's total emissions, whereas the
11 OECD emissions, totaling about 14 billion metric tons, represent the balance [2]. Emissions reductions are
12 vital for the world and the developing nations with their ever increasing populations and energy demands.

13 The International Energy Agency (IEA) has identified carbon capture and storage (CCS) as one of the
14 important strategies in reducing CO₂ emissions [3]. In this technology, CO₂ released from power plants
15 is separated, compressed and transported to a site for underground injection in secure geological forma-
16 tions, including natural underground reservoirs, or depleted oil and gas fields. The integration of these
17 CCS technologies with the power generation plants has not yet been fully demonstrated commercially at a
18 large enough scale that can overcome the technological risk and cost barriers [4, 5]. Nonetheless, oxy-fuel
19 combustion is one of the promising CCS options [6], the other ones being post-combustion CO₂ capture and
20 pre-combustion CO₂ capture [7]. The main difference between these technologies is the location at which
21 the CO₂ is removed in the cycle. In oxy-fuel combustion, the fuel is burned in oxygen diluted with CO₂ or
22 water, at near stoichiometric conditions so that the products consist of only carbon dioxide and water. The
23 water can then be easily separated from the carbon dioxide by condensation. A diluent is added to the fuel
24 and oxidizer to moderate the temperatures in the combustion chamber. **Due to the simplicity of the carbon
25 capture system in oxy-fuel combustion, the CO₂ capture efficiency is very high (90% +) [7]. This is one of
26 the main reasons for the recent interest in this CCS technology.**

27 Oxy-combustion has often been associated with coal since coal power plants produce about two times
28 as much CO₂ per MWh than natural gas power plants [1]. On the other hand, natural gas' share of the
29 world's electricity generation is expected to grow from 22% in 2010 to 24% in 2040 [2] and applying the same
30 concept to this fuel has been suggested. Estimated cost of electricities by source [8], suggest that natural gas
31 cycles for CCS are competitive with other zero carbon energy sources. For our analysis we will be focusing
32 on an even cheaper source of natural gas, namely sour gas.

33 Sour gas consists of three major components: methane (CH₄), hydrogen sulfide (H₂S) and CO₂. This is
34 the form of natural gas extracted, from a growing number of gas fields, prior to the purification process [9].
35 Typical volume fractions of the H₂S and CO₂ compounds are between 0-30% each, the exact composition
36 changes depending on the life of the well, location and geography [10, 11]. Nearly 40% of the world's gas
37 reserves can be classified as being sour [11]. Currently, for conventional natural gas power plants, expensive
38 and energy-intensive purification processes are done to remove H₂S and CO₂ before the methane is burned
39 for power generation. At high concentrations of both, it is not practical or economical to extract the gas.
40 As a result of this, a large fraction of world wide natural gas resources are currently unusable [10].

41 The objective of this work is to explore the use of sour gas directly as the fuel in an oxy-combustion power

42 plant for CCS. The utilization of this unusual fuel directly saves on the energy utilized for the purification
43 process [9]. Furthermore, enhanced oil recovery (EOR) using CO₂ injection can increase the life of the
44 reservoir by about 5-15% [12]. Currently in the literature, there is a lack of research on the utilization of
45 sour gas as the fuel directly in power plants, and especially in oxy-combustion cycles. The issue of corrosion
46 is the main hindrance to the progress and interest of using this gas directly. Once this knowledge gap has
47 been addressed and tackled, these thermodynamic performance studies can then be used as inputs for future
48 work focusing on the combustion behavior (ex. flame dynamics, reaction zone structures, stability) in these
49 sour gas combustors, similar to what has been done for methane oxy-combustion [6, 13, 14].

50 Sour gas combustion produces SO_x and H₂SO₄ which can cause corrosion and also affect the trans-
51 portation and storage of the CO₂ stream for EOR. Therefore an important part of the design is limiting
52 the concentrations of these compounds in the products. Since sour gas technologies, specifically for oxy-
53 combustion, has not received much attention, it is necessary to investigate different options for using this
54 fuel to determine their viability and evaluate their potentials.

55 Oxy-fuel cycles have the great advantages of almost eliminating NO_x emissions, and also providing a
56 simpler mechanism to capture CO₂ [14, 15]. The flame temperature in pure oxygen is very high and so
57 a diluent is needed. The diluent used is usually some form of a recycled flue gases. For methane oxy-
58 fuel cycles, several configurations have been studied in the literature. Semi-Closed Oxy-fuel Combustion
59 Combined Cycles (SCOC-CC) [16, 17, 18, 19, 20] recycle part of the CO₂. In Water cycles [19, 21, 22, 23],
60 H₂O is separated from the flue gases and recycled back to the combustor. The Graz cycle [16, 19, 24] adopts
61 both CO₂ and H₂O recirculations.

62 The focus of this paper is on sour gas water cycles [21] with some modifications due to the presence of
63 the sulfur compounds. As mentioned, this type of analysis is completely novel and has not been addressed
64 in the literature before for sour gas. The water cycle can be categorized as a Rankine-type cycle with reheat
65 and regeneration. Methane based water cycle working fluid consists mainly of H₂O and CO₂ (90/10 %vol.),
66 whereas the sour gas case (as will be shown later), also has SO_x compounds in the working fluid which will
67 affect the heat capacity of that stream and thus the power output in the turbines and also the performance
68 of the whole cycle. These SO_x compounds affect the dew point temperature of the working fluid which can
69 then cause acids to form and condense leading to corrosion issues in the low temperature components (ex.
70 condenser, regenerator). Therefore, the sour gas cycles need to be modified.

71 There are four main areas and stages where the sulfur compounds pose problems in the cycle: fuel
72 compression, expansion in the turbines, low temperature and condensation equipment, and in the CO₂
73 purification unit (CPU) for EOR. Since the fuel contains high levels of H₂S, corrosion is an extremely serious
74 issue for the fuel compressors. Thomas et al. [25] suggested the use of the corrosion resistant alloy 20Cb-3
75 (Carpenter Alloy[®]), and the material's data sheet also recommends this alloy as having good corrosion

76 resistance to sour gas [26]. This material was used in the cost of electricity calculation as will be explained
77 later on. After combustion, SO_x compounds are formed which can cause hot corrosion in the turbines (as
78 will be explained in section 4.8), and the corrosive agent's deposition rate was found to be independent of the
79 sulfur content [27]. Once again certain materials will have to be used which will further increase costs. The
80 condensation of the working fluid, containing SO_x , leads to the formation of sulfuric acid which corrodes the
81 components unless an acid-resistant material is used. The dew point temperature of the working fluid affects
82 when this condensation occurs and so this is a critical design feature of these sour gas cycles. Finally, these
83 sulfur compounds need to be removed from the system either through condensation and liquid separation
84 or by a SO_x removal system (described in section 3.3.1). This is done in order to meet CO_2 transport and
85 EOR constraints (ex. $\text{SO}_2 < 100$ ppm) and so these sulfur compounds need to be removed.

86 This paper is organized as follows: the methodology used in this analysis is briefly described in Section
87 2. In Section 3 the different sour gas water cycle configurations are described. In Section 4, the results of the
88 analysis and cycle simulations are presented and discussed. Finally, Section 5 wraps up with the summary
89 and conclusions of this work.

90 2. Methodology

91 Aspen Plus[®][28] was used in the modeling analysis of the sour gas cycles. The PR-BM (Peng Robinson
92 cubic equation of state with Boston-Mathias alpha function) property method [28] was used to model the
93 combustors. When modeling other components, a different property method had to be used because of the
94 unusual components in the working fluid: the sulfur compounds. The SR-Polar (Schwarzentruber and Renon
95 equation-of-state model) property method was chosen to be the best fit for our application because it can be
96 applied to highly polar components, e.g. SO_2 and SO_3 , and recommended for high temperature and pressure
97 applications.

98 Following an extensive literature review on methane oxy-fuel power cycles (since sour gas has never been
99 addressed), two different configurations were considered for the sour gas water cycles: an acid resistance
100 and a SO_x removal cycle. The Acid Resistance cycle is where we allow the working fluid, containing sulfur
101 compounds, to condense. In this case, sulfuric acids form, which can corrode the components. Therefore,
102 we assume that we use acid-resistant materials in order to protect the components where the acid is present.
103 Acid resistant materials represent a major economic burden and can significantly increase costs, as will
104 be shown later. The second type is the SO_x removal configuration, applied just before the working fluid
105 condenses in the main cycle. All sulfur compounds are removed in this system and the exiting gas stream
106 contains only mainly CO_2 with some Ar, N_2 (since the oxidizer is 95% O_2) and H_2O (similar to the pure
107 methane cycles). This purified stream is used in the rest of the cycle. This configuration solves the problem
108 of acid condensation, but as will be shown later there is an efficiency penalty associated with this SO_x

109 removal process.

110 These two cycles were simulated and a comparison was made with respect to the overall net efficiency,
111 working fluid compositions and cost of electricity estimates. The modeling details and the results of this
112 analysis are shown and discussed in the following sections.

113 3. Sour Gas Water Cycles

114 3.1. Modeling Assumptions

115 The important assumptions made when performing the thermodynamic modeling and simulations of the
116 two sour gas water cycles are shown in Table 1. The same assumptions were applied to the Acid-Resistance
117 and SO_x Removal cycles.

118 When modeling the combustor, the “RGibbs reactor” model was used [28]. RGibbs models single-phase
119 chemical equilibrium, or simultaneous phase and chemical equilibria. The reaction kinetics are not taken into
120 account. A Gibbs free energy minimization is done to determine the product composition. It is commonly
121 used in the literature to model combustors when reactions occurring are not known, or are high in number
122 due to the many components participating in the reactions. The combustion process was also assumed to
123 be stoichiometric.

124 Because the combustors were modeled as equilibrium type combustors, this will grossly under-predict the
125 SO₃ concentrations at the exit. Since the amount of SO₃ dictates how much acid forms in the latter stages
126 of the cycle, it is important to try and improve the accuracy of the concentration of SO₃ in the working fluid.
127 To do this, an additional reactor was added to model the SO₃ formation after the equilibrium combustor
128 reactor (not shown in the cycle diagrams). The second reactor was modeled as an “RStoic” reactor in Aspen
129 Plus [28]. This reactor was used only to model the formation of SO₃ from SO₂ using the single reaction:
130 $\text{SO}_2 + \frac{1}{2} \text{O}_2 \rightarrow \text{SO}_3$ with a specified conversion rate (SO₂/SO₃) of 1.5% obtained from [29]. The conditions
131 that they tested were significantly different than ours (250-1000 ppm vs 11% SO₂), and so the conversion
132 percentage chosen is really an upper limit and represents the most conservative estimate since the conversion
133 ratio was found to decrease with increasing SO₂ concentration.

134 However in our cycle since the combustors are stoichiometric, there is not enough excess oxygen for
135 the SO₂ to react with to achieve that conversion percentage. Thus the SO₃ concentration only increases by
136 about 10-30 ppm (by volume) across that second reactor. In reality however, this SO₃ concentration actually
137 decreases during expansion of the gas in the turbine [30] making our prediction even more conservative,
138 since the mixture is essentially frozen during the expansions process. Also hot corrosion was found to be
139 independent of the sulfur content in the working fluid [27] and so our SO_x concentrations estimations will
140 not affect material selection.

