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Comparison of coal IGCC with and without CO_2 capture and storage: Shell gasification with standard vs. partial water quench

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Abstract

This work provides a techno-economic assessment of Shell coal gasification-based IGCC, with and without CO_2 capture and storage (CCS), focusing on the comparison between the standard Shell configuration with dry gas quench and syngas coolers versus partial water quench cooling.

Keywords: coal IGCC, CO_2 capture and storage, CCS, dry feed, Shell gasification, partial water quench, modeling, economics

1. Introduction

In a world with a rapidly expanding appetite for energy and rising concentrations of greenhouse gases, the use of coal as a primary energy source engenders both heightened interest and concern. Coal is the most abundant and least expensive fossil fuel, but also the most carbon intensive. Various gasification technologies enable the conversion of coal into a synthesis gas that can be further processed into common energy carriers such as electricity and synthetic fuels (e.g. hydrogen, natural gas, and liquid transportation fuels). Gasification also provides some of the least costly methods for large scale CO_2 capture for sequestration in deep geologic formations away from the atmosphere.

Numerous studies indicate that bituminous coal-based electric power with CO_2 capture and storage (CCS) is significantly less expensive using integrated gasification combined cycles (IGCC) instead of standard pulverized coal (PC) steam electric plants [1, 2]. For lower rank subbituminous coals and lignites, which comprise fully half of the world's coal reserves [3], the relative economics are much less clear. In recognition of this important issue, this work investigates the thermodynamic and economic performance of three different variants of one particular type of coal IGCC that is likely to be able to economically convert *all* coals into electricity and other energy carriers: pressurized, entrained-flow, oxygen-blown gasification, with coal drying and dry feeding into the gasifier.

Commercial plants of this type (e.g. the Shell Coal Gasification Process) typically employ high temperature heat exchangers to cool down the hot (~900 C) synthesis gas by generating high pressure steam prior to syngas cleaning

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and chemical processing. In plants with CO_2 venting, the high cost of these "syngas coolers" is generally offset by significantly increased plant efficiency. However, costly syngas coolers are often not well matched to CCS, which requires a relatively moist syngas; much of the generated steam must be used for syngas humidification required by the downstream water-gas shift (WGS) reaction necessary for high levels of carbon capture. In this regard, dry feed gasifiers are at a disadvantage relative to coal-water slurry fed gasifiers (e.g. GE and ConocoPhillips E-Gas) which generate a more humid syngas; often, additional steam is not required prior to WGS. To address this issue, Shell recently filed a patent application for a "partial water quench" whereby the hot raw syngas is cooled by direct water injection [4]. This system both humidifies the syngas and eliminates the costly high temperature syngas coolers.

This study compares the thermodynamic and economic performance of Shell IGCC fueled with bituminous coal – with and without CCS – using either the standard or partial water quench syngas cooling methods. Our goal is to understand if partial water quench cooling is the preferred design when capturing CO_2.

2. Methodology

We model three cases: ***SV*** - a standard Shell coal gasifier-based IGCC with syngas coolers and CO_2 venting, ***SC*** - a standard Shell IGCC with CCS, and ***QC*** - the Shell partial water quench IGCC with CCS.[1] This research entailed seven major tasks: 1) building a detailed model of the Shell coal gasification process using Aspen Plus chemical process modeling software [5], 2) calibrating the model by matching key component data and process flows to the detailed information provided in refs. 6 and 7 which describe standard Shell- and Prenflo-based IGCC fueled with

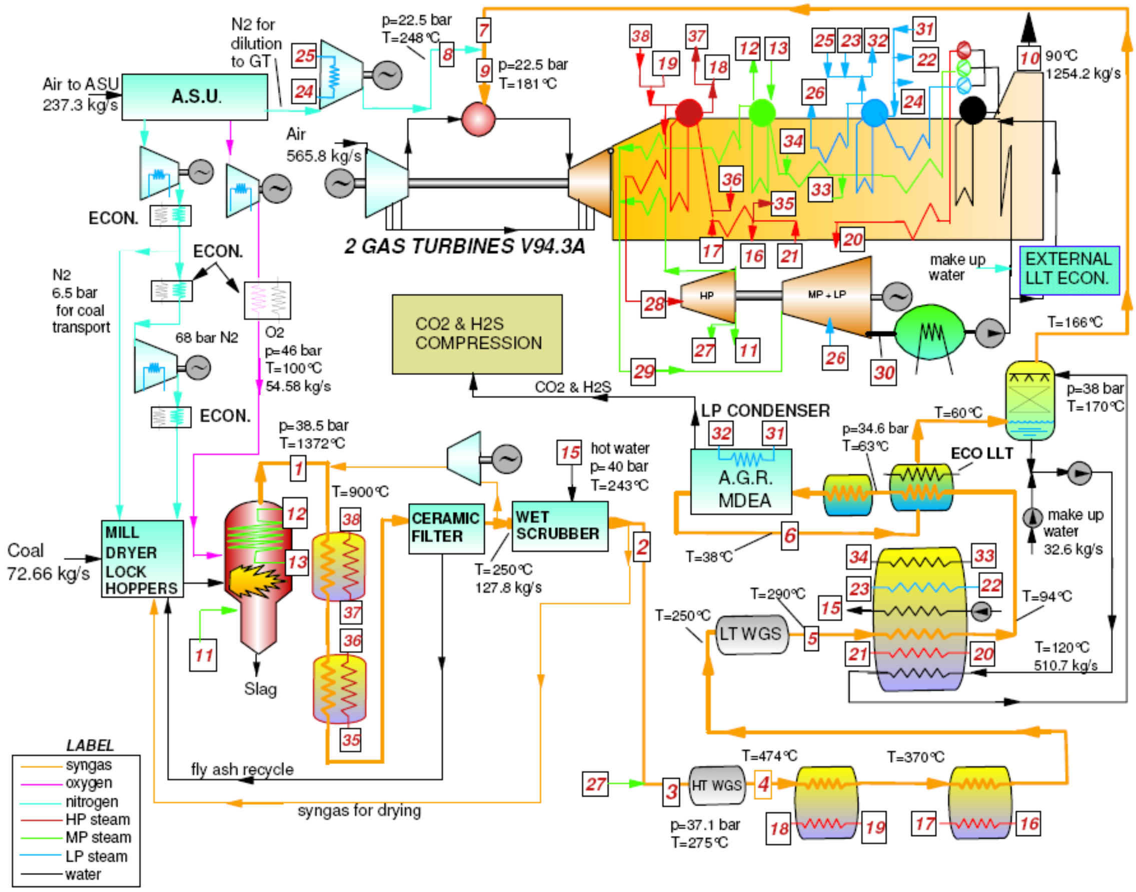

Fig. 1. Plant schematic for case ***SC***, the standard Shell IGCC with CO_2 capture.

[1] Hypothetical case ***QV***, the partial water quench with CO_2 venting, was deemed to be too inefficient to be of general interest.

bituminous coal, 3) tuning/updating the model to match the more recent but less detailed Shell coal IGCCs modeled by Foster-Wheeler (FW) for the IEA Greenhouse Gas R&D Programme [8], 4) investigating the optimal design of a partial water quench/wet scrubber/water gas shift system for Shell IGCC+CCS, 5) simulating the Siemens 94.3A-based combined cycle using the "Gas/Steam" (GS) modeling framework developed at Politecnico di Milano [9,10], 6) optimizing the plant heat integration to maximize the power output of the steam cycle, and 7) adding the cost framework required for a full techno-economic comparison between cases.

C	64.60	Moisture	9.5
H	4.38	Ash	12.2
O	7.05		
N	1.39		*MJ/kg*
S	0.86	HHV	27.063
Cl	0.02	LHV	25.874

Table 1. Composition (%wt) and heating value of as received (AR) East Australian bituminous coal used here [9].

3. System Design Overview

Gasifier Island. The basic IGCC design is illustrated in Fig. 1; calculation details are given in Appendix Table A1. East Australian bituminous coal (Table 1) is milled, dried to a moisture level of 2%wt, and fed into the gasifier via lockhopper pressurization using N_2 as a transport gas. The coal is gasified in the presence of medium pressure (MP) steam and 95% oxygen from a stand-alone cryogenic air separation unit (ASU). Gasification is modeled using full chemical equilibrium at 38.5 bara and 1390 C. Steam and oxidant flows are set by maximizing the LHV of the raw synthesis gas (SG) exiting the gasifier while fixing the heat loss to the membrane wall at 1.4% of the input coal HHV. The single-pass carbon conversion is 97.3%; with recycled fly ash (minus 5% bleed), the overall carbon conversion is 99.8%. Much of the input mineral matter (34.5%) exits the bottom of the gasifier as a vitreous slag; the remainder is captured as fly ash (after syngas cooling) by a ceramic filter and recycled back to the coal milling/drying unit. Heat for this unit is provided by burning 1% of the scrubbed syngas. All gasifier island parameters (Table A1) were "tuned" in order to closely match the detailed data on syngas flow and composition from the gasification island provided by Shell [8].

Case *SV*. In the standard Shell IGCC, the raw syngas exiting the gasifier is first quenched to 900 C (to solidify molten ash) by a stream of recycled, cooled, ash-free syngas and is then cooled to 250 C in syngas coolers that raise both MP and high pressure (HP) steam for the bottoming cycle. Dry particulate filters remove fly ash from the syngas, which is then divided (~45% is sent to the recycle compressor for the gas quench) and sent to a countercurrent flow wet scrubber that removes trace particulate matter and water soluble contaminants. The scrubbed syngas is then warmed to 200 C and passed through a COS hydrolysis unit that converts COS to H_2S, and HCN to NH_3. The syngas is cooled to 40 C and sent to an MDEA-based acid gas removal (AGR) system that strips out virtually all of the H_2S (and 16% of the CO_2) which is sent to an O_2-blown Claus unit for conversion to elemental sulfur. The Claus tailgas is hydrogenated and recycled to the AGR. The sweet syngas exiting the AGR is heated to 181 C and burned in two Siemens V94.3a gas turbines (GT).[2] NO_x emissions are limited to ~25 ppmv (15% O_2) by diluting the syngas with HP N_2 in order to lower the stoichiometric flame temperature to 2300 K [11].[3] Heat is efficiently recovered from the turbine exhaust in a 3 pressure level (166/42/3.5 bar; + reheat and deaerator) heat recovery steam generator (HRSG) coupled to a single steam turbine (0.005 bar condenser). A high degree of heat integration is employed between the syngas train and the steam cycle, and design is optimized to achieve maximum efficiency.

Case *SC*. Our design for the standard Shell IGCC with CCS (Fig. 1) mirrors that of ref. 8 to facilitate model calibration and verification; however, we adopt the somewhat higher steam-to-CO (S/CO) ratio of 2.5 in order to capture ~90% of the total input carbon.[4] In case ***SC***, the scrubbed syngas is preheated, combined with a large flow (423 MW) of superheated MP steam bled from the steam turbine, and sent to a sour water-gas shift (WGS) unit that converts 97% of CO to CO_2 and H_2. Ref. 8 employs a traditional dual-reactor sour WGS design[5] with sulfur-tolerant Co-Mo catalyst and fairly high equilibrium approach temperatures (see Table A1). The syngas enters/exits the high temperature (HT) WGS reactor at 275/475 C; it is then cooled and enters/exits the low temperature (LT) WGS

[2] The assumptions and accuracy of the GT model are described in Ref. 15.

[3] In case ***SV***, the most efficient method of syngas dilution involves using all of the available N_2 and a small amount of steam. In cases ***SC*** and ***QC***, we first saturate the syngas using low temperature heat that is not otherwise well utilized in the bottoming cycle, and then add N_2 as needed.

[4] This relatively high S/CO ratio also prevents carbon formation on the WGS catalyst.

[5] This design is suitable for high S/CO ratios (e.g. 2.5), but less optimal for lower values. Novel WGS designs that include upstream saturators and downstream desaturators, or WGS reactor bypass streams, are not considered here [12].

reactor at 250/290 C. The shifted syngas is cooled, sent to the AGR for H_2S and CO_2 co -capture, saturated with water[2], diluted with N_2, heated to 181 C, and burned in the gas turbines. The FW AGR [8] captures 91.6% of the input CO_2, yielding an overall carbon capture fraction of 87.6%. The captured CO_2+H_2S stream is dehydrated and compressed from 1.8 to 110 bar for pipeline transport and storage in geologic formations.

Case *QC*. In the partial water quench system with CCS (Fig. 2), the syngas recycle loop and syngas coolers of case ***SC*** are replaced by a quench cooling design [4] in which the raw syngas is quenched by a spray of hot (243 C) water, cooling it to a temperature suitable for the downstream particulate filter. The quenched syngas traverses the filter and is sent to the wet scrubber operating with 243 C wash water. As a result of the partial water quench, the syngas enters/exits the scrubber with a S/CO ratio of 1.96/2.2; as a result, the flow of MP steam required to achieve the target S/CO ratio of 2.5 is less than 15% (62 MW) of that in the standard Shell configuration, case ***SC***.

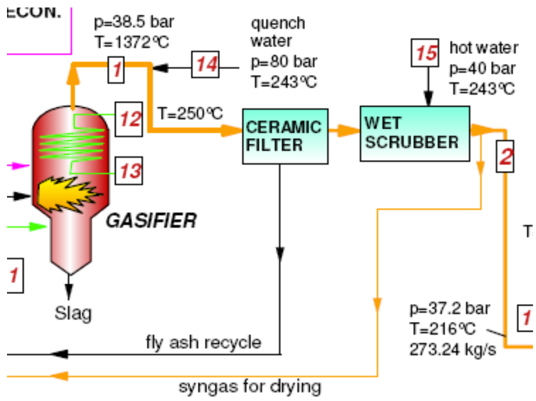

Fig. 2. Partial plant schematic for case ***QC***, the Shell IGCC with partial water quench and CO_2 capture.

4. Partial Water Quench/Scrubber Design

Significant effort was spent optimizing the performance of the partial water quench and wet scrubbing system. Our design goal was to minimize the flow of MP steam to the WGS unit needed to maintain a fixed S/CO ratio of 2.5 in the humidified syngas stream entering the HT-WGS reactor (a value which yields an overall carbon capture fraction of 87.7%). A secondary goal is minimizing water vaporization in the wet scrubber (to minimize cost). Five free parameters must be specified: the temperature and flow rate of the quench water, the temperature of an optional scrubber pre-cooler, and the temperature and flow rate of the wash water in the wet scrubber. To speed the evaporation of the quench water droplets into the syngas, and also minimize exergy loss in the quench process, the quench water temperature is set to 243 C, only a few degrees below the 38.5 bara saturation temperature, T_{sat}=248.1 C.[6] As seen in Fig. 3, as the flow of quench water increases, the temperature of the quenched syngas drops and its S/CO increases. A maximum S/CO value of ~2.1 is reached when the syngas is cooled down to the mixture dewpoint, ~210 C (when using 243 C quench water); this clearly represents an absolute lower temperature limit because of the *dry* particulate filter downstream. However, in order to insure sufficiently rapid evaporation of the quench water droplets, a more plausible lower temperature limit is the saturation temperature of the quench water, T_{sat}. Note that an illustrative quench temperature used in Shell's patent application is 400 C [4].

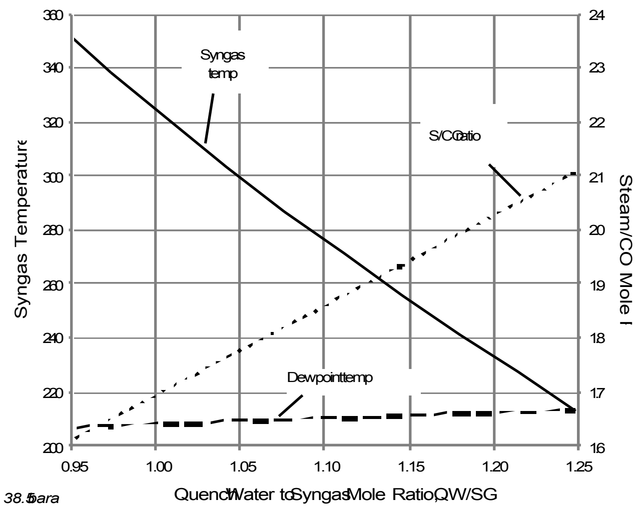

Fig. 3. Temperature and S/CO ratio of quenched syngas as a function of the amount of quench water.

The wet scrubber was modeled as an adiabatic, countercurrent absorption column with 5 equilibrium stages and a fixed liquid-to-gas mass ratio[7] (*L/G*) of 0.25 [13,14]. Thus constrained, the scrubber has only a

[6] The raw syngas could also be quenched in part or totally with steam, but our investigation shows that adding steam to the raw syngas is essentially equivalent to adding it just prior to the WGS unit. In short, adding steam to the quench does not reduce overall steam consumption.

[7] Because a significant fraction of the input water can evaporate into the syngas within the scrubber, we define this quantity as the mass of water *exiting* the scrubber divided by the mass of syngas *entering* it.

limited capability to increase the S/CO ratio of the syngas (Fig. 4). For example, even at the highest wash water temperature of 243 C (just below T_{sat}), the S/CO ratio of dry syngas from a standard Shell gasifier rises from ~0.08 to only ~0.3, requiring the addition of 382 MW of MP steam to achieve the target S/CO ratio of 2.5. In contrast, raw syngas that is partially quenched with water at *QW/SG* = 1.2 can exit the scrubber with S/CO = 2.2, requiring only 53 MW MP steam.

In summary, the partial water quench, scrubber, and WGS steam addition are all methods of humidifying the syngas; minimizing the latter two requires maximizing the partial water quench. This is achieved by using quench water that is as hot as possible (i.e. close to T_{sat}), and quenching down close to T_{sat}. Syngas coolers (such as an optional syngas precooler), which reduce the temperature of the syngas without humidifying it, work *against* these goals. With a fixed *L/G* ratio, the wet scrubber has only a limited ability to alter the humidity of the syngas, and thus using wash water as hot as possible (i.e. close to T_{sat}) is optimal. Comparing the standard Shell vs. partial water quench configurations, the former entails more than three times the amount of heat transfer (440+220 vs. 205 MW).

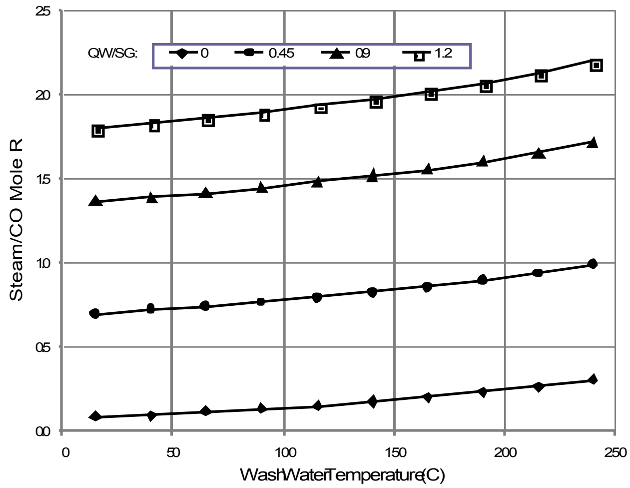

Fig. 4. Steam-to-CO ratio of scrubbed syngas as a function of wash water temperature and quench water-to-syngas mole ratio, *QW/SG*. The scrubber *L/G* ratio is fixed at 0.25.

5. Comparative Plant Performance

The performance of all three cases is given in Tables 2 and 3. The more than 20% drop in LHV efficiency between cases ***SV*** and ***SC*** reflects the well-known losses via WGS (steam consumption and reduction in syngas heating value) and CO_2 compression. (Note that the N_2 compression power for NO_x control is smaller in case ***SC***.) The drop in efficiency from case ***SC*** to ***QC***, only 2.7%, is surprising in light of the significant (39%/54%) reduction in HP/MP steam generated in the syngas coolers (Table 3). The explanation lies in the large steam flow required by the WGS unit; 55% of the HP steam is bled off at 42 bar in case ***SC*** versus only 13% in case ***QC***. This latter case also produces more LT heat and thus more low pressure (LP) steam because it produces more low temperature heat. In summary, case ***SC*** has higher mass flows in the HP section of the steam cycle, and lower in the MP and LP sections.

In more technical terms, since the mechanical power of the steam turbine is equal to the integral of the product $m*\eta*v*dp$ along all the expansion (where m is the mass flow, v the specific volume, dp the infinitesimal pressure drop, and η the polytropic efficiency of the infinitesimal expansion), the difference in HP steam mass flow between cases ***SC*** and ***QC*** (39%) must be weighed together with the terms v and η. Although the mass flow of HP steam in case ***SC*** is higher, the product $v*\eta$ is lower because the specific volume is relatively small (HP section inlet: v=0.023 m^3/kg, MP section inlet: v=0.098 m^3/kg, LP section inlet: v=0.435 m^3/kg), and the polytropic efficiency η is penalized because of its relatively small ratios of blade height to diameter.

Case	*SV*	*SC*	*QC*
Coal input, MW LHV	1,699	1,880	1,880
GT power, MW_e	587.1	580.6	580.6
ST power, MW_e	359.9	293.8	276.8
Coal handling, gasifier	-17.0	-18.8	-18.8
ASU, O_2 & N_2 compr.	-140.8	-113.5	-113.5
AGR	-0.2	-12.0	-12.0
Quench pump	0	0	-1.5
CO_2 dry, compression	0	-35.0	-35.0
Saturator pumps	0	-0.4	-0.4
Auxiliary power, MW_e	*-158.1*	*-179.6*	*-181.1*
Net power, MW_e	789.0	694.8	676.3
LHV efficiency, %	46.44	36.96	35.98
Emissions, g CO_2/kWh	702.8	102.5	105.3

Table 2. Comparative plant performance.

Flow, [kg/s]	*SC*	*QC*
HP level, SH steam produced	270.00	164.10
MP level, SH steam produced	66.40	30.67
LP level, SH steam produced	4.80	13.30
HP section inlet mass flow	270.00	164.10
MP steam for gasifier and WGS	157.27	30.11
MP section inlet/outlet mass flow	179.60	186.60
LP section inlet/outlet mass flow	184.40	199.90

Table 3. Heat Recovery Steam Cycle details.

6. Comparative Plant Economics

Coal price [1]	1.71 $/GJ LHV
Capacity factor	85%
Capital charge rate (CCR)	15% per year
Interest during construction	16.0% of overnight capital
Operation & maintenance	4% of overnight capital / yr
CO_2 transport+storage costs	7.1 $/tonne CO_2
U.S. dollars valued in year	2008 (mid -year)

Table 4. Economic assumptions employed here.[7]

Economic parameters used to estimate the cost of producing electricity are given in Table 4.[8] At these plant sizes, CO_2 removal rates are high (542.3 tonnes/hr in case ***SC***), and so transport and storage (T+S) costs are potentially modest.[9] Our model for estimating the capital cost of each major plant component is derived from the detailed capital cost data for Shell IGCCs given in a May 2007 study by NETL [1]. Costs are escalated to mid-2008 US dollars using the Chemical Engineering Plant Cost Index [17,18]. The total plant cost (TPC), or "overnight construction cost", given in Table 5 for each case, includes engineering and overhead, general facilities, balance of plant, and both process and project contingencies (3.2 and 17% of the bare erected cost, respectively).

						Case SV		*Case SC*		*Case QC*	
Plant component	***Scaling parameter***	***So***	***n***	***f***	***Co (M$)***	***S***	***C (M$)***	***S***	***C (M$)***	***S***	***C (M$)***
Coal and sorbent handling	AR coal, tonne/day	5,447	1	0.67	40.4	5674	41.5	6,278	44.4	6,278	44.4
Coal preparation & feeding	AR coal, tonne/day	2,464	2	0.67	101.6	5674	208.5	6,278	223.1	6,278	223.1
Ash handling	Coal ash, tonne/day	477.8	1	0.67	38.1	692	48.8	765.9	52.2	765.9	52.2
Stand-alone ASU, O_2 compressor	Pure O_2, tonne/day	2,035	2	0.50	106.7	3942	196.0	4,361	206.1	4,361	206.1
Standard gasifier, SG coolers	AR coal, MW LHV	737.4	2	0.67	178.1	1699	365.4	1,880	391.0	-	-
Partial water quench gasifier	AR coal, MW LHV	770.9	2	0.67	139.5	-	-	-	-	1,880	297.4
LT heat recovery, FG saturation	AR coal, MW LHV	737.4	2	0.67	17.3	1699	35.5	1,880	38.0	1,880	38.0
COS hydrolysis	AR coal, MW LHV	797.7	2	0.67	4.7	1699	9.1	-	-	-	-
Water-gas shift reactors	AR coal, MW LHV	815.2	2	0.67	9.3	-	-	1,880	19.1	1,880	19.1
Gas cleanup balance of plant	AR coal, MW LHV	815.2	2	0.67	6.1	1699	11.7	1,880	12.5	1,880	12.5
MDEA AGR (H_2S capture)	S input, tonne/day	23.7	2	0.67	15.9	49	30.3	-	-	-	-
MDEA AGR (H_2S+CO_2 capture)	CO_2 captured, tonne/hr	275.0	2	0.67	43.2	-	-	542.3	79.9	542.3	79.9
Claus plant	S input, tonne/day	136.5	1	0.67	37.6	49	18.9	-	-	-	-
CO_2 compression and drying	Compressor pwr, MW_e	27.4	1	0.67	43.0	-	-	35.0	50.7	35.0	50.7
Siemens 94.3A gas turbine (GT)	-	295.9	2	-	92.8	-	173.2	-	173.2	-	173.2
HRSG, ductwork, & stack	GT net power, MW_e	232.0	2	0.67	33.8	587	73.8	580.6	73.3	580.6	73.3
Steam turbine, condenser, aux.	ST gross power, MW_e	274.7	1	0.67	74.0	360	88.7	293.8	77.4	276.8	74.4
Balance of plant	15.5% of plant cost						237.8		263.3		245.7
Total Plant Cost (TPC)							***1,539***		***1,704***		***1,590***
Specific Total Plant Cost ($/kW$_e$)							**1,951**		**2,453**		**2,351**

Table 5. "Overnight" capital costs for major plant components, and the total plant cost (TPC) for each case. The overnight cost, C, of a component having size, S, is related to the cost, C_o, of a single train of a reference component of size S_o by the relationship: $C = n^e\, C_o\, [S/(n S_o)]^f$, where n is the number of equally sized equipment trains operating at a capacity of 100%/n, f is the cost scaling factor, and e=0.9 is the cost scaling exponent for multiple trains of equipment. (Note: AGR costs from ref. 8.)

Cost of Electricity. The levelized cost of electricity (LCOE) for each plant is given in Table 6 for two prices on CO_2 emissions: zero and 35 $/tonne CO_2, the "crossover" value at which the LCOE for CO_2 capture case ***QC*** equals that of CO_2 venting case ***SV***. Note that the LCOE for case ***QC*** is just slightly (~2.5%) lower than that for case ***SC***, i.e. the small drop in efficiency between cases ***SC*** and ***QC*** (Table 2) is

Cost component, mid 2008 $/MWh	***SV***	***SC***	***QC***	***QC****
Installed capital (at 15% of TPI)	45.6	57.3	54.9	49.2
O&M (at 4% of TPC per yr)	10.5	13.2	12.6	11.3
Coal (at 1.71 $/GJ, HHV)	13.9	17.4	17.9	17.9
CO_2 disposal (at 7.1 $/tonne CO_2)	0.0	5.6	5.7	5.7
LCOE (no carbon price)	*69.9*	*93.5*	*91.2*	*84.1*
CO_2 emissions (at 35.2 $/tonne CO_2)	25.2	3.9	4.0	4.0
LCOE with CO_2 price (35.2 $/tonne)	***95.2***	***97.4***	***95.2***	***88.1***

Table 6. Levelized cost of electricity for each case.

[8] Interest during construction (IDC) is based on a 4-year construction schedule with equal, annual payments, and a discount rate of 10%/yr. The capital charge rate is applied to total plant cost (TCP) + IDC.

[9] CO_2 T+S costs are based on a 100 km pipeline, aquifer depth of 2 km, CO_2 injectivity of 2500 tonne/day per well, and a 19%/yr CCR. [16]

more than offset by the reduction in capital cost (Table 5) associated with the syngas coolers. We also anticipate a higher availability for the partial water quench case due to lack of syngas cooler fouling, leakage, creep, and metal dusting. In short, this analysis suggests that, from the standpoint of overall economics of Shell coal IGCC with CCS, the standard and partial water quench configurations are quite similar.

In Table 6 we have included an additional (economic) case ***QC**** that uses an alternative gasifier capital cost model based on a highly disaggregated vender quote for the ATI Sulcis Shell IGCC which suggests that the syngas coolers are twice as costly as the gasifier [19]. This implies a relatively high steam generation cost of ~1000 $\$/kW_{th}$ (compared to ~400 $\$/kW_{th}$ in Table 5), comparable to the cost of convective syngas coolers used by the GE coal gasifier [20,14]. If case ***QC**** is a more accurate reflection of Shell syngas cooler costs, then the partial water quench may be significantly more competitive than the standard Shell IGCC with CCS; note that the specific TCP of case ***QC**** is $2,351 $\$/kW_e$ and its LCOE is 9-10% less than case ***SC***. This result mirrors the conclusion of a previous analysis of GE-based coal IGCC (with and without CCS), in which the total water quench design was found to be less efficient but economically superior to plants with radiant+convective syngas coolers [15,20].

Finally, we note that the relative advantage of the partial water quench over the standard Shell configuration is likely to be smaller at lower S/CO ratios, as might be found in systems with a lower overall carbon capture fraction, a more advanced WGS design, and/or an AGR unit with a higher CO_2 capture fraction.

7. Conclusion

In a conventional Shell coal IGCC with syngas coolers, adding CO_2 capture reduces plant efficiency by nearly 10 percentage points. The partial water quench (with CCS) further decreases the efficiency by ~1 percentage point. The cost of Shell coal IGCC with CCS is estimated in the range 2300-2500 $\$/kW_e$, with the partial water quench near the lower end and the conventional design at the upper end. With the partial water quench, the levelized cost of "decarbonized" electricity is likely to be equal or less than – perhaps by as much as 10% – the standard Shell configuration with syngas coolers.

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Appendix

SHELL GASIFICATION ISLAND		***WATER GAS SHIFT (WGS) UNIT***	
Dried coal moisture content (wt%)	2	HT/LT WGS reactor approach temperatures (°C)	82.5/22.0
Syngas for drying (% of total flow)	1.0	HT/LT WGS syngas input temperatures (°C)	275/250
Gasification pressure (bara)	38.5	***CO_2 COMPRESSION AND DRYING***	
H_2O/O_2 molar ratio (steam as moderator)	0.266	Specific electricity use: P_{elec}/CO_2 mass flow (kJ/kg)	231.23
Gasification temperature	1372	***AIR SEPARATION UNIT (ASU)***	
Carbon conversion (with fly ash recycle)	99.79	Air compressor (axial), polytropic efficiency	0.86
O_2 purity (% molar fraction)	95	Pressure of O_2 and N_2 delivered by ASU (bara)	1.05
HP N_2 for pressurization /coal mass flow (wt ratio)	0.3193	Excess air	0.06
LP N_2 for coal transport /coal mass flow (wt ratio)	0.1244	Compressor electrical*mechanical efficiency	0.92
HP N_2 into syngas/dried coal (wt ratio)	0.103	O_2 compressor (3 ICs), 4 radial stages, average η_{poly}	0.845
Maximum oxidant temperature (°C)	100	LP N_2 compressor (1 IC), 2 radial stages, average η_{poly}	0.858
Syngas coolers: pinch points gas-steam (°C)	20	HP N_2 compressor (2 ICs), 3 radial stages, average η_{poly}	0.79
Steam/CO value in the WGS reactor	2.5	Dilution N_2 compressor (axial machine,1 IC), avg. η_{poly}	0.887
Heat loss from heat exchangers (%)	0.5	Intercooler exit temperature (°C)	45
Heat to membrane wall (% coal LHV thermal power)	1.50	***POWER ISLAND***	
Syngas coolers & wet scrubber pressure drop (%)	4.00	(2 Siemens V94.3A GTs + 2 HRSGs + 1 steam turbine)	
Elec. use: coal handling+water sys. (% of coal LHV)	1.00	HRSC, 3 pressure levels: condenser 0.005 - 3.5 - 42 - 166	
SYNGAS TREATMENT & CONDITIONING LINE		Steam temperature at admission SH HP and RH (°C)	565
Heat exchangers pressure drops - gas side (%)	0.05	Steam temperature at admission SH LP (°C)	230
Heat exchanger heat losses (%)	0.5	Delta T pinch points and sub cooling in HRSG (°C)	10
Quench water pump hydraulic efficiency	0.8	Pumps: hydraulic efficiency	0.84
Quench water pump elec. * mechanical efficiency	0.9	Pumps: electrical * organic efficiency	0.9
Saturator pump hydraulic efficiency	0.75	Steam turbine, HP section (166-42 bar), η_{iso}	0.845
Saturator pump electrical * mechanical efficiency	0.9	Steam turbine, LP section (42-2 bar), η_{iso}	0.92
ACID GAS REMOVAL (AGR) UNIT		Steam turbine, LLPP section (2-0.005 bar), η_{iso}	0.84
(UOP/DOW - Amine Guard MDEA, 26 barg)		Steam turbine mechanical efficiency	0.98
Gas temperature at AGR inlet (°C)	38	Steam turbine generator electrical efficiency	0.99
LP steam for stripping (MW per kg/s of stripped CO_2)	0.538	Economizers: pressure losses (%)	16
Specific electricity use: P_{elec} /SG mass flow (kJ/ kg)	79.219	Superheater and reheater pressure losses (%)	8
CO co-absorbed/ CO_2 mass flow (%)	0.011	Electricity for condenser cooling (% of condens. heat)	0.5
H_2 co-absorbed/ CO_2 mass flow (%)	0.038		

Table A1. Technical assumptions used in calculations of plant performance