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SHELL GASIFIER-BASED COAL IGCC WITH CO_2 CAPTURE: PARTIAL WATER QUENCH VS. NOVEL WATER-GAS SHIFT

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ABSTRACT

We provide a detailed thermodynamic and economic analysis of two novel alternatives to the "standard" Shell coal IGCC with CO_2 capture. In the first, the syngas coolers are replaced by a "partial water quench" where the raw syngas stream is cooled and humidified via direct injection of hot water. This design is less costly, but also less efficient. The second approach retains syngas coolers but instead employs a novel WGS configuration, developed at the Energy Research Centre of the Netherlands (ECN), that requires substantially less steam to obtain the same degree of CO conversion to CO_2, and thus increases the overall plant efficiency. We investigate how both of these innovations alter the plant's cost and complexity, and ultimately, the levelized cost of electricity.

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 is less costly 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 less clear. To help clarify this issue, we investigate the thermodynamic and economic performance of three different variants of one particular type of coal-based IGCC plants 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. All plants in this work use bituminous coal; a forthcoming study addresses the effect of coal rank on plant performance and economics.

Commercial plants of this type (e.g. that use the Shell Coal Gasification or Siemens Fuel 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 and chemical processing. In plants with CO_2 venting, the high cost of these "syngas coolers" (SC) is generally offset by significantly increased plant efficiency. However, costly syngas coolers are often not well matched to CO_2 capture, 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 CO_2 capture. In this regard, dry feed gasifiers are at a disadvantage relative to coal-water slurry fed gasifiers (e.g. GE Energy and ConocoPhillips E-Gas^{TM}) 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 SCs.

Researchers at ECN have recently developed an advanced WGS design that significantly reduces the flow of steam required for conversion of CO and H_2O to CO_2 and H_2 [5]. This system has recently been implemented at pilot scale at NUON's Buggenum IGCC plant in the Netherlands.

This study compares the thermodynamic and economic performance of a bituminous coal-based IGCC plant using Shell gasification technology – with and without CO_2 capture – using either the standard gas quench or the partial water quench as syngas cooling method and either the conventional two-stage sour WGS or the advanced ECN WGS design. Our goal is to understand what the preferred IGCC design is for dry feed, entrained flow gasifiers with CO_2 capture.

METHODOLOGY

We model four cases, three with CO_2 capture:

SV - a ***S***tandard (i.e. with syngas coolers) Shell coal gasifier-based IGCC plant with syngas coolers and CO_2 ***V***enting,

SC – a ***S***tandard Shell IGCC plant with CO_2 capture that uses a ***C***onventional two-stage WGS unit,

SE **–** a ***S***tandard Shell IGCC plant with CO_2 capture that uses the advanced ***E***CN WGS design, and

QC – a partial water ***Q***uench Shell IGCC plant with CO_2 capture, using a ***C***onventional two-stage WGS unit.

This research entailed seven primary tasks: 1) building a detailed model of the Shell coal gasification process using Aspen Plus chemical process modeling software [6], 2) calibrating the model by matching key component data and process flows to the detailed information provided in refs. 7-9 which describe standard Shell- and Prenflo-based IGCC plants using bituminous coal, 3) investigating the optimal design of a partial water quench + wet scrubber + WGS system for Shell IGCC with CO_2 capture, 4) building the ECN WGS and coupling it to a standard Shell IGCC plant, 5) simulating the General Electric (GE) 9FB gas turbine (burning H_2-rich syngas) using the "Gas/Steam" (GS) simulation code developed at Politecnico di Milano [10,11], 6) configuring and optimizing the layout of the heat recovery steam cycle (HRSC) for each plant using a new method developed by Martelli [12] that maximizes the power output of the steam cycle, and 7) adding the cost framework required for a full techno-economic comparison between cases.

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 to 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