	Sour Gas Water Cycles
Fuel	
Composition (mol%)	70% CH ₄ , 15% H ₂ S, 15% CO ₂
Combustors	
Operating Pressures (bar)	100 & 15 (Main & Reheater)
Pressure Drops (%)	10 & 6 (Main & Reheater)
Turbines	
TIT's (°C)	600 & 1200 (HPT & LPT)
Isentropic Efficiencies (%)	87 & 90 (HPT & LPT)
Pumps	
Isentropic Efficiency (%)	75
Heat Exchangers	
Minimum Internal Temperature Approach (°C)	20
Pressure Drops (%)	5
ASU	
Specific Power (kWh/kg-O ₂)	0.225
O ₂ Stream Composition (mol%)	95% O ₂ , 4.2% Ar, 0.8% N ₂
SO_x Removal System	
Gas Exit SO ₂ Concentration	< 100 ppm
Liquid Exit pH	≈ 7
CPU	
CO ₂ Delivery Pressure (bar)	110
Exit CO ₂ Stream Composition (mol%)	> 99% CO ₂ (EOR Ready)

Table 1: Sour gas water cycles modeling assumptions

141 For the low pressure turbine (LPT), we have chosen the reheater temperature to be 1200°C¹. It was
142 found that increasing the LPT inlet temperature had a bigger impact on efficiency than increasing the high
143 pressure turbine (HPT) inlet temperature. A fixed combustor exit temperature of 600°C was chosen, and
144 was controlled by the certain proportion of working fluid (water) that was recycled back to the combustor
145 (stream 1 in Fig. 1). Turbine blade cooling for the LPT was not considered in this study. However, in reality,
146 these high temperature turbines will definitely require cooling to be able to handle these temperatures. The
147 abundance of water and cool steam (around 200°C) streams, could be used as the cooling fluids for these
148 high temperature systems. But nonetheless, the main conclusions from this study are not expected to be
149 significantly impacted.

150 The operating pressure of the combustor for the water cycle is 100 bar. Combustor pressure sensitivity
151 did not impact the cycle efficiency or SO_x concentrations significantly enough to cause for changing this
152 operating pressure away from what is commonly used in the literature [23, 32]. The reheater operates at
153 15 bar, which was found using a pressure sensitivity analysis (discussed later) to determine the optimum
154 reheater pressure. The pressure drops for the two burners were taken as 10% and 6% for the combustor²
155 and reheater respectively.

156 One key aspect of oxy-fuel combustion is the oxygen production process. Air separation units (ASU)
157 using cryogenic separation is the only available option to produce the large amounts of oxygen required by
158 these plants [34]. Cryogenic ASUs have significant energy penalties equivalent to 7-10% efficiency points.

159 The air separation unit model is similar to that of Hong et al. [35]. This ASU model produces an oxygen
160 stream with an outlet oxygen purity (by volume) of 95% O₂, 4.2% Ar, 0.8% N₂ at a pressure of 1.24 bars,
161 while requiring a specific power of 0.225 kWh/kg-O₂ (0.812 MJ/kg-O₂). This value is close to what is used
162 in the literature [19].

163 The excess working fluid from the two cycles ('VAP' stream in Figures 1 and 5), is sent to a CO₂
164 purification unit (CPU) where the non-condensable gases (Ar & N₂) are removed and the capture-ready
165 CO₂ stream is compressed up to 110 bar.

166 The incoming stream (mostly CO₂) to the CPU has the inert gases removed using low temperature sep-
167 aration techniques and the purified CO₂ stream is extracted as a liquid and pumped up to the sequestration
168 pressure, and an exhaust stream consisting of mainly the inert gases is also produced. The separation process
169 was modeled based on the layout of gas removal configuration B in [36]. This process was chosen because
170 it delivers a liquid stream, thus eliminating the cost and energy penalty of gas phase compression of the
171 purified stream. The separation technique also requires external refrigeration to provide the cooling load to

¹As a reference, Clean Energy Systems (CES) has implemented an oxy-fuel water cycle for turbines with turbine inlet temperatures (TIT's) of 1080-1260°C [22, 31].

²As a reference, CES's gas generator is rated with a pressure drop of 10-15% [33]

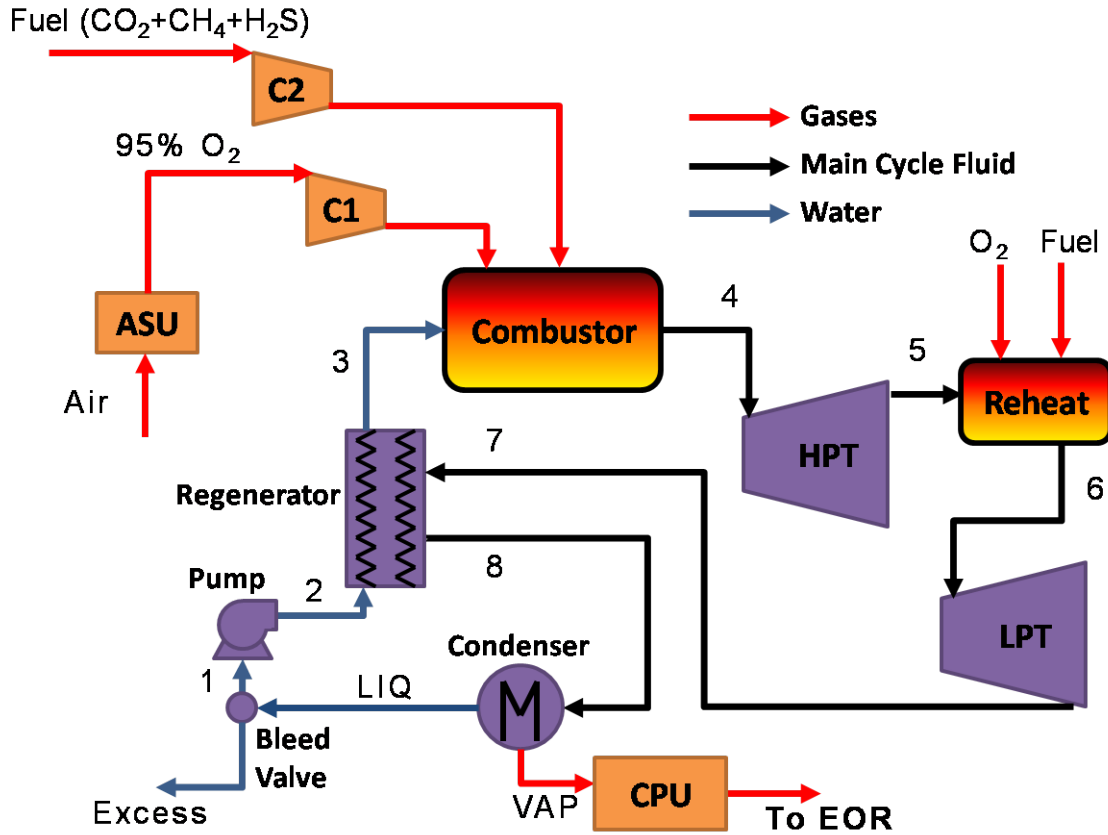


Figure 1: Overall process layout for the sour gas water cycle with acid resistance

172 the unit.

173 3.2. Acid Resistance Cycle

174 The “Acid Resistance Cycle” configuration is similar to the water cycle described in [19, 21, 22]. A
 175 similar component layout is adopted here with the difference being the fuel (70% CH₄, 15% H₂S, 15% CO₂)
 176 and the fact that acid resistant materials are used for all the cycle components where condensation occurs.
 177 Figure 1 shows the cycle diagram and components, with the corresponding T-s diagram in Figure 2. It is
 178 based on a Rankine cycle with reheat and regeneration.

179 Water at state 1 is pumped to 100 bar where it is preheated in the regenerator to about 247°C before
 180 entering the combustor. On the gas side, the oxygen stream from the air separation unit is sent to the
 181 combustor along with the fuel (70% CH₄, 15% H₂S, 15% CO₂), and the recycled working fluid (water).
 182 Water acts as a diluent in the combustor, and so the recycle ratio (\dot{m}_1/\dot{m}_{LIQ}) of the working fluid fixes
 183 the combustor exit temperature to 600°C. The main combustor flue gases (5% CO₂, 93% H₂O, 1% SO₂
 184 by volume), state 4, are expanded in the high pressure turbine (HPT) to 15 bars to produce power. The

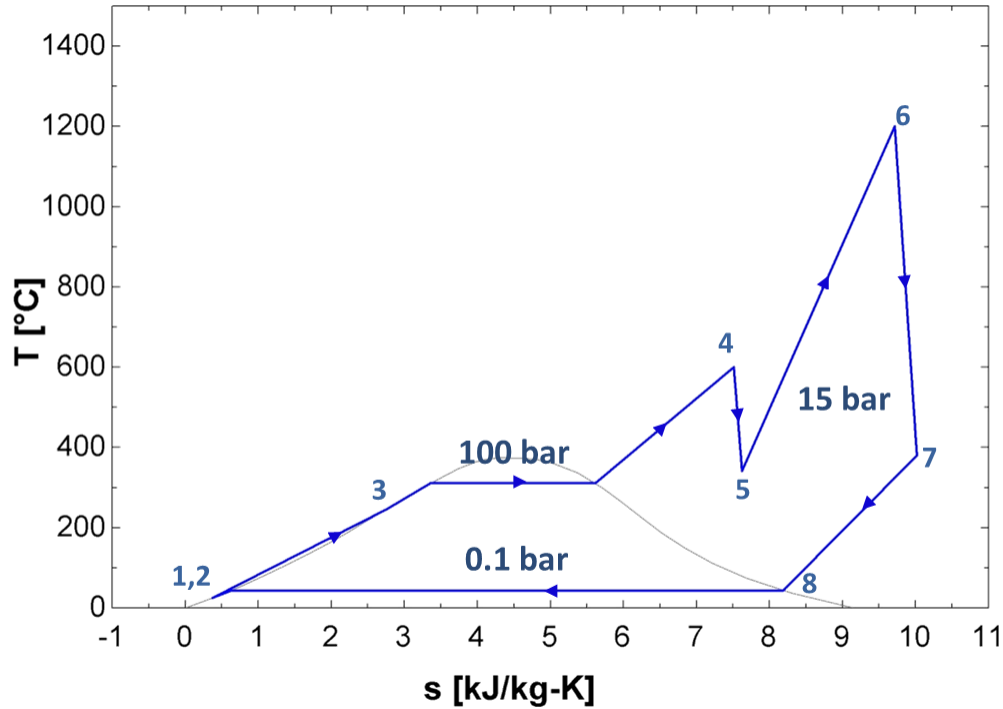


Figure 2: T-s diagram of the sour gas water cycle with acid resistance

185 working fluid at state 5 is reheated in the reheater where more fuel and oxygen are combusted to achieve a
 186 temperature of 1200°C. The reheater flue gases (10% CO₂, 88% H₂O, 2% SO₂, <1%Ar by volume), state 6,
 187 are expanded in the low pressure turbine (LPT) down to 0.1 bar.

188 Next, the hot working fluid enters the regenerator where it transfers its thermal energy to the water
 189 stream going to the combustor while being cooled down to state 8. The regenerator was divided up into
 190 two parts: a non-condensing heat exchanger and condensing heat exchanger. This was done to minimize
 191 the cost of acid resistance material needed in the regenerator; standard materials could be used for the
 192 non-condensing part and the expensive acid-resistant materials would only be required for the condensing
 193 section.

194 The remaining working fluid at state 8 is condensed to 25°C in the condenser and the vapor is separated
 195 out to be sent for EOR. Since the working fluid, containing sulfur compounds, is allowed to condense in
 196 the regenerator and condenser, sulfuric acid forms in those components, requiring acid resistant materials.
 197 After the condenser, 87% of the remaining liquid (water) is recycled back to the pump to be used as the
 198 dilution medium in the combustor. The vapor from the condenser is sent to the CPU and compressed to
 199 110 bars. The CPU removes the inert compounds (Ar & N₂) but before this is done, the sulfur compounds
 200 are also removed. This SO_x removal system is described in detail in section 3.3.1. For these systems, no

201 extra water is needed to make up for the water that leaves with the vapor stream in the condenser and the
202 'Excess' stream in the bleed valve. At steady state, all the water formed in the combustor and reheater due
203 to combustion, leave in the 'VAP' and 'Excess' streams.

204 The efficiency of this cycle with these conditions was found to be 40.9%. This is almost 0.5% points lower
205 than the methane water cycle, which has the same layout and operating conditions but different fuel. The
206 slight difference in efficiency can be attributed to the fact that the methane cycle has a working fluid with
207 a slightly higher heat capacity (because of the higher CO₂ fraction) and as such produces more work in the
208 turbines, increasing the efficiency.

209 A pressure sensitivity analysis was performed to investigate the effect of reheat pressure on the important
210 cycle parameters. This analysis was performed on the sour gas and methane water cycles by varying the
211 reheater pressure between 6-30 bars, and the results are shown in Figures 3 and 4. The reheater pressure was
212 varied, instead of the combustor, because it was found to have a higher impact on cycle efficiency and SO_x
213 concentrations. Because of the higher pressure ratio across the LPT and the higher TIT, a larger proportion
214 of the power output came from the LPT.

215 Figure 3 shows the effect of varying pressure on the net cycle efficiencies. The efficiencies of both cycles
216 increase with the pressure until a maximum is reached at about 15 bar. This is mainly because when the
217 reheat pressure is changing, the fuel (and oxidizer) flow rates are continuously adjusted in order to maintain
218 a 1200°C reheat exit temperature. This affects the total heat input to the cycle which in turn affects the
219 efficiency. However, the pressure sensitivity analysis revealed that the efficiency did not vary by more than
220 0.5% when changing the pressure. The methane cycle also has about a 0.5% efficiency gain over the sour gas
221 cycle, this is because the methane cycle's working fluid has a slightly larger heat capacity which produces
222 more work and increases efficiency.

223 SO₂ and SO₃ concentrations (at the exit of the reheater) versus the pressure are shown in Figure 4.
224 As can be seen, the reheater pressure had very little effect on both only causing a slight drop in the SO₃
225 fraction which are expressed in parts per million (ppm). This means that the system design in regards to
226 acid formation and condensation will not be impacted, and if this cycle is implemented, changes in that
227 reheater pressure during operation will not be a major concern.

228 3.3. SO_x Removal Cycle

229 This cycle was modeled in order to determine how the impact of removing the SO_x compounds from the
230 working fluid affects the performance and cost. Figure 5 shows the cycle diagram and components, with the
231 corresponding T-s diagram in Figure 6.

232 Water at state 1 is pumped to 100 bar, then preheated in the regenerator to about 260°C before entering
233 the combustor. Oxygen from the ASU is sent to the combustor along with the fuel (70% CH₄, 15% H₂S,
234 15% CO₂), and the recycled working fluid (water). The recycle ratio (\dot{m}_1/\dot{m}_{LIQ}) fixes the combustor exit

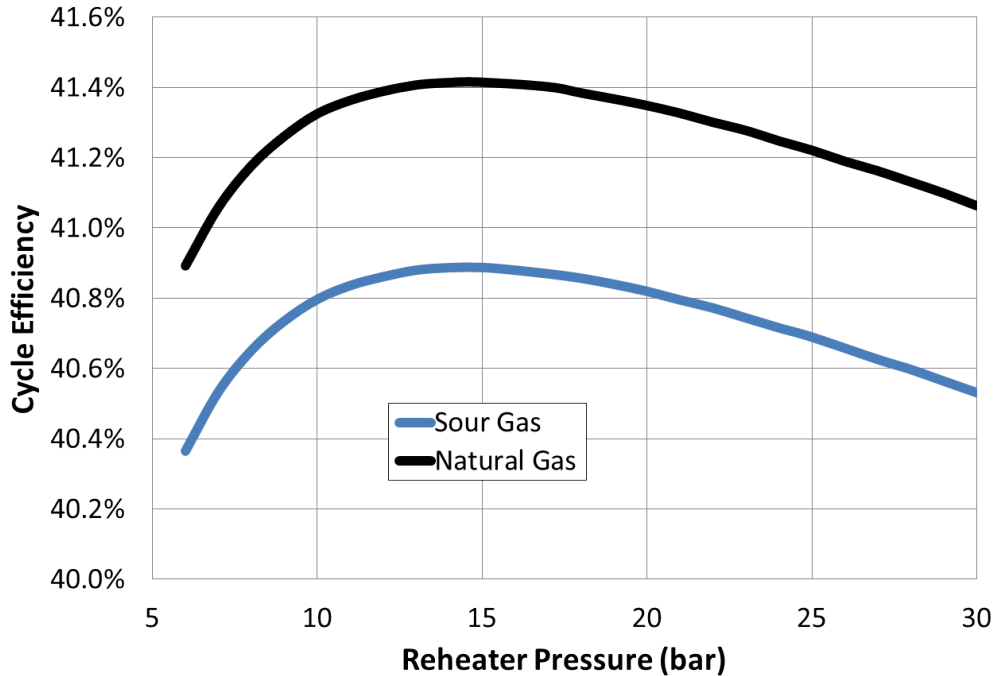


Figure 3: Effect of varying reheater pressure on the net cycle efficiency for the sour gas (acid resistance) and methane water cycles

235 temperature at 600°C. The combustor gases (5% CO₂, 93% H₂O, 1% SO₂ by volume), state 4, are expanded
 236 in the HPT to 15 bars. The fluid is then reheated where more fuel and oxygen are combusted to reach
 237 1200°C. The exit stream (10% CO₂, 88% H₂O, 2% SO₂, <1%Ar by volume), state 6, are expanded in the
 238 low pressure turbine (LPT) down to 0.28 bar.

239 The low pressure fluid enters the regenerator where it transfers its thermal energy to the water stream
 240 going to the combustor while being cooled down to state 8. In this case, the working fluid is not allowed to
 241 condense in the regenerator by limiting the exit temperature to values higher than the dew point. The dew
 242 point of this cycle's working fluid (10% CO₂, 88% H₂O, 2% SO₂, <1%Ar by volume) was close to 203°C,
 243 and thus the hot stream exit temperature was fixed at 208°C while the cold stream's exit temperature was
 244 calculated such that the minimum internal temperature approach inside the heat exchanger was 20°C. At
 245 the exit of the regenerator the hot stream is sent to the SO_x removal system, leaving CO₂ with some Ar
 246 and N₂. The SO_x removal system is similar to the traditional flue gas desulfurization systems found in coal
 247 power plants where the flue gases are sprayed with a mixture of lime (CaO) and water which condenses
 248 and neutralizes the acidic mixture. SO₂ dissolves in the liquid and is separated from the gas stream. This
 249 process will be explained in greater detail later.

250 At the exit of the SO_x removal system, 82% of the liquid water is recycled back to the pump to act as

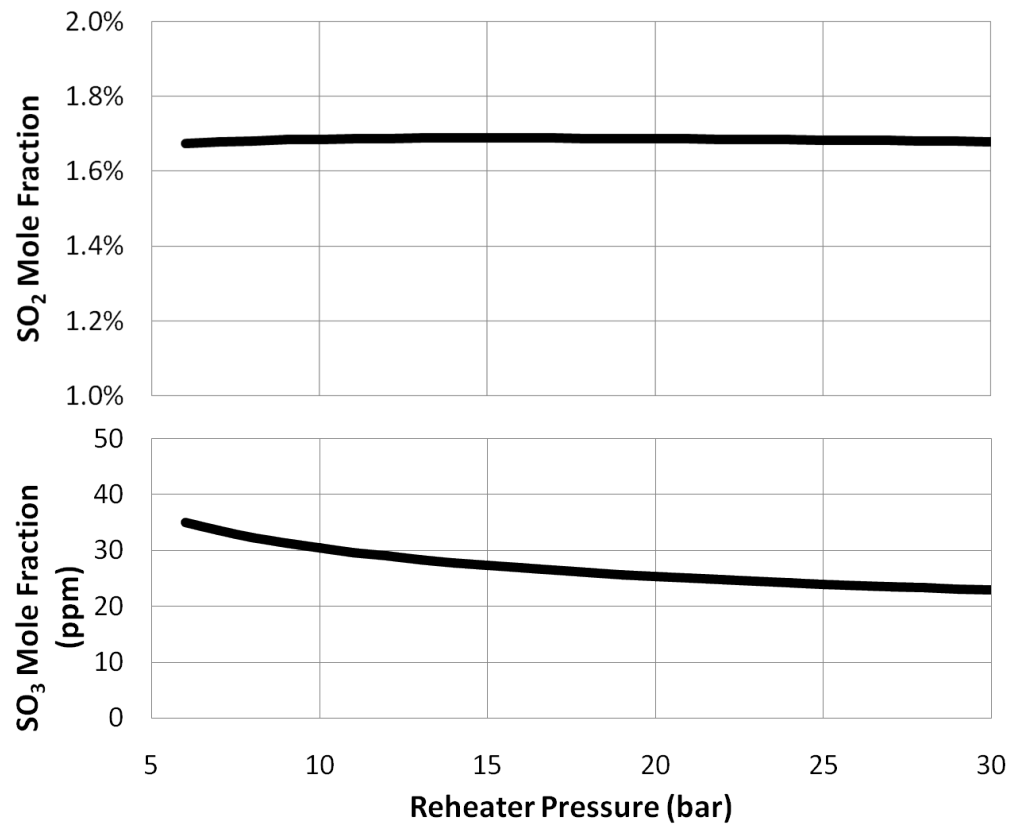


Figure 4: Effect of varying reheat pressure on the SO₂ and SO₃ concentrations at the exit of the reheater

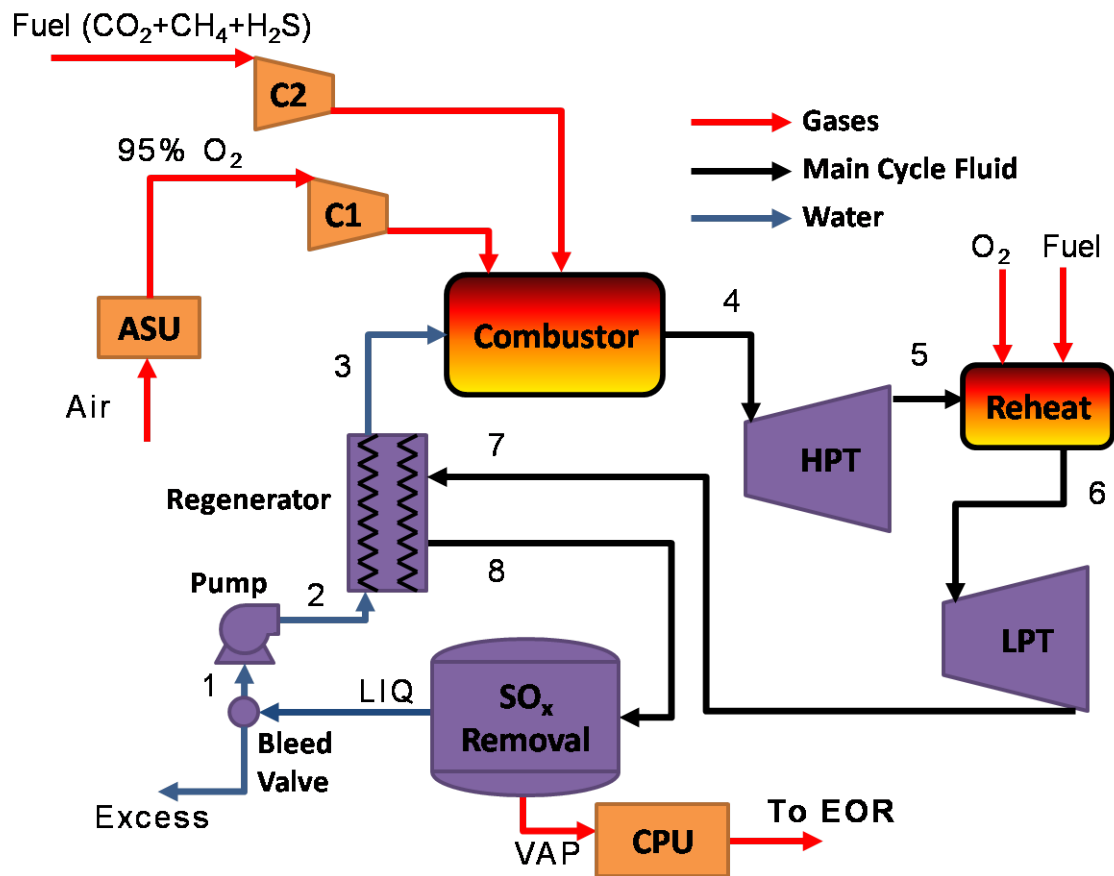


Figure 5: Overall process layout for the sour gas water cycle with SO_x removal

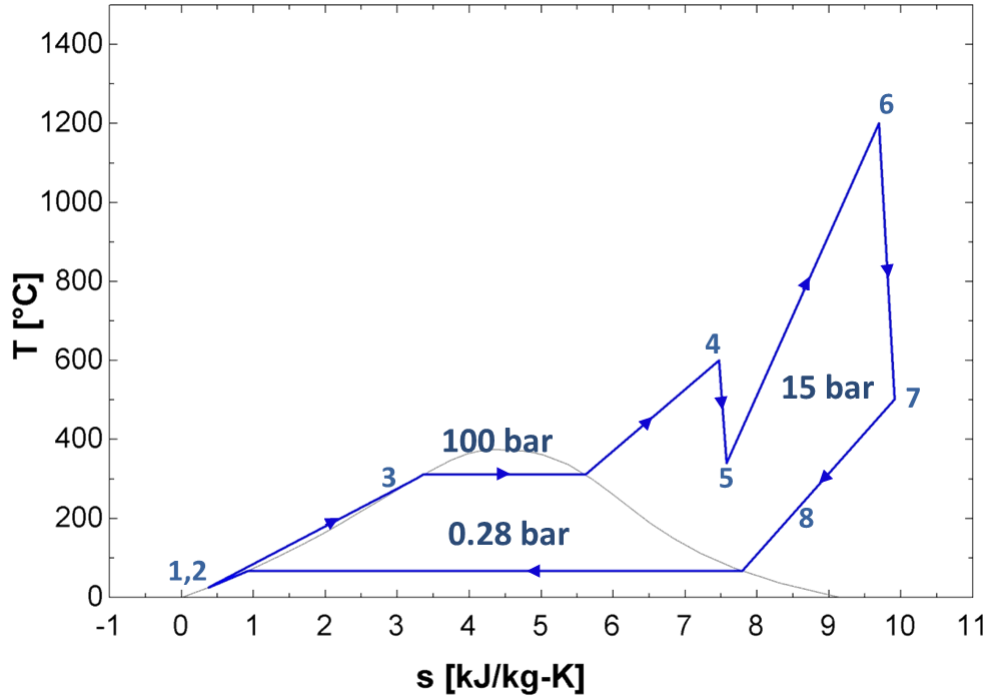


Figure 6: T-s diagram of the sour gas water cycle with SO_x removal

251 the dilution medium in the combustor. The vapor stream exiting the SO_x removal system, is sent to the
 252 CO_2 purification unit (CPU) and compressed up to 110 bars to a capture-ready carbon dioxide stream. The
 253 CPU removes the inert compounds from the working fluid (Ar & N_2).

254 The efficiency of this cycle was found to be 36.1%, about 4.5% points below the acid resistance cycle.
 255 Reasons include the slightly lower LPT pressure ratio and the energy requirement for the SO_x removal.

256 3.3.1. SO_x Removal System

257 The SO_x removal system modeled for the sour gas cycles, was based on the wet flue gas desulfurization
 258 (FGD) techniques [37]. The removal of these sulfur compounds prior to the working fluid condensing, allows
 259 us to limit the use of expensive acid resistant materials.

260 This system removes the S-compounds from the working fluid by reacting it with a lime solution (CaO
 261 + H_2O) and removing the byproducts as solid salts. The lime solution comes into direct contact with the
 262 working fluid and condenses the water and some SO_2 and SO_3 . The SO_x compounds dissociate in the water
 263 to form ions and these react with the calcium ions present in the lime solution resulting in the formation of
 264 salts which eventually neutralize the effect of the acid. The salt formation creates a concentration gradient
 265 which drives more SO_2 and SO_3 to condense and dissolve in the water, thus prompting further flue gas
 266 desulfurization.

267 Figure 7 shows the system layout with all the components of the SO_x removal system. This system is
 268 similar to the direct contact condenser described by Zebian et al. [38]. The main goals of the design are to:

- 269 1. Keep the gas exit SO_2 concentration < 100 ppm (EOR constraints [39])
 270 2. Keep the liquid exit $\text{pH} \simeq 7$

271 Before the CaO can react with the SO_2 and SO_3 , both must be broken down into their respective ions.
 272 This is accomplished by dissolving the lime in water, which dissociates into Ca^{2+} , and spraying it into the
 273 flue gases to dissolve the SO_2 . When the SO_2 condenses, it ionizes to form SO_3^{2-} . Similarly, when the SO_3
 274 reacts with water it forms H_2SO_4 which then ionizes and forms SO_4^{2-} . These ions react with the Ca^{2+} and
 275 water to form salts. The corresponding reactions are shown below. These reactions, commonly used in FGD
 276 systems, were found to be mainly dependent on the amount of lime (CaO) input to the system. This amount
 277 is adjusted in order to achieve the two goals mentioned above. Reactions 9 and 10 are very important in
 278 the desulfurization process because the calcium salts are formed which are then removed as solids from the
 279 system, creating a gradient which furthers the dissolution of the calcium and sulfur compounds. The H_2SO_4
 280 is formed through reaction 1 by reacting all of the SO_3 with the water in the working fluid. The water is an
 281 essential part of the FGD process and as such the appropriate amount was chosen to allow for all of these
 282 ionization reactions to take place. The system's operating pressure had a major impact on the efficiency and
 283 as such was chosen to maximize the cycle efficiency.



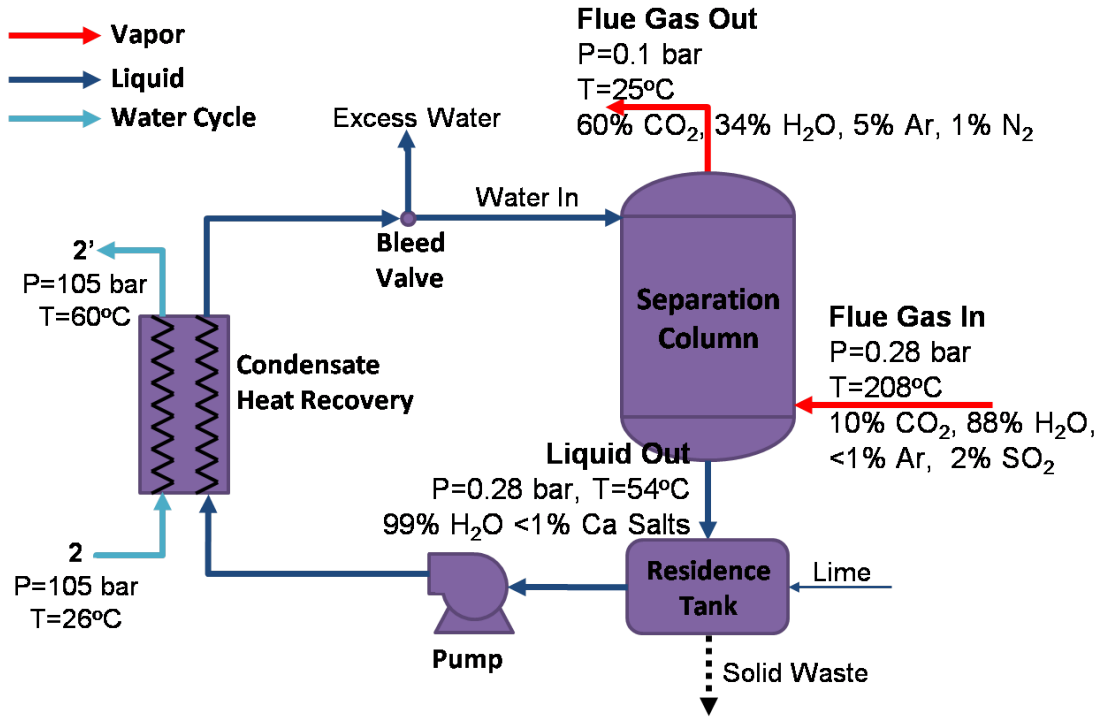
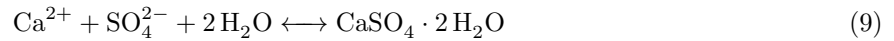


Figure 7: SO_x removal system implemented in the water cycle showing the operating conditions



284 The conditions and stream compositions for the system are also shown in Figure 7. The flue gas comes in
 285 directly from the exit of the regenerator, with a pressure that was chosen in order to maximize cycle efficiency
 286 and minimize pressure drop through the column. They exit the column at a temperature of around 49°C
 287 before being cooled further down to the condenser (not shown) temperature of 25°C . The gas flow rate
 288 coming in, for this case is 41 kg/s . On the other side, the water (+ lime) enters the column at the top with a
 289 mass flow rate of 625 kg/s after 4% of it was removed as excess in the bleed valve. The exiting liquid mixture
 290 at 54°C is sent to the residence tank where lime is added and the solids are removed. These reactions are
 291 exothermic and so the liquid temperature increases as it exits the tank. The amount of lime necessary for
 292 this cycle at these conditions was found to be 1.9 kg/s , this is the amount necessary for the “Liquid Out”
 293 stream to have neutral pH. The reagent stoichiometry, defined as $\frac{\text{moles}_{\text{reagent}}}{\text{moles}_{\text{S-removed}}}$, for the SO_x removal system
 294 was found to be 1.03 which is also exactly what traditional wet FGD systems operate at [40].

295 3.4. $H_2O + CO_2$ Recycle

296 In addition to the two sour gas water cycles, a cycle utilizing both H_2O and CO_2 recirculations was
297 modeled to determine its performance when using sour gas as the fuel. This cycle has a similar layout to
298 the Graz cycle [24], but with some modifications due to the new fuel composition and nature of the working
299 fluid.

300 This cycle consists of a high temperature Brayton cycle and a low temperature Rankine cycle. In this
301 case the main cycle working fluid exiting the compressor along with pure steam from the rankine cycle, is
302 recycled to the combustor to act as the dilution medium. This results in a mixture of about 15% CO_2 , 80%
303 H_2O , 3% SO_2 , 1% Ar by volume leaving the combustor at state 3. Since some CO_2 and SO_2 is also recycled,
304 unlike the previous two cycles, this results in higher compositions of both of these in the main cycle fluid:
305 15% vs. 10% CO_2 and 3% vs. 2% SO_2 .

306 The efficiency of this cycle with sour gas as the fuel, was calculated to be 24.8%, compared to a methane-
307 based cycle which had an efficiency of 46.0%. This significant difference in efficiencies was due to the sulfur
308 present in the working fluid which had a significant impact on the dew point of the sour gas cycle fluid.
309 In the condenser, not all of the water is sent to the pump because depending on that recycle ratio, the
310 composition of the working fluid of the cycle changes, and in turn affects the dew point of that stream. If
311 that recycle ratio is too high then condensation occurs in the LPT which for this working fluid (containing
312 H_2SO_4) would be severe. To prevent this condensation, a lower recycle ratio was chosen which resulted in a
313 significant decrease in efficiency as was shown. Using this result, we concluded that this type of cycle with
314 both H_2O and CO_2 recirculation is not a good option to use with sour gas as the fuel and as such this cycle
315 was not discussed further.

316 4. Results and Discussion

317 4.1. T - s Diagrams

318 Comparing the T - s diagrams of the two sour gas water cycles, shown in Figures 2 and 6, the low pressure
319 line is slightly higher for the SO_x Removal cycle than for the Acid Resistance cycle because of the higher
320 pressure required for the SO_x removal system (0.3 vs 0.1 bar). As can be seen, the area inside the T - s
321 diagram for the SO_x removal cycle is smaller and so we can estimate the efficiency to be lower, which is
322 indeed the case. However, this may not always be the case since work is proportional to the area inside a
323 T - s diagram only for ideal cycles.

324 4.2. Recycle Ratio

325 Figure 8 compares the recycle ratios for the systems. Since the Acid Resistance cycle has latent heat
326 recovery in the regenerator, there is more energy available to transfer to the water stream being preheated.
327 Thus the recycle ratio is higher for that cycle in order to recuperate that energy.

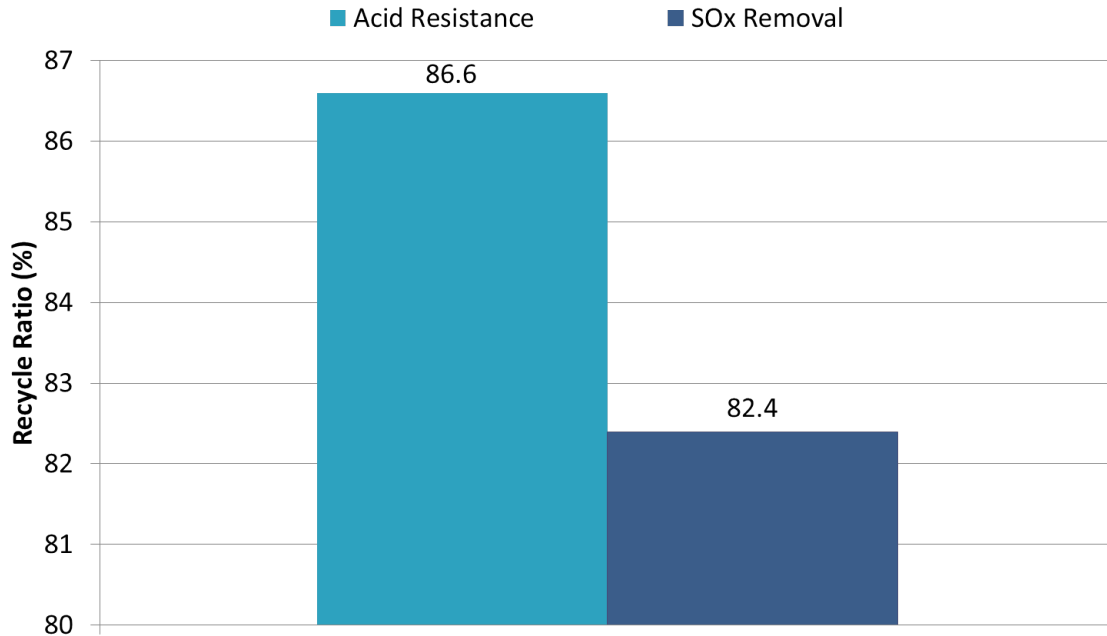


Figure 8: Recycle ratio comparison for the sour gas water cycles

328 4.3. Working Fluid

329 The working fluid of both cycles is the same at both the combustor and reheater exits. At the combustor
 330 exit, the composition is mainly: 5% CO₂, 93% H₂O, 1% SO₂, whereas at the reheater exit, it is mainly:
 331 10% CO₂, 88% H₂O, 2% SO₂. Since the same working fluid is being recycled (liquid water) the exit of the
 332 combustor for both cycles also have the same compositions. At the reheater exit, the working fluids had
 333 higher SO₂ and CO₂ concentrations because the product gases from the combustion of fuel and oxygen are
 334 high in those two compounds. Therefore, when it mixes with the incoming working fluid, the total mole
 335 fraction of SO₂ and CO₂ goes up and H₂O goes down at the exit.

336 4.4. Sulfur Compounds Formation

337 The important sulfur compounds concentrations at every point in the cycle for the two sour gas water
 338 cycles are shown in Table 2. The state numbers shown refer to those in Figures 1 and 5. Firstly, it is evident
 339 that SO₂ is the major sulfur compound formed in these cycles with SO₃ and H₂SO₄ concentrations at ppm
 340 levels. All of the S-compounds fractions are about the same for both cycles at most points in the cycles, since
 341 the same type of diluent is recycled to the combustor. One important difference in the two cycles happens
 342 at states 7-8. For the acid resistance cycle, process 7-8 is the condensation step occurring in the regenerator
 343 which as can be seen, significantly increases the H₂SO₄ concentration in the working fluid by about 2 orders
 344 of magnitude due to the reaction of SO₃ with H₂O.

States	P (bar)		T (°C)		SO ₂ (%)		SO ₃ (ppm)		H ₂ SO ₄ (ppm)	
	(a)	(b)	(a)	(b)	(a)	(b)	(a)	(b)	(a)	(b)
1	0.09	0.27	25	25	0	0	0	0	0	0
2	105.3	105.3	26.0	26.1	0	0	0	0	0	0
3	100	100	247.1	260.5	0	0	0	0	0	0
4	90	90	600	600	0.94%	0.92%	0.034	0.034	0.070	0.069
5	15	15	341.4	341.4	0.94%	0.92%	0.034	0.034	0.070	0.069
6	14.1	14.1	1200	1200	1.70%	1.68%	27	27	0.057	0.056
7	0.10	0.28	380.1	502.0	1.70%	1.68%	27	27	0.057	0.056
8	0.10	0.275	47	208	1.70%	1.68%	0	27	27.5	0.056

Table 2: Stream results and sulfur compounds compositions (mole fractions): (a) acid resistance cycle and (b) SO_x removal cycle

345 For the other cycle, however the H₂SO₄ concentration stays the same during process 7-8 since this cycle
346 prevents the working fluid from condensing in order to limit the use of acid-resistant equipment. But after
347 state 8, the stream is then sent to the SO_x removal system which is why the temperature and pressure of
348 this stream are higher than those of the acid resistance cycle. This removes all of these sulfur compounds
349 from that cycle's working fluid before then condensing the sulfur-free stream and then sending it to the CPU
350 for inert gas removal. In the acid resistance cycle however, after state 8 the fluid is condensed even further
351 and the gas then sent to the CPU where the S-compounds are removed. But at that point, the H₂SO₄
352 concentration increases even further due to the lower temperature and lower vapor quality. Finally it can
353 also be seen that the SO_x fractions at the exit of the combustor (state 4) increase by about 1-2 orders of
354 magnitudes after the reheat step (state 6) due to the combustion of more H₂S.

355 4.5. Pressure Drop Sensitivity

356 The following analysis presents the results from a combustor and reheater pressure drop sensitivity study
357 that was done on the two cycles to determine their effect on the cycles' efficiencies. As can be seen from
358 Figures 9 and 10, the combustor and reheater pressure drops were found not to have a significant effect on
359 the efficiency.

360 The average slopes of the two graphs for the combustor analysis were determined to be the same at
361 about -0.02 Eff%/Pdrop%. For the reheater analysis, the values were different with slopes of -0.03 and -0.04
362 Eff%/Pdrop% observed for the Acid Resistance and SO_x Removal cycles respectively. As can be seen, the
363 reheater pressure drop had a slightly bigger impact on the efficiency for the two cycles. This is because the
364 LPT contributes more to the net power output and so varying its inlet pressure (by changing the reheater

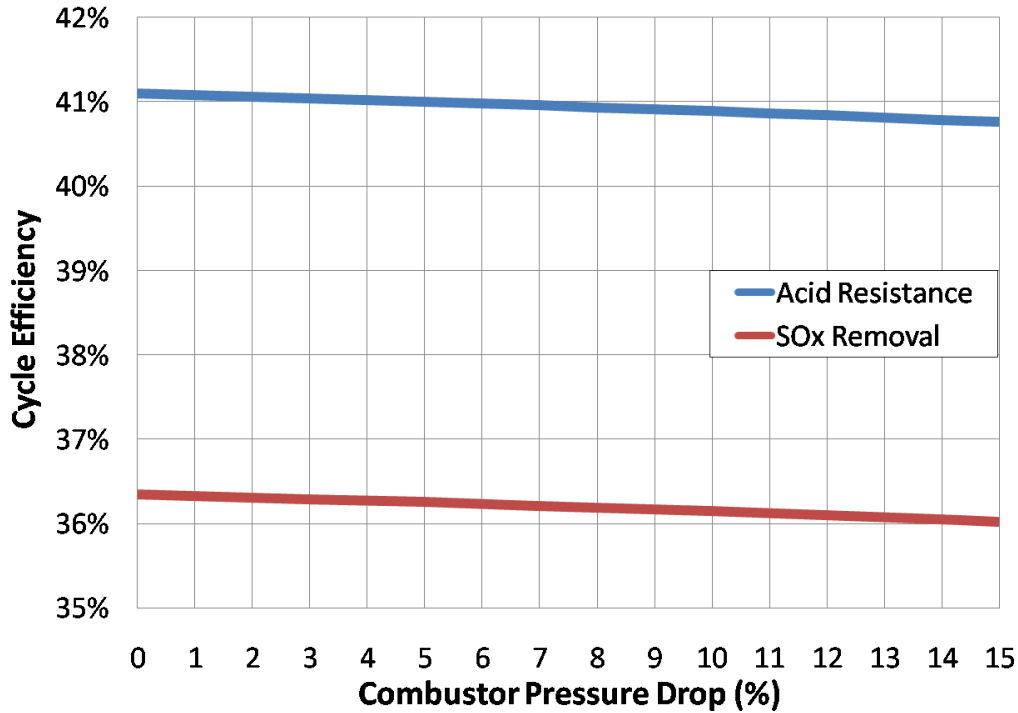


Figure 9: Effect of combustor pressure drop on net cycle efficiency for the sour gas water cycles

365 pressure drop), impacts the efficiency more significantly. In our analysis and for the results shown next,
 366 the default values of the pressure drops for the combustors and reheaters were taken to be 10% and 6%
 367 respectively.

368 4.6. Efficiency and Power Breakdown

369 The final and most important technical comparison of these cycles is shown in Figure 11. Details of the
 370 power generated and consumed by the different components in the cycles are shown where they are expressed
 371 as a function of the (total) heat input to the cycle (based on the fuel's LHV) in order to non-dimensionalize
 372 the results. The heat input to the two cycles were about 137 MW and 139 MW respectively. These values
 373 were calculated by Aspen Plus using a technical constraint in order to fix the exit temperature of the reheater.

374 The turbine work for the SO_x Removal cycle is lower than the Acid Resistance cycle because the low
 375 pressure is 0.3 compared to 0.1 bar. Therefore, there is a smaller pressure ratio across the turbine (LPT) and
 376 as a result less power is produced in the turbines. Due to the very low pressures that these turbines expand
 377 to, small changes in that outlet pressure can have a big impact on the power produced by the turbine. For
 378 these cycles, the density of the LPT outlet stream for the SO_x Removal cycle is about 2.4 times greater
 379 than that of the Acid Resistance cycle. This very large density difference at the turbine outlet results in a

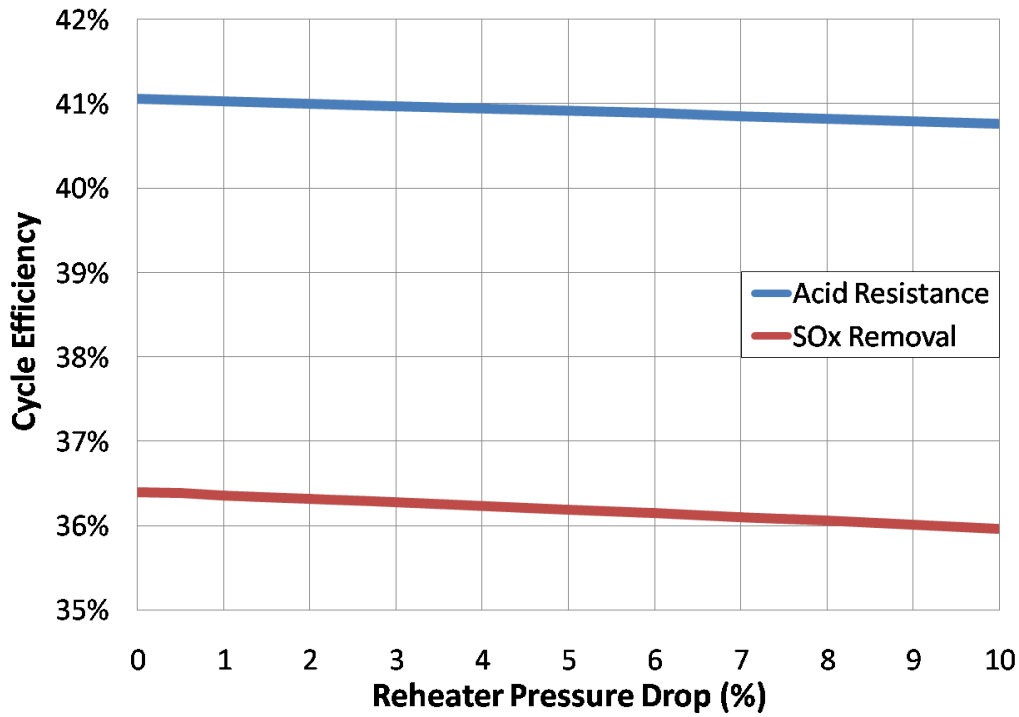


Figure 10: Effect of reheater pressure drop on net cycle efficiency for the sour gas water cycles

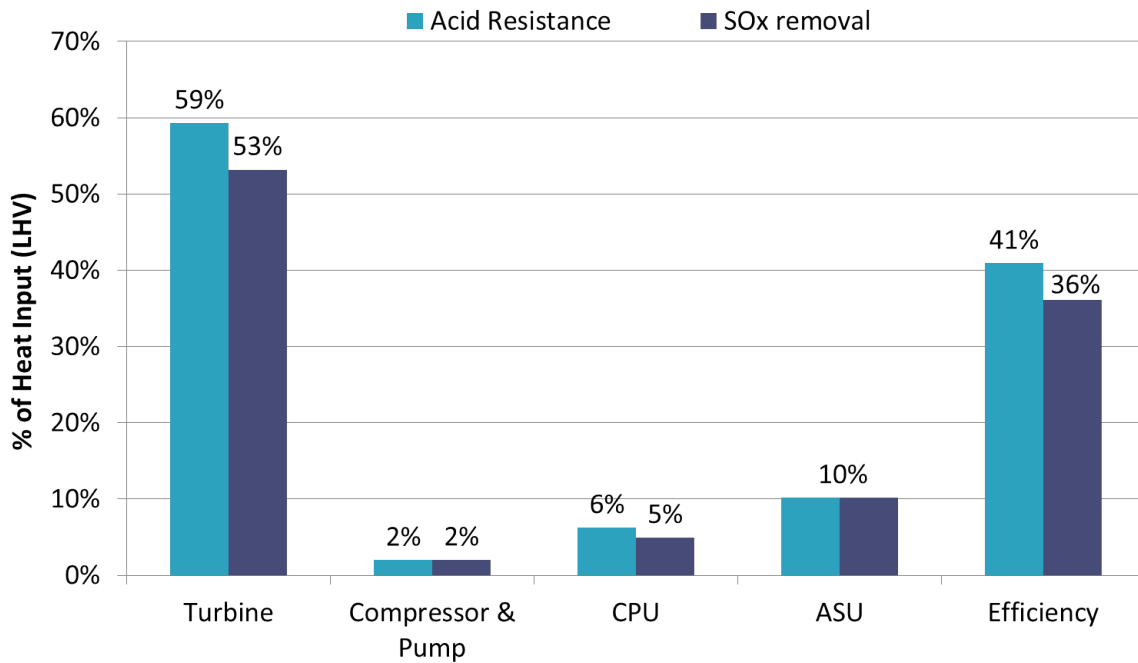


Figure 11: Power breakdown for the sour gas water cycles

380 large difference in the turbine works which results in the 6% efficiency reduction shown in the figure. The
381 compressors and pump work for the two cycles are both very low, only resulting in a 2% efficiency loss.

382 The CPU (CO₂ Purification Unit) and ASU (Air Separation Unit) power inputs are fairly similar for
383 both cycles. But a slightly smaller power is required in the CPU for the SO_x Removal cycle because the
384 SO_x compounds are already being removed in the main cycle before entering the CPU. Therefore, there is
385 no further efficiency penalty associated with this process, unlike the other cycle. The ASU is also the largest
386 power consumer for both cycles as is common in oxy-combustion systems [7].

387 There is also an efficiency drop for the SO_x Removal cycle because of the inability to recuperate all of the
388 latent energy from the hot working fluid in the regenerator. Since the hot working fluid doesn't condense for
389 this cycle, less heat is transferred to the liquid water being preheated and so this overall leads to a smaller
390 net power output and also lower efficiency.

391 4.7. Fuel Composition Sensitivity

392 4.7.1. H₂S Variations

393 Another important technical assessment of the sour gas cycle is the effect of fuel composition on the cycle
394 performance. The most critical components in sour gas are H₂S and CO₂ since these two compositions vary
395 widely depending on the lifetime and geography of the gas field. The acid resistance sour gas cycle perfor-
396 mance was studied using the assumptions shown previously in Table 1 but with varying fuel compositions.
397 Firstly, the results of H₂S variations in the fuel on the cycle are shown in Table 3. The cycle cost was found
398 not to be sensitive to fuel composition changes and as such those results are not reported.

399 As the H₂S content in the fuel was increased, the recycle ratio also went up due to the change in heat
400 capacity of the working fluid. As can be seen, the CO₂ and H₂O fractions decrease because the amount
401 of methane in the fuel is decreasing which lowers the heat capacity as well. To compensate for that slight
402 decrease in heat capacity, more of the diluent needs to be recycled to achieve the combustor and reheater
403 exit temperatures. The amount of SO₂ in the working fluid increases due to the increase in H₂S content
404 in the fuel. This increase in SO₂ in the working fluid did not have a significant impact on the efficiency as
405 can be seen. The very slight increase is attributed to the fact that a smaller amount of CO₂ is now in the
406 working fluid, as the H₂S increases, which slightly decreases the power requirement in the CPU. However
407 the bigger impact from that increase in SO₂ content in the fluid, is in the SO_x removal system in the CPU
408 where now more lime has to be used to remove more SO₂ from the working fluid.

409 4.7.2. CO₂ Variations

410 Next, variations in the CO₂ content in the fuel was studied and the results are shown in Table 4. Once
411 again, the recycle ratio goes up with CO₂ content as the working fluid's heat capacity decreases due to the
412 decrease in the H₂O fraction. Since the fuel's methane content is lower, the H₂O fraction in the working

Parameter		Cycle Results		
Fuel Composition (mol%)	CH ₄	84	70	55
	H ₂ S	1	15	30
	CO ₂	15	15	15
Recycle Ratio (%)		85.7	86.6	87.7
Working Fluid Composition (mol%)	CO ₂	10.57	9.60	8.47
	H ₂ O	88.4	87.7	86.9
	SO ₂	0.11	1.70	3.63
	Ar	0.80	0.81	0.83
	N ₂	0.15	0.16	0.16
Net Efficiency (%)		40.8	40.9	41.1

Table 3: Fuel composition sensitivity results to H₂S variations

fluid also goes down even though the recycle ratio is increasing. However, the CO₂ and SO₂ fractions both increase with increasing CO₂ composition with the CO₂ fraction in the fluid reaching 16% at a CO₂ fuel composition of 50%. As in the previous fuel sensitivity, when the CO₂ content in the working fluid increases, the CPU power requirement goes up which, in this case, greatly affects the net cycle efficiency. The efficiency decreases by more than 3% as the CO₂ content in the fuel increases from 1 to 50%. Also the working fluid's heat capacity at lower CO₂ fuel concentrations is higher which increases the turbine power outputs and also further increasing the net efficiency.

4.8. Cost of Electricity

A preliminary cost analysis was also performed to estimate the capital cost of these cycles. Using this, the levelized cost of electricity (LCOE) was then calculated to bring in the impact of the cycle efficiency.

A critical part of the cost estimation procedure was the selection of the material for the different components. All of the sour gas cycles have SO_x in the working fluid and sulfuric acid forms where this working fluid condenses. Therefore, in order to protect the equipment from corrosion, certain material must be used. The selection was made based on literature recommendations. An important consideration is the problem of hot corrosion, defined as “the accelerated corrosion, resulting from the presence of salt contaminants, such as Na₂SO₄, that combined to form molten deposits, which damage the protective surface oxides” [41]. This Na₂SO₄ comes from the reaction of SO₂ in the working fluid with small concentrations of NaCl which is usually present in the combustion air if the plant is located near a sea, or from other industrial pollutants present in air. To combat this issue, it was found that increasing the chromium content in the metal alloys or coatings would significantly improve the resistance of the material. More specifically, nickel-based alloys

Parameter		Cycle Results		
Fuel Composition (mol%)	CH ₄	84	70	35
	H ₂ S	15	15	15
	CO ₂	1	15	50
Recycle Ratio (%)		85.7	86.6	92.3
Working Fluid Composition (mol%)	CO ₂	8.27	9.60	16.30
	H ₂ O	89.30	87.70	79.88
	SO ₂	1.46	1.70	2.88
	Ar	0.82	0.81	0.79
	N ₂	0.16	0.16	0.15
Net Efficiency (%)		41.4	40.9	38.1

Table 4: Fuel composition sensitivity results to CO₂ variations

433 with chromium content greater than 15 wt.% were found to be more resistant to hot corrosion. The data
434 also suggested that increasing titanium helps improve the material's hot corrosion resistance [42]. Based on
435 these results, the best available material was chosen for the turbines and compressors that had SO₂ in the
436 working fluid. This choice of material affects the equipment costs as will be seen later in Table 6.

437 Another important part of the material selection was that for the acid equipment. This refers to the
438 components where the working fluid has condensed and sulfuric acid has formed. This is mainly for the
439 condensing heat exchangers (condenser and regenerator) and the absorber column in the SO_x removal system.
440 The model results presented show that the pH levels in these systems are expected to be very low thus making
441 their environments very aggressive. To combat acid corrosion, a corrosion resistant material must be chosen
442 that can withstand these extreme conditions. In a study by Shoemaker et al. [43] comparing the corrosion
443 resistance of stainless steel metals, Alloy 686 (Inconel®) was found to be very stable in highly corrosive
444 environments with sulfuric acid. As such, this material was chosen for the absorber shell cladding, and the
445 shell and tube materials in the heat exchangers.

446 Using these material selections, the total equipment costs were calculated for the two cycles, and these
447 were used as the Bare Erected Costs (BEC) when calculating the Levelized Cost of Electricity (LCOE). The
448 Levelized Cost of Electricity (LCOE) calculation was performed based on the guidelines and assumptions
449 discussed in the NETL report, Quality Guidelines for Energy System Studies: Cost Estimation Methodology
450 for NETL Assessments of Power Plant Performance [44] and the assumptions that were used in this study
451 are listed in Table 5. The results from this calculation are shown in Table 6 where the LCOE is shown for
452 all of the cycles. The impact of the cycle efficiency plays a big role because the fuel cost is included in the
453 LCOE calculations. We also considered what would happen in the limit that the fuel cost is very minimal

Parameter	Value
Engineering, Procurement and Construction Cost (EPCC)	
Engineering, Procurement and Construction (EPC) Contractor Services	9% of BEC
Total Plant Cost (TPC)	
Process Contingency	30% of EPCC
Project Contingency	25% of EPCC + Process Contingency
Total Overnight Cost (TOC)	
Owner's Costs	17.5% of TPC
Global Economic Assumptions	
Operational Period	25 years
Plant Capacity Factor	90%
Internal Rate of Return on Equity	10%
Income Tax Rate	38% Effective
Capital Depreciation	25 years, 200% declining balance
Variable O&M Costs Factor	1.5% of EPCC
Fixed O&M Costs Factor	3.5% of EPCC
Fuel Cost (only natural gas)	3 \$/MMBTU [47]
Escalation of COE (revenue), O&M Costs, Fuel Costs (nominal annual rate)	3%

Table 5: LCOE economic modeling assumptions

454 and thus ≈ 0 , and those results are also shown in the table.

455 Finally the costs of avoiding CO₂ emissions are shown in Table 6 for the two cycles and the two fuel
456 cost scenarios. The cost of CO₂ avoided is calculated as the difference in the LCOE between the plant with
457 CO₂ capture and that of a baseline plant without capture, divided by the difference in their CO₂ emissions
458 in kg/MWh [45]. This value represents the average cost (\$/ton) of reducing atmospheric CO₂ emissions by
459 one ton while producing one MWh of electricity. The baseline plant chosen in this analysis is the gas-fired
460 combined cycle without capture whose performance and costs were presented by Davison [46]. The choice of
461 the reference plant for this calculation is very important as it can greatly impact the avoided cost and thus
462 careful attention must be paid when comparing values from different technologies.

463 The two sour gas water cycles condense down to low pressures and so large heat exchangers are needed

464 for both which increases the equipment cost. Another issue with this low pressure is that in the CPU, more
465 compression is needed to recompress the CO₂ for EOR which also results in further cost penalties. The
466 SO_x removal cycle was found to have the overall lower cost than the acid resistance cycle. Costs savings are
467 achieved from the fact that the expensive acid resistant materials don't need to be used for this cycle since
468 the working fluid doesn't condense. Therefore, especially for the heat exchangers, the acid resistance cycle
469 requires much more expensive equipment. As can be seen in the table, these choice of materials for the acid
470 resistance cycle results in about a 20% increase in the equipment costs (BEC) over the SO_x removal case.
471 It is known that errors and uncertainties in the cost estimation will inevitably be introduced depending on
472 the method used or the source of the costs for components and materials. Thus we calculated the sensitivity
473 of the LCOE to errors in the BEC. It was found that 10% variations in the BEC, resulted in an 8% change
474 in the LCOE for both cycles. Similarly, 20% variations in the BEC caused the LCOE to change by 15% for
475 the acid resistance cycle and 16% for the SO_x removal cycle.

476 Also as it turns out, although the acid resistance cycle has a higher efficiency than the SO_x removal one,
477 its cycle cost is higher. Therefore a tradeoff would have to be made between cost and cycle performance.
478 From a purely economic point of view, the LCOE result suggests that the SO_x Removal cycle is a better
479 option than the Acid Resistance cycle. Since we don't have an exact number for the cost of the sour gas fuel,
480 we also considered the case where the fuel cost is taken be 0. The LCOE results go down as expected and this
481 helps give us a range of what one might expect depending on what the fuel cost is. Worst case, the LCOE
482 for the SO_x removal cycle would be around 126 \$/MWh and in the best case scenario this cost goes down
483 to 101 \$/MWh, a 20% decrease. The LCOE of a methane oxy-fuel cycle was found to be around 117 (2014
484 \$/MWh) [46]. Since the majority of the methane oxy-fuel cycle modeling in the literature don't include the
485 energy and cost penalties of the natural gas processing step, it is hard to make an accurate recommendation
486 on which sour gas treatment option is more feasible: pre-combustion gas sweetening vs burning gas directly.
487 However as can be seen, when the fuel is assumed to be very cheap (cost ≈ 0) the LCOE of the sour gas
488 cycles are actually both cheaper than the methane oxy-fuel cycle.

489 The cost of CO₂ avoided for the two cycles is calculated to be 184 \$/ton and 151 \$/ton for the acid
490 resistance and SO_x removal cycles respectively. Just as a reference, Davison [46] found the CO₂ avoided cost
491 for a natural gas oxy-fuel cycle to be 120 \$/ton (after conversion to 2014 \$). Therefore the cost of reducing
492 CO₂ emissions while supplying the same amount of electricity for the two sour gas cycles, only becomes
493 competitive and even cheaper than the natural gas cycle when the sour gas fuel cost is considered negligible,
494 as shown in Table 6. However due to the uncertainties and assumptions that go into the CO₂ avoided cost
495 calculation (ex. choice of reference plant), arguably the measure that is most relevant for technical, economic
496 and policy analyses, is the levelized cost of electricity.

Parameter	Unit	Acid Resistance	SO _x Removal
Net Power Output	MW	55.98	50.04
Net Efficiency	%	40.9	36.1
BEC	MM\$	131	103
EPCC	MM\$	143	112
TOC	MM\$	260	203
LCOE	2014 \$/MWh	138	126
LCOE (Fuel Cost = 0)	2014 \$/MWh	116	101
Cost of CO ₂ Avoided	\$/ton	184	151
Cost of CO ₂ Avoided (Fuel Cost = 0)	\$/ton	121	80

Table 6: Costing analysis results for the sour gas water cycles

5. Conclusions

Methane oxy-fuel water cycles have been extensively studied in the literature, whereas sour gas cycles have received no attention thus far. A detailed analysis of oxy-fuel water cycles fueled by sour gas has been performed in this study. The water cycles were subdivided into two configurations for addressing issues and limitations associated with the presence of sulfur compounds in the fuel.

An Acid Resistance and a SO_x Removal cycle were considered. It was found that the Acid Resistance cycle had the better efficiency of 40.9%, the main reason being the fact that the working fluid is allowed to condense in the regenerator and so some of its latent heat is recuperated. Changes in the CO₂ composition in the fuel was found to have a bigger impact on system performance than H₂S variations as it significantly affected the CPU power requirement and thus the efficiency.

A preliminary cost analysis was also done and the levelized cost of electricity was calculated to bring in the impact of cycle efficiency. It was found that the SO_x Removal cycle to be cheaper, as we had predicted, since the working fluid doesn't condense in the cycle and so less amounts of expensive acid resistant equipment need to be used. The LCOE of the SO_x Removal cycle was found to be 126 \$/MWh compared to 138 \$/MWh for the Acid Resistance cycle.

Since the fuel is cheap, sacrificing some efficiency points at the expense of a less costly system would not be a major issue. Therefore from this whole analysis it seems that the best process cycle to use is the SO_x Removal water cycle.

In future work, a similar analysis will be done that focuses on another type of cycle configuration namely, the sour gas combined cycles.

517 6. Acknowledgments

518 The authors would like to thank Siemens for sponsoring this work. Aspen Plus[®] was generously provided
519 by Aspen Technology.

520 References

- 521 [1] United States Environmental Protection Agency (EPA), Global Greenhouse Gas Emissions Data, [http:](http://www.epa.gov/climatechange/ghgemissions/global.html)
522 [//www.epa.gov/climatechange/ghgemissions/global.html](http://www.epa.gov/climatechange/ghgemissions/global.html) (2013).
- 523 [2] Energy Information Administration, International Energy Outlook, report DOE/EIA-0484 (2013).
- 524 [3] D. Gielen, The future role of CO₂ capture and storage: results of the IEA-ETP Model, report no.
525 EET/2003/04, <http://www.iea.org/textbase/papers/2003/eet04.pdf> (2003).
- 526 [4] V. Tola, A. Pettinau, Power generation plants with carbon capture and storage: A techno-economic
527 comparison between coal combustion and gasification technologies, Applied Energy 113 (0) (2014) 1461
528 – 1474. doi:<http://dx.doi.org/10.1016/j.apenergy.2013.09.007>.
529 URL <http://www.sciencedirect.com/science/article/pii/S0306261913007538>
- 530 [5] A. Pettinau, F. Ferrara, C. Amorino, Techno-economic comparison between different technologies for
531 a {CCS} power generation plant integrated with a sub-bituminous coal mine in italy, Applied Energy
532 99 (0) (2012) 32 – 39. doi:<http://dx.doi.org/10.1016/j.apenergy.2012.05.008>.
533 URL <http://www.sciencedirect.com/science/article/pii/S030626191200356X>
- 534 [6] M. A. Nemitallah, M. A. Habib, Experimental and numerical investigations of an atmospheric diffusion
535 oxy-combustion flame in a gas turbine model combustor, Applied Energy 111 (0) (2013) 401 – 415.
536 doi:<http://dx.doi.org/10.1016/j.apenergy.2013.05.027>.
537 URL <http://www.sciencedirect.com/science/article/pii/S0306261913004285>
- 538 [7] B. Metz, O. Davidson, H. De Coninck, M. Loos, L. Meyer, IPCC, 2005: IPCC special report on carbon
539 dioxide capture and storage. Prepared by Working Group III of the Intergovernmental Panel on Climate
540 Change, Cambridge, United Kingdom and New York, NY, USA, 442 pp.
- 541 [8] US Energy Information Administration, Average national levelized costs for generating technologies in
542 2018 (2011).
- 543 [9] F. Garca-Labiano, L. de Diego, P. Gayn, A. Abad, A. Cabello, J. Adnez, G. Sprachmann, Energy
544 exploitation of acid gas with high {H₂S} content by means of a chemical looping combustion system,
545 Applied Energy 136 (0) (2014) 242 – 249. doi:<http://dx.doi.org/10.1016/j.apenergy.2014.09>.

- 546 041.
547 URL <http://www.sciencedirect.com/science/article/pii/S0306261914009891>
- 548 [10] W. Burgers, P. Northrop, H. Kheshgi, J. Valencia, Worldwide development potential for sour gas,
549 Energy Procedia 4 (0) (2011) 2178 – 2184, 10th International Conference on Greenhouse Gas Control
550 Technologies. doi:<http://dx.doi.org/10.1016/j.egypro.2011.02.104>.
551 URL <http://www.sciencedirect.com/science/article/pii/S1876610211003018>
- 552 [11] Total, France, Sour gas: A history of expertise (2007).
- 553 [12] ICO2N, Enhanced Oil Recovery - a significant economic opportunity to help jump-start Carbon Cap-
554 ture and Storage, [http://www.ico2n.com/wp-content/uploads/2011/08/ICO2N-EOR-Fact-Sheet.](http://www.ico2n.com/wp-content/uploads/2011/08/ICO2N-EOR-Fact-Sheet.pdf)
555 pdf (2011).
- 556 [13] A. P. Shroll, S. J. Shanbhogue, A. F. Ghoniem, Dynamic-stability characteristics of premixed methane
557 oxy-combustion, Journal of Engineering for Gas Turbines and Power 134 (5) (2012) 051504.
- 558 [14] C. Liu, G. Chen, N. Sipczyk, M. Assadi, X. Bai, Characteristics of oxy-fuel combustion in gas turbines,
559 Applied Energy 89 (1) (2012) 387 – 394, special issue on Thermal Energy Management in the Process
560 Industries. doi:<http://dx.doi.org/10.1016/j.apenergy.2011.08.004>.
561 URL <http://www.sciencedirect.com/science/article/pii/S0306261911004995>
- 562 [15] J. D. Figueroa, T. Fout, S. Plasynski, H. McIlvried, R. D. Srivastava, Advances in {CO₂} capture tech-
563 nologythe u.s. department of energy’s carbon sequestration program, International Journal of Green-
564 house Gas Control 2 (1) (2008) 9 – 20. doi:[http://dx.doi.org/10.1016/S1750-5836\(07\)00094-1](http://dx.doi.org/10.1016/S1750-5836(07)00094-1).
565 URL <http://www.sciencedirect.com/science/article/pii/S1750583607000941>
- 566 [16] W. Sanz, H. Jericha, B. Bauer, E. Gattlich, Qualitative and quantitative comparison of two promising
567 oxy-fuel power cycles for CO₂ capture, Journal of Engineering for Gas Turbines and Power 130 (3)
568 (2008) 031702.
- 569 [17] G. Corchero, V. Timon, J. Montanes, A natural gas oxy-fuel semiclosed combined cycle for zero CO₂
570 emissions: a thermodynamic optimization, Proceedings of the Institution of Mechanical Engineers, Part
571 A: Journal of Power and Energy 225 (4) (2011) 377–388.
- 572 [18] O. Bolland, P. Mathieu, Comparison of two CO₂ removal options in combined cycle power plants,
573 Energy Conversion and Management 39 (1618) (1998) 1653 – 1663. doi:[http://dx.doi.org/10.](http://dx.doi.org/10.1016/S0196-8904(98)00078-8)
574 [1016/S0196-8904\(98\)00078-8](http://dx.doi.org/10.1016/S0196-8904(98)00078-8).
575 URL <http://www.sciencedirect.com/science/article/pii/S0196890498000788>

- 576 [19] H. M. Kvamsdal, K. Jordal, O. Bolland, A quantitative comparison of gas turbine cycles with capture,
577 Energy 32 (1) (2007) 10 – 24. doi:<http://dx.doi.org/10.1016/j.energy.2006.02.006>.
578 URL <http://www.sciencedirect.com/science/article/pii/S036054420600048X>
- 579 [20] F. Franco, T. Mina, G. Woolatt, M. Rost, O. Bolland, Characteristics of cycle components for CO₂
580 capture, in: Proceedings of 8th International Conference on Greenhouse Gas Control Technologies,
581 Trondheim, Norway, 2006.
- 582 [21] R. Anderson, H. Brandt, H. Mueggenburg, A power plant concept which minimizes the cost of carbon
583 dioxide sequestration and eliminates the emission of atmospheric pollutants, Greenhouse Gas Control
584 Technologies (1999) 59.
- 585 [22] R. E. Anderson, S. MacAdam, F. Viteri, D. O. Davies, J. P. Downs, A. Paliszewski, Adapting gas
586 turbines to zero emission oxy-fuel power plants, in: ASME Turbo Expo 2008: Power for Land, Sea, and
587 Air, American Society of Mechanical Engineers, 2008, pp. 781–791.
- 588 [23] C. Gou, R. Cai, H. Hong, An advanced oxy-fuel power cycle with high efficiency, Proceedings of the
589 Institution of Mechanical Engineers, Part A: Journal of Power and Energy 220 (4) (2006) 315–325.
- 590 [24] W. Sanz, H. Jericha, F. Luckel, F. Heitmeir, A further step towards a Graz cycle power plant for CO₂
591 capture, ASME Paper GT2005-68456, ASME Turbo Expo.
- 592 [25] D. Thomas, S. Benson, Carbon Dioxide Capture for Storage in Deep Geologic Formations-Results from
593 the CO₂ Capture Project: Vol 1-Capture and Separation of Carbon Dioxide from Combustion, Vol
594 2-Geologic Storage of Carbon Dioxide with Monitoring and Verification, Elsevier, 2005.
- 595 [26] Carpenter, 20Cb-3 Stainless Technical Data Sheet (2011).
- 596 [27] M. Steinberg, K. Schofield, The controlling chemistry of surface deposition from sodium and potassium
597 seeded flames free of sulfur or chlorine impurities, Combustion and Flame 129 (4) (2002) 453 – 470.
598 doi:[http://dx.doi.org/10.1016/S0010-2180\(02\)00363-2](http://dx.doi.org/10.1016/S0010-2180(02)00363-2).
599 URL <http://www.sciencedirect.com/science/article/pii/S0010218002003632>
- 600 [28] Aspen Technologies Inc, Aspen Plus (2012).
- 601 [29] D. Fleig, M. U. Alzueta, F. Normann, M. Abin, K. Andersson, F. Johnsson, Measurement and modeling
602 of sulfur trioxide formation in a flow reactor under post-flame conditions, Combustion and Flame 160 (6)
603 (2013) 1142 – 1151. doi:<http://dx.doi.org/10.1016/j.combustflame.2013.02.002>.
604 URL <http://www.sciencedirect.com/science/article/pii/S0010218013000448>

- 605 [30] D. Bongartz, A. F. Ghoniem, Chemical kinetics mechanism for oxy-fuel combustion of mixtures of
606 hydrogen sulfide and methane, *Combustion and Flame* (0) (2014) -. doi:[http://dx.doi.org/10.](http://dx.doi.org/10.1016/j.combustflame.2014.08.019)
607 [1016/j.combustflame.2014.08.019](http://dx.doi.org/10.1016/j.combustflame.2014.08.019).
608 URL <http://www.sciencedirect.com/science/article/pii/S0010218014002661>
- 609 [31] R. Anderson, F. Viteri, R. Hollis, A. Keating, J. Shipper, G. Merrill, C. Schillig, S. Shinde, J. Downs,
610 D. Davies, et al., Oxy-fuel gas turbine, gas generator and reheat combustor technology development
611 and demonstration, in: *ASME Turbo Expo 2010: Power for Land, Sea, and Air*, American Society of
612 Mechanical Engineers, 2010, pp. 733–743.
- 613 [32] S. H. Tak, S. K. Park, T. S. Kim, J. L. Sohn, Y. D. Lee, Performance analyses of oxy-fuel power
614 generation systems including CO₂ capture: comparison of two cycles using different recirculation fluids,
615 *Journal of mechanical science and technology* 24 (9) (2010) 1947–1954.
- 616 [33] Clean Energy Systems, <http://www.cleanenergysystems.com>, accessed: 02-24-2014.
- 617 [34] C. Fu, T. Gundersen, Using exergy analysis to reduce power consumption in air separation units for
618 oxy-combustion processes, *Energy* 44 (1) (2012) 60–68.
- 619 [35] J. Hong, G. Chaudhry, J. Brisson, R. Field, M. Gazzino, A. F. Ghoniem, Analysis of oxy-fuel combustion
620 power cycle utilizing a pressurized coal combustor, *Energy* 34 (9) (2009) 1332–1340.
- 621 [36] C. O. Iloeje, Process modeling and analysis of CO₂ purification for oxy-coal combustion, Master’s thesis,
622 Massachusetts Institute of Technology, Cambridge, MA, USA (2011).
- 623 [37] R. K. Srivastava, W. Jozewicz, Flue gas desulfurization: The state of the art, *Journal of the Air &*
624 *Waste Management Association* 51 (12) (2001) 1676–1688.
- 625 [38] H. Zebian, N. Rossi, M. Gazzino, D. Cumbo, A. Mitsos, Optimal design and operation of pressurized
626 oxy-coal combustion with a direct contact separation column, *Energy* 49 (2013) 268–278.
- 627 [39] U.S. Department of Energy - National Energy Technology Laboratory, Advanced oxy-combustion tech-
628 nology development and scale-up for new and existing coal-fired power plants (DE-FOA-0000636).
- 629 [40] R. Gaikwad, W. Boward, W. DePriest, Wet flue gas desulfurization technology evaluation, Sargent and
630 Lundy, LLC, prepared for National Lime Association, project (11311-000).
- 631 [41] N. Eliaz, G. Shemesh, R. Latanision, Hot corrosion in gas turbine components, *Engineering failure*
632 *analysis* 9 (1) (2002) 31–43.
- 633 [42] G. Y. Lai, High-temperature corrosion and materials applications, ASM International, 2007.

- 634 [43] L. Shoemaker, J. R. Crum, Experience in effective application of metallic materials for construction of
635 fgd systems (2010).
- 636 [44] National Energy Technology Laboratory, Quality guidelines for energy system studies: cost estimation
637 methodology for NETL assessments of power plant performance, U.S. Department of Energy (2011).
- 638 [45] M. R. Abu-Zahra, J. P. Niederer, P. H. Feron, G. F. Versteeg, {CO₂} capture from power plants: Part ii.
639 a parametric study of the economical performance based on mono-ethanolamine, International Journal
640 of Greenhouse Gas Control 1 (2) (2007) 135 – 142, 8th International Conference on Greenhouse Gas
641 Control Technologies GHGT-8. doi:[http://dx.doi.org/10.1016/S1750-5836\(07\)00032-1](http://dx.doi.org/10.1016/S1750-5836(07)00032-1).
642 URL <http://www.sciencedirect.com/science/article/pii/S1750583607000321>
- 643 [46] J. Davison, Performance and costs of power plants with capture and storage of {CO₂}, Energy 32 (7)
644 (2007) 1163 – 1176. doi:<http://dx.doi.org/10.1016/j.energy.2006.07.039>.
645 URL <http://www.sciencedirect.com/science/article/pii/S0360544206002155>
- 646 [47] US Energy Information Administration, Natural Gas, <http://www.eia.gov/naturalgas/weekly/>
647 (2014).