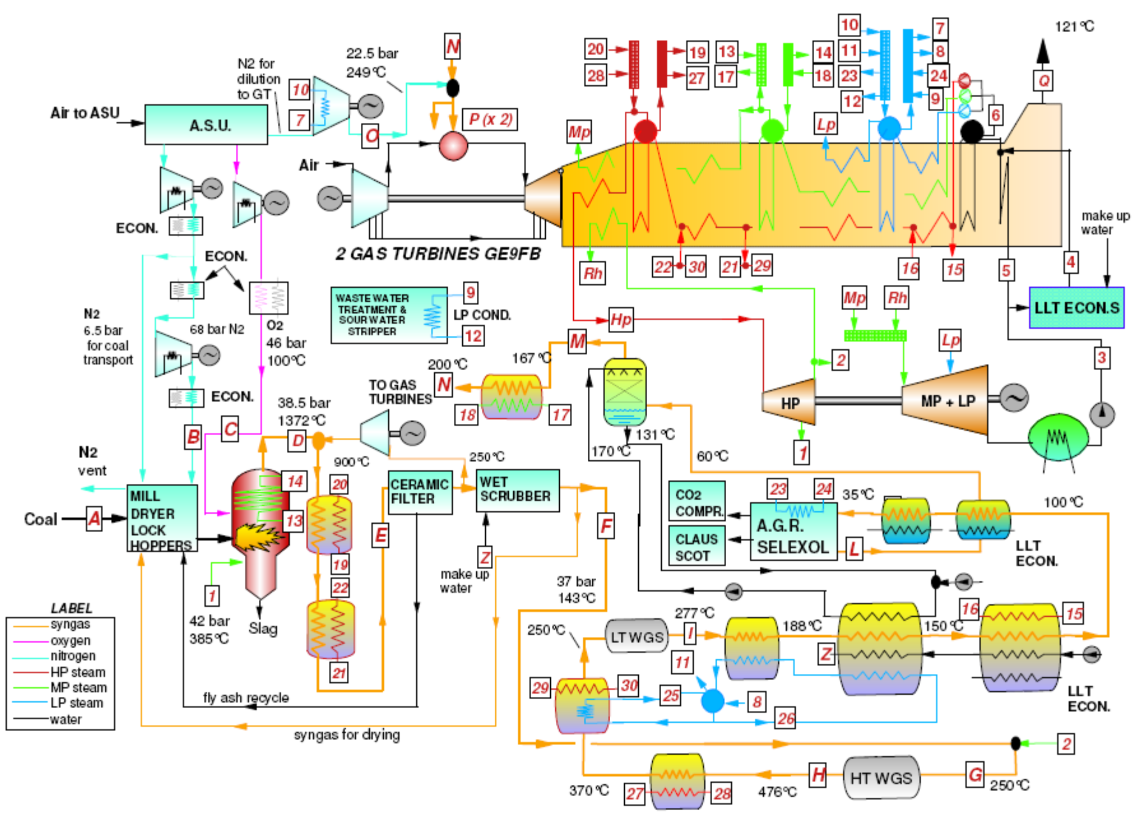

Fig. 1. Plant schematic for case ***SC***, the standard Shell IGCC+CO_2 capture with a conventional WGS.

and recycled back to the coal milling/drying unit. Heat for drying 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 [9].

Case *SV*. In the standard Shell IGCC plant, 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 economize and evaporate high pressure (HP) feedwater to generate 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. 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 a Sulfinol M™ 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 350°C and burned in two GE 9FB gas turbines (GT). NO_x emissions from the GT are limited to ~25 ppmv (15% O_2) by diluting the syngas with all the available N_2 and some steam in order to lower the stoichiometric flame temperature to 2027°C [14].[1] Heat is efficiently recovered from the turbine exhaust in a 3 pressure level (plus reheat) heat recovery steam generator (HRSG) coupled to a single steam turbine. A high degree of heat integration is employed between the syngas train and the steam cycle, and design is optimized to achieve maximum efficiency [12].

Case *SC*. Our design for the standard Shell IGCC with CO_2 capture (Fig. 1) mirrors that of ref. 8 to facilitate model calibration and verification; however, we adopt a somewhat higher minimum input steam-to-CO (S/CO) ratio of 2.5 in order to avoid carbon formation on the WGS catalysis [13] and also to achieve an “overall carbon capture fraction”[2] of 93.1%. The scrubbed syngas is preheated, combined with a large flow of superheated MP steam bled from the steam turbine, and sent to a conventional two-stage sour WGS unit (with sulfur-tolerant Co-Mo catalysts) that converts 98% of CO to CO_2 and H_2. The syngas enters/exits the high temperature (HT) WGS reactor at 250/ 466°C; it is then cooled and enters/exits the low temperature (LT) WGS reactor at 250/276°C. The shifted syngas is cooled to 38°C, sent to the Selexol™ process for H_2S and CO_2 selective removal, saturated with water, diluted with N_2, heated to 200°C, and burned in the gas turbines. The Selexol AGR [9] captures 96.54% of the inlet CO_2, 0.53% of H_2 and 0.44% of CO. Thus, the overall carbon capture fraction is 93.0% for the cases with 98% CO conversion and 90.4% for the cases with 95% CO conversion. The captured CO_2 stream is dehydrated and compressed from 1.8 to 150 bar for pipeline transport and storage in geologic formations; the H_2S rich stream is treated in a conventional Claus unit followed by a Shell Claus Offgas Treating (SCOT) process. In the cases with only 95% CO conversion, 19.3% of the syngas is split off from the primary flow, bypasses the HT-WGS reactor (still operating at S/CO 2.5), and is fed directly into the LT-WGS reactor[3]. As a consequence, the second reactor has a higher reaction heat and outlet temperature (321 C), suitable for raising both MP and LP steam.

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 East Australian bituminous coal used here [9].

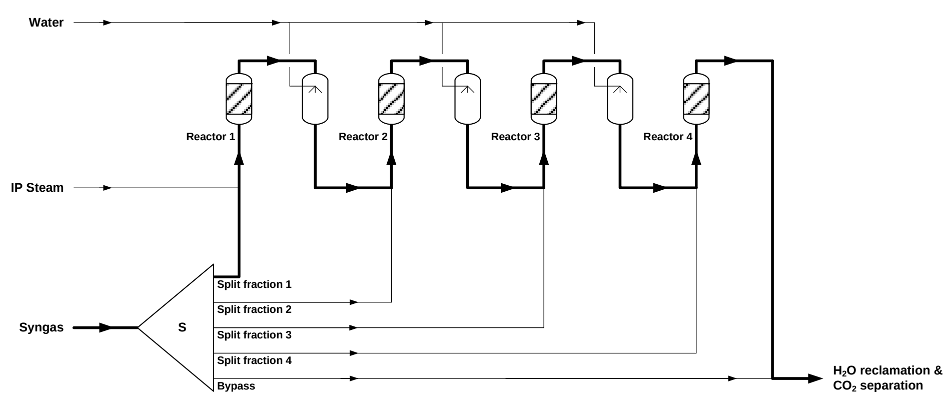

Fig. 2. Layout of the ECN advanced WGS.

Case *SE*. In contrast to the two-stage conventional WGS unit,[4] the advanced ECN system [5], as implemented here (Fig. 2), consists of four sequential, adiabatic sour shift reactors, each of which is fed a fraction of the original syngas stream plus the requisite amount of either MP steam (reactor 1) or 160°C water (reactors 2-4) needed to satisfy the requirement

[1] 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 all CO_2 capture cases 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 to control NO_x.

[2] Defined here as the fraction of the carbon in the input coal that is retained either as carbon in the gasifier slag/flyslag or as CO_2 stored underground.

[3] Bypassing both WGS reactors would require an additional COS hydrolysis reactor for adequate sulfur capture in the AGR.

[4] Novel WGS designs that include upstream saturators and downstream desaturators [13] are not considered in this work.

adopted here that S/CO=2.5 at the input of each reactor (in order to avoid carbon formation on the catalyst). Designed specifically to minimize the amount of steam required for the WGS reaction, the ECN system utilizes the concept that, since the WGS reaction consumes only one mole of H_2O per CO, the H_2O/CO ratio within the bed (along the reactant flow path) rapidly exceeds 2.5 as "excess" water accumulates. By splitting the original syngas into multiple fractions and combining the water-laden streams exiting upstream WGS reactors with the feed streams of downstream WGS reactors, the total water consumption is significantly reduced. As discussed below, this has important consequences on both the efficiency and economics of the system.

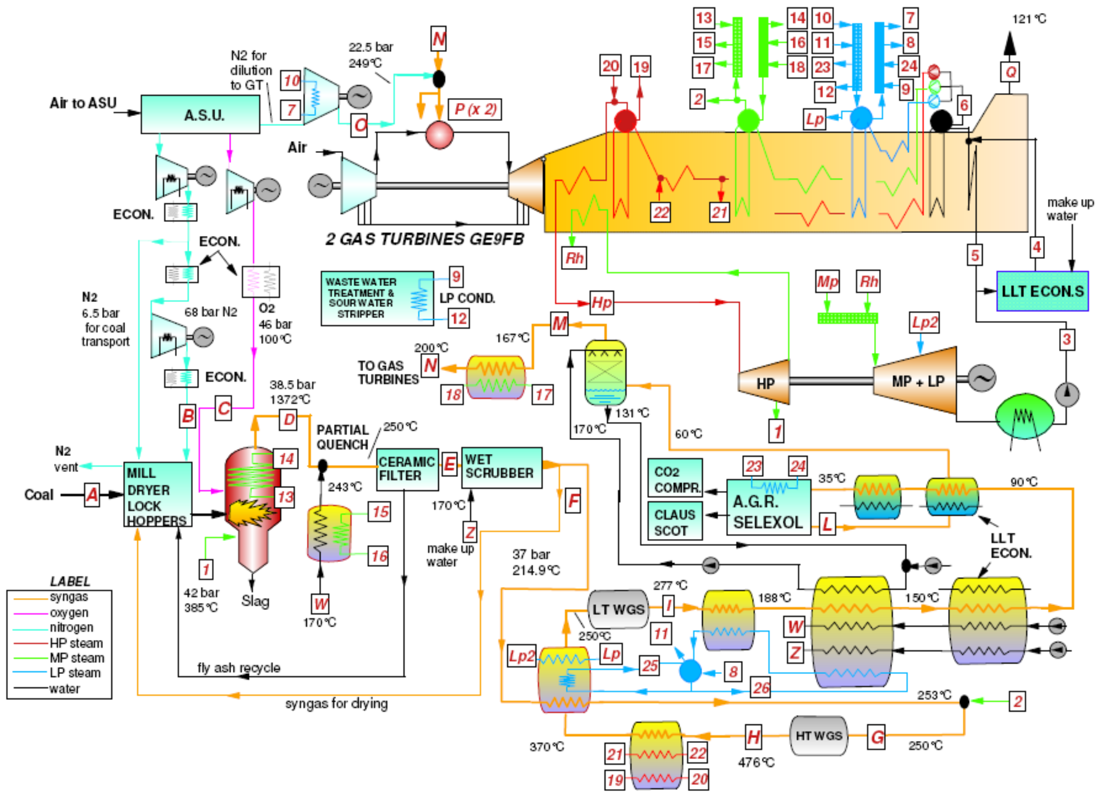

Fig. 3. Plant schematic for case ***QC***, the partial water quench IGCC with the conventional WGS.

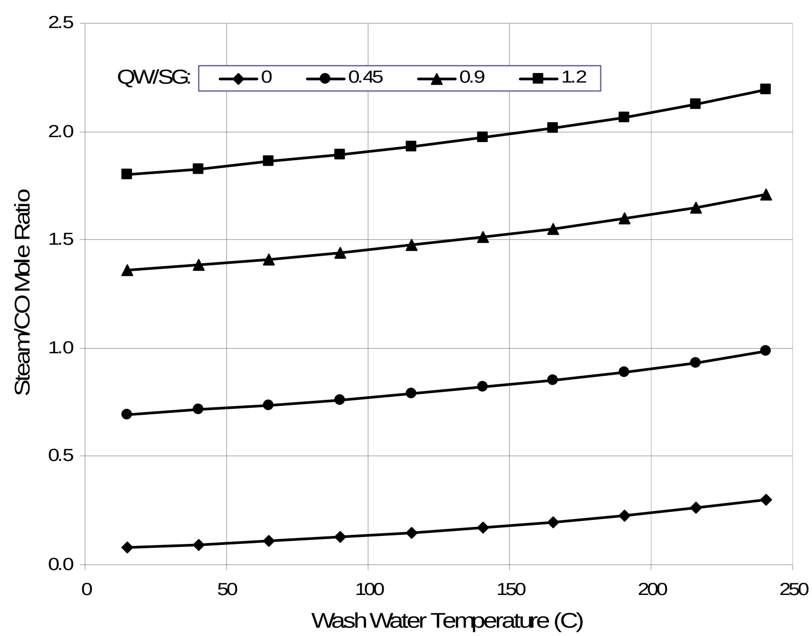

Fig. 4. Steam-to-CO ratio of scrubbed syngas as a function of wash water temperature and quench water-to-syngas mole ratio, *QW/SG*. The *L/G* ratio of the scrubber is fixed at 0.25.

Configuring the advanced WGS system shown in Fig. 2 is relatively straightforward, with an equal number of free parameters and constraints. Making the simplifying assumption that all water input streams have the same temperature, there are ten free parameters to be set: four syngas split fractions, four steam/water flows, and the water and steam temperatures. The condition of S/CO=2.5, and T=250°C at the input of each reactors yields eight constraints; the remaining two are the temperature of MP steam from the steam turbine and the overall CO conversion efficiency. The parameter values for case ***SC*** are given in Table 1a. The plant layout for case ***SE*** is essentially identical to that shown in Fig. 1, with conventional WGS unit replaced by the ECN system shown in Fig. 2.

Case *QC*. In the partial water quench system with CO_2 capture (Fig. 3), the standard syngas cooling system (gas quench via syngas recycle, followed by syngas coolers) of case ***SC*** is replaced by a water 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 170°C wash water. The syngas enters/exits the scrubber with a S/CO ratio of 1.96/2.2, and as a result, the flow of MP steam required to achieve the target S/CO ratio of 2.5 is less than 15% of that required by the standard Shell configuration.

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 achieve a S/CO ratio of 2.5 in the humidified syngas stream entering the HT-WGS reactor while limiting water vaporization in the wet scrubber (to minimize costs). 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 was set to 243°C, only a few degrees below the 38.5 bara saturation temperature, T_{sat}=248.1°C.[5]

The wet scrubber was modeled as an adiabatic, countercurrent absorption column with 5 equilibrium stages and a fixed liquid-to-gas mass ratio[6] (*L/G*) of 0.25 [15,16]. Thus constrained, the scrubber has only a 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_{th} of MP steam to achieve the target S/CO ratio of 2.5. In contrast, raw syngas that is partially quenched with water at quench water-to-syngas mass ratio (*QW/SG*) of 1.2 can exit the scrubber with S/CO = 2.2, requiring only 53 MW_{th} 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. However, the gas quench process with syngas coolers and steam injection for humidification is much less irreversible than the direct water quench: the high temperature heat of the raw syngas can produce high pressure steam that expands in the first section of the steam turbine up to the bleed for the WGS; in the water quench process the high temperature raw syngas is immediately degraded into low temperature heat which may produce only low pressure steam. Comparing the standard Shell (***SC***) vs. partial water quench configuration (***QC***), the former entails more than three times the amount of heat transfer (440+220 vs. 205 MW_{th}).

With a fixed *L/G* ratio, the wet scrubber has only a limited ability to alter the humidity of the syngas. Thus, to limit the water vaporization and to optimize the recovery of low temperature heat in the cooling section, the temperature of the wash water is fixed slightly below the syngas dew point (170°C for cases ***SC*** and ***QC***, and 160°C for case ***SE***).

Table 2. Details of the GE 9FB gas turbine model used here.

Fuel / Case	*GE 9FB*	*Calibrated Gas Turbine Model (using "Gas-Steam" software [10,11])*			
	Natural Gas	***Natural Gas***	***SV***	***SE***	***SC, QC***
TIT (estimated), °C	unknown	1,360[a]	1,327[b]	1,260[b]	1,260[b]
Turbine outlet (*TOT*), °C	622.8	622.7	613.3	565.1	566.4
Air mass flow, kg/s	637.13	637.13	530.45	573.5	575.43
Fuel mass flow, kg/s	16.187	16.198	132.73	69.212	65.432
1st row cooling, kg/s	unknown	61.022	55.507	46.262	46.516
Flue gas, kg/s	653.317	653.33	663.26	642.72	640.86
Net electric power, MW	272.6	272.44	313.8	296.96	296.46

[a] Estimate of the real *TIT*, calibrated to reproduce the GT output.
[b] *TIT* de-rated using Eqn. 1.

POWER ISLAND DESIGN AND MODELING

The power unit consists of two GE 9FB gas turbines, each with a three pressure level HRSG integrated with the process (syngas coolers, Selexol reboiler, and sour water stripper.), and a single steam turbine. While the GE 7FB is commercialized for operation on syngas, the 9FB has not yet been so proven. Nevertheless, since the 9FB is a scaled-up version of the 7FB, it is expected to run on syngas without significant issues. The GT has been modelled using the "Gas/Steam" simulation code [10,11] and the performance estimation method of Chiesa [17]. The GT model was calibrated to reproduce the same output of a GE 9FB burning natural gas (NG), and then adjusted for operating on syngas (see Table 2). We adopt the control strategy discussed in ref. 14 for operating a syngas-fired GT without major modifications: the turbine inlet temperature (TIT) is fixed at a value consistent with blade lifetime equal to that for NG operation (TIT de-rating discussed below), and the variable inlet guide vanes (VGV) are partially closed to reduce the compressor air flow. We assume here a constant pressure ratio - that employed for NG; the actual pressure ratio will depend on the actual compressor map (pressure ratio versus non-dimensional mass flow), details of which are confidential and not readily predicted. In general, variations in pressure ratio can be neglected if the compressor air mass flow exceeds ~85% of its design point (NG) value.

[5] 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.

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

We de-rate the *TIT* to offset the relatively high heat transfer between the H_2O-rich gases and the turbine blades, and thus insure standard blade lifetimes. Rather than using the method of ref. 14 which adjusts the TIT until critical blade temperatures obtained during syngas operation match those during NG firing (using the same cooling flows), we employ instead the more conservative correlation of Oluyede [18]:

$$\Delta T_{TIT}(°F) = 13.312*(\text{Vol. \% } H_2)^{0.69} \quad (1)$$

which, for our shifted syngas, de-rates the TIT by 100°C (compared to 35°C using the method of ref. 14).

The GT fuel inlet temperature is another key parameter. High fuel inlet temperatures improve overall cycle efficiency because the thermal energy of pre-heating replaces the chemical energy of syngas. Standard CO-rich syngas can be heated up to 350°C without major issues [19], but H_2-rich syngas can be safely pre-heated to only 200-210°C, just enough to avoid water condensation during gas expansion through the combustor control valve. For these reasons, we adopt an inlet fuel temperature of 350°C in case ***SV***, and 200°C in the CO_2 capture cases.

HRSC and Heat Integration. Downstream of each gas turbine is a three pressure level HRSG with reheat. The heat integration of each plant was individually optimized using a new methodology developed by Martelli [12] in which the arrangement of the steam network, pressure levels, and pinch point ΔTs were chosen to maximize the net electric power of the HRSC. Capital costs are not explicitly considered in the objective function, but economic feasibility is obtained by adopting realistic design constraints. For example, all syngas coolers are evaporators or economizers and have a minimum pinch point temperature difference, ΔT_{pp}, of 15°C, and 50°C for the HT SC. Within the HRSG, $\Delta T_{pp} \geq 8$°C, sub-cooling ΔT = 3°C, and the superheat/reheat approach $\Delta T \geq 25$°C. The condenser pressure is assumed to be 0.05 bar (e.g. cold river or sea water). We generally employ a "conservative" set of HRSC parameters in which $P_{HP} \leq 140$ bara, $P_{MP} > 39$ bara, $P_{LP} \geq 5$ bara, and superheat temperature limit, T_{SH}, is 540°C. Detail results of each optimized HRSC are given in Table 3. We also briefly discuss the results of an "aggressive" HRSC parameter set in which $P_{HP} \leq 170$ bara, P_{MP} is free, $P_{LP} \geq 3$ bara, and T_{SH} = 565°C (or TOT - 20°C).

PLANT PERFORMANCE

Plant performance is given in Tables 3 and 4.

SV: CO_2 Venting Reference Case. The efficiency of the plant without capture, ~48% LHV, is higher than a similar case presented previously [20] for three reasons: the GE 9FB is more efficient than the Siemens V94.3A[7], the clean syngas is highly pre-heated before entering the combustor, and the heat recovery steam cycle is fully optimized (pressure levels, steam/gas temperatures and syngas cooler integration). Our results predict that such high efficiencies are possible using well optimized, commercial technology.

[7] The pressure ratio is higher: 18.3 vs. 17.0, the *TIT* is estimated to be 10°C higher, and the clean syngas is highly pre-heated before entering the combustor.

Table 3. Details of the optimized HRSCc for each case. (All steam flows are given in kg/s.)

	SV	***SC***	***SE***	***QC***
HP-MP-LP pressures, bara	140-39-5	140-39-5	140-39-5	140-42-5
HP steam produced	248.27	271.24	237.22	161.64
MP steam produced	25.99	60.48	38.19	0.00
LP steam produced	7.79	35.74	15.63	55.66
HP steam at turbine inlet	248.27	271.24	237.22	161.64
MP steam into gasifier	7.24	7.73	7.71	7.73
MP steam bleed into WGS	0.00	153.66	71.25	28.97
MP steam bleed into GT	21.39	0.00	0.00	0.00
MP steam at turbine inlet	243.17	171.31	195.75	152.31
LP steam at turbine inlet	250.97	184.62	211.39	207.98
Net electric power, MW	359.84	281.08	300.72	254.34

Table 4. Plant performance, with breakdown of power consumption by unit.

	SV	***SC***	***SE***	***QC***
Power Consumption, MW_e:				
Coal handling, gasifier	17.30	18.47	18.42	18.47
ASU, O_2 & N_2 compr., pumps	144.55	123.03	128.69	123.65
AGR, Claus, SCOT units	1.04	19.69	19.75	19.69
CO_2 drying & compression	0.00	35.08	35.18	35.08
Auxiliary power, MW_e	*162.89*	*196.27*	*202.04*	*196.89*
GT net power (2 × 9FB), MW_e	627.60	592.92	593.92	592.92
ST net power, MW_e	359.76	279.57	300.58	253.91
Net electric power, MW_e	*824.47*	*676.21*	*692.46*	*649.94*
Coal input, MW_{th} LHV	1,729.7	1,846.7	1,841.6	1,846.7
LHV efficiency, %	47.66	36.62	37.60	35.19
Overall carbon capture, %	0.00	93.09	93.09	93.09
CO_2 emissions, g CO_2/kWh	667.75	52.01	51.70	54.81

SC, SE, QC: CO_2 Capture Cases. These plants share the same 98% CO conversion efficiency and AGR, and thus achieve the same overall degree of carbon capture, 93.1%. The difference between them is the syngas humidification process and the WGS layout.

SV vs. SC: CO_2 Capture Penalty. The more than 23% 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***.) A primary goal of this work is to reduce this efficiency loss by novel syngas humidification schemes.

SC* vs. *QC*: Partial Water Quench.** As seen in Table 3, the partial water quench amply fulfills its design goal of reducing the steam needed for syngas shifting; the WGS unit in ***QC consumes less than 1/5 of the MP steam required by ***SC***.

However, the loss of syngas coolers exacts a thermodynamic toll; the overall efficiency of case ***QC*** is 3.9% (or 1.4 percentage points, pp) less than ***SC***. The reason is that quenching ~1400°C syngas with hot water is the most irreversible method (in the plants studied here) of increasing syngas humidification, and consequently case ***QC*** has the lowest efficiency of all cases, 35.2%. In contrast, case ***SC*** employs the least irreversible method, steam injection.

At first, the 3.9% drop in efficiency from case ***SC*** to ***QC*** seems surprisingly small in light of the 40% reduction in HP steam available to the steam turbine (Table 3). The reason is that the mass flow of steam is only one of three important factors governing the steam turbine's mechanical power, equal to the integral of $m*\eta*v*dp$ along the expansion (where m is the mass flow, v is the specific volume, η is the polytropic efficiency, and dp the infinitesimal pressure drop). Despite the higher mass flow of HP steam in case ***SC***, the product $v*\eta$ is lower because v is relatively small (HP section inlet: v=0.0243 m^3/kg, MP section inlet: v=0.0937 m^3/kg, LP section inlet: v=0.522 m^3/kg) and the polytropic efficiency η is lower because of the smaller ratio of blade height to diameter. Thus, the fact that the quench plant expands 12.6% *more* low pressure (LP) steam than in case ***SC*** (and only 11% less MP steam) goes a long way toward mitigating the effects of reduced HP steam.

SC* vs. *SE*: *ECN Advanced WGS*.** The ECN design reduces the WGS steam requirement by 54% (vs. 81% in ***QC). By humidifying syngas with both steam and water, case ***SE*** would seem to be – exergetically – an intermediate between ***SC*** and ***QC.*** However, its 37.6% overall efficiency is the highest of all three 98% CO conversion cases, 2.7% (1 pp) higher than the conventional WGS case ***SC***. The comparison between the steam cycles in ***SE*** and ***SC*** mirrors the previous discussion of ***QC*** vs. ***SC***. Although ***SE*** generates 12.5% less HP steam than ***SC***, it expands 14% *more* MP steam and 14.5% *more* LP steam. Not surprisingly, its steam turbine produces 20 MW_e (7.5%) more power. This discrepancy is the most important factor underlying the difference in overall plant efficiency between the two cases.[8]

To investigate in detail why the bottoming cycle of ***SE*** generates more power than that of ***SC***, we use a comparative exergy analysis (Table 5) on all units related to the HRSC: HT syngas coolers, WGS unit, LT syngas coolers (downstream of the WGS) and HRSC. The primary differences between the two cases lie in the WGS unit and LT syngas cooling; the remaining units (the syngas coolers and HRSGs) are virtually identical. Since the exergy input (raw syngas) and output (cold syngas at 140°C leaving the cooling section) are the same for the two cases, the difference of the total reversible power lost must be equal to difference of the HRSC electric power. Table 5 indicates that the process of mixing syngas with water in the ECN WGS is ~13 MW more dissipative than mixing syngas with steam in the conventional WGS. However, because the ECN arrangement requires much less water overall, that loss is more than repaid in the LT cooling section where less latent heat of condensing water is wasted (exergetically).[9] This simplified exergy analysis accounts almost exactly for the difference in HRSC power between the two cases.

Table 5. Comparison of exergy losses (MW) within the WGS unit, syngas cooling, and steam turbine. The difference in the total exergy loss equals the difference in the net power of the two steam cycles.

Plant Section	***Process***	***SC***	***SE***	***Diff.***
WGS Unit	Mix syngas + water	23.12	35.85	-12.74
	WGS reaction	32.90	26.64	6.26
	Heat transfer	2.20	0.00	2.20
	Total loss	*58.21*	*62.50*	*-4.28*
Cooling post WGS	Heat transfer	16.69	5.37	11.32
	Cooling: 140°C to 38°C	37.47	23.10	14.37
Overall	***Total exergy loss***	112.37	90.97	21.41

If we employ the "aggressive" set of HRSC parameters rather than the default "conservative" assumptions, the gain in plant efficiency offered by case ***SE*** is markedly reduced, from 2.7% to only 1% (0.37 pp) because the improved bottoming cycle benefits case ***SC*** more than ***SE***. Case ***SC*** makes more LT heat because it adds more H_2O to the syngas, and efficient recovery of this heat is critical to high overall efficiency. The LP pressure level is particularly important [10] in this regard; at P_{LP} = 3.7 bara, the exergy loss during LT cooling drops from 17 to 9 MW because, instead of inefficient economizing at very low temperatures, LP steam now can be evaporated using LT heat (below the dew point, 190°C-170°C).

It is worth noting that our calculation of the efficiency gain offered by case ***SE*** is smaller than that calculated in a recent study by ECN [5]. This appears to have two causes. First, compared with ref. 5, this work examines relatively high (98%) CO conversion efficiencies, which tend to reduce the advantages of the ECN design. Second, we employ here a new methodology [12] for optimizing the configuration of each plant's heat integration/bottoming cycle (including pressure levels), assuming modern greenfield steam cycle design. As we have shown above, the relative performance of the two systems depends critically upon the details of the HRSC.

Finally, the exergy analysis highlights the deleterious effect of mixing water with syngas in the ECN WGS. One might instead use steam instead of water, followed by syngas cooling

[8] The difference in overall net power is smaller than that of the HRSC power because the syngas saturation scheme varies between the two plants; case ***SC*** can recover more low temperature heat (because of its higher syngas dew point) and heat up the saturator water up to 170°C, while case ***SE*** can reach only 160°C (using higher temperature heat). As a consequence, case ***SC*** needs more N_2 (thus more compression power) as a diluent to meet the stoichiometric flame temperature required for NO_x control.

[9] No power can be produced by cooling syngas from 140 to 25°C because the plants already have a large surplus of unrecoverable low temperature heat.

[10] Reducing the lower limit of P_{LP} appears to be significantly more beneficial than raising the upper limit on P_{HP} to 170 bara.

to achieve the required WGS reactor inlet temperature. Preliminary calculations indicate that this strategy could raise plant efficiency by 0.4-0.5 pp, but at the cost of heat integration complexity that exceeds the original WGS design.

COMPARATIVE PLANT ECONOMICS

Economic parameters used to estimate the costs of producing electricity are given in Table 6.[11] At these plant sizes, CO_2 removal rates are high (566.1 tonnes/hr in case ***SC***), and so transport and storage (T+S) costs are potentially modest.[12] Our model for estimating the capital costs of each major plant component is derived from the detailed capital cost data for Shell IGCC plants given in a May 2007 study by NETL [1]. Costs are escalated to mid-2008 US dollars using the Chemical Engineering Plant Cost Index [22,23]. The total plant costs (TPC), or "overnight construction costs", given in Table 8 for each case, includes engineering and overhead, general facilities, balance of plant (BOP), and both process and project contingencies (3.2 and 17% of the bare erected costs, respectively).

A comparison of capital costs between the standard and novel plants with CO_2 capture requires a careful assessment of the cost of various heat exchangers, both standard and costly syngas coolers. This work attempts only to provide a rough estimate of these costs in order to give the reader an idea of the economic impact of such changes, and the ultimate effect on the levelized cost of electricity (LCOE). The costs of syngas coolers, ~400 $/kW$_{th}$, in case ***QC*** is derived from NETL [1]. In Table 7 we have included a variant case labeled ***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 [24]. This implies a relatively high steam generation cost of ~1000 $/kW$_{th}$, comparable to the cost of convective syngas coolers used by the GE coal gasifier [25, 26].

With regard to the cost of less exotic heat exchangers used in the WGS unit, the capital cost of case ***SE*** was reduced relative to case ***SC*** based on 165 MW$_{th}$ less WGS heat exchanged, assuming an installed capital cost of $100/kW$_{th}$. The cost of additional catalyst in the ECN WGS system (estimated in ref. 5 to be ~1.5 times greater than that used in conventional WGS units) was neglected.

Cost of Electricity. The levelized cost of electricity (LCOE) for each plant is given in Table 7 for two prices on CO_2 emissions: zero and 38 $/tonne, the lowest "crossover" carbon price, i.e. the smallest CO_2 emissions price at which the LCOE for a CO_2 capture plant (in this case ***QC****) equals that of CO_2 venting case ***SV***.

Table 6. Economic assumptions employed here.

Coal price [1]	1.71 $/GJ LHV
Capacity factor	85%
Capital charge rate (CCR) CO_2	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)

At a CO_2 emissions price of 38 $/tonne, despite the lower efficiency caused by lack of syngas coolers, the LCOE for case ***QC*** is 1% less than that of ***SC***, and ***QC**** is smaller by more than 8%. This is because the TPC is lower by 6.5% and 16%, respectively, than in case ***SC***. We also might expect a higher availability for the partial water quench plants due to lack of syngas cooler fouling, leakage, and corrosion.

The ECN WGS also yields greater efficiency and lower cost (TPC is reduced by ~1%) relative to the conventional unit, causing a 3.1% reduction in LCOE of case ***SE*** relative to ***SC***. (Using the "aggressive" HRSC assumptions, however, the reduction in LCOE is roughly halved to 1.6%, as case ***SC*** benefits more from the increased bottoming cycle efficiency.) The plant integration between the WGS and steam cycle is notably simplified, promising improved reliability and ease of plant operation.

Table 7. Levelized cost of electricity for each case.

Cost component, mid 2008 $/MWh	***SV***	***SC***	***SE***	***QC***	***QC****
Installed capital (at 15% of TPI)	43.2	60.9	58.9	59.2	53.3
O&M (at 4% of TPC per yr)	9.9	14.0	13.5	13.6	12.3
Coal (at 1.71 $/GJ, HHV)	13.5	17.6	17.1	18.3	18.3
CO_2 disposal (at 7.1 $/tonne CO_2)	0.0	5.8	5.7	6.0	6.0
LCOE (no carbon price)	*66.6*	*98.3*	*95.2*	*97.2*	*89.9*
CO_2 emissions (at 38 $/tonne CO_2)	25.4	2.0	2.0	2.1	2.1
LCOE with CO_2 price (38 $/tonne)	*92.0*	*100.3*	*97.2*	*99.3*	*92.0*
Cost of avoided CO_2, $/tonne	–	51.5	46.5	49.9	38.0

CONCLUSION

This work has investigated two novel methods of humidifying syngas in a dry feed coal IGCC in order to make it more efficient when employing WGS and CO_2 capture. Compared to the reference case, the partial water quench is simpler, less expensive, but less efficient; however, it offers potentially large economic benefits. The advanced ECN WGS brings increased efficiency, a significantly reduced WGS-steam cycle integration, and somewhat improved economics.

[11] 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.

[12] Estimated costs for CO_2 T+S 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 [21].

Table 8. "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.)

						SV		***SC***		***SE***		***QC***	
Plant component	***Scaling parameter***	***So***	***n***	***f***	***Co (M$)***	***S***	***C (M$)***	***S***	***C (M$)***	***S***	***C (M$)***	***S***	***C (M$)***
Coal+sorbent handling	AR coal, tonne/day	5,447	1	0.67	40.4	5776	42.0	6,167	43.9	6,150	43.8	6,167	43.9
Coal prep & feeding	AR coal, tonne/day	2,464	2	0.67	101.6	5776	211.0	6,167	220.4	6,150	220.0	6,167	220.4
Ash handling	Coal ash, tonne/day	477.8	1	0.67	38.1	705	49.4	752.3	51.6	750	51.5	752.3	51.6
ASU, O_2 & N_2 compr.	Pure O_2, tonne/day	2,035	2	0.50	106.7	4013	197.7	4,284	204.3	4,272	204.0	4,284	204.3
Std. gasifier, SG coolers	AR coal, MW LHV	737.4	2	0.67	178.1	1730	369.8	1,847	386.3	1,842	385.6	-	-
Partial quench gasifier	AR coal, MW LHV	770.9	2	0.67	139.5	-	-	-	-	-	-	1,847	293.8
LT heat recov, saturator	AR coal, MW LHV	737.4	2	0.67	17.3	1730	36.0	1,847	37.6	1,842	37.5	1,847	37.6
COS hydrolysis	AR coal, MW LHV	797.7	2	0.67	4.7	1730	9.2	-	-	-	-	-	-
Water-gas shift reactors	AR coal, MW LHV	815.2	2	0.67	9.3	-	-	1,847	6.1	1,842	2.3*	1,847	18.9
Gas cleanup BOP	AR coal, MW LHV	815.2	2	0.67	6.1	1730	11.8	1,847	12.4	1,842	12.3	1,847	12.4
AGR (H_2S capture only)	S input, tonne/day	66.8	2	0.67	49.6	50	47.7	53.1	49.9	53.0	49.8	53.1	49.9
2nd stage CO_2 capture	CO_2 captured, tonne/hr	234.3	2	0.67	55.2	-	-	566.1	117.0	564.6	116.8	566.1	117.0
Claus plant	S input, tonne/day	136.5	1	0.67	37.6	50	19.1	53.1	19.9	53.0	19.9	53.1	19.9
CO_2 compress & dry	Compressor pwr, MW_e	27.4	1	0.67	43.0	-	-	35.1	50.8	35	50.9	35.1	50.8
General Electric 9FB GT	-	-	2	-	68.5	-	127.8	-	127.8	-	127.8	-	127.8
HRSG, ductwork, stack	GT net power, MW_e	232.0	2	0.67	33.8	628	77.2	592.9	74.3	594	74.4	592.9	74.3
ST, condenser, aux.	ST gross power, MW_e	274.7	1	0.67	74.0	360	88.6	279.6	74.9	300.6	78.6	253.9	70.2
Balance of plant	15.5% of plant cost						235.3		269.9		269.6		254.5
Total Plant Cost (TPC)							***1,523***		***1,747***		***1,745***		***1,647***
Specific Total Plant Cost ($/net kW_e)							**1,847**		**2,584**		**2,520**		**2,535**

* Cost reduced by 16.5 M$ because of reduction in 165 MW_{th} of heat exchangers.

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ANNEX A: TECHNICAL ASSUMPTIONS

Table A1. Technical assumptions used in calculations of plant performance.

SHELL GASIFICATION ISLAND	
Dried coal moisture content (wt%)	2
Syngas for drying (% of total flow)	1.0
Gasification pressure (bara)	38.5
H_2O/O_2 molar ratio (steam as moderator)	0.266
Gasification temperature	1372
Carbon conversion (with fly ash recycle)	99.79
O_2 purity (% molar fraction)	95
HP N_2 for pressurization /coal mass flow (wt ratio)	0.3193
LP N_2 for coal transport /coal mass flow (wt ratio)	0.1244
HP N_2 into syngas/dried coal (wt ratio)	0.103
Maximum oxidant temperature (°C)	100
Syngas coolers: pinch points gas-steam (°C)	15
Steam/CO value in the WGS reactor	2.5
Heat loss from heat exchangers (%)	0.5
Heat to membrane wall (% coal LHV thermal power)	1.50
Syngas coolers & wet scrubber pressure drop (%)	4.00
Electricity use: coal handling+water system (% of coal LHV)	1.00
SYNGAS TREATMENT & CONDITIONING LINE	
Heat exchangers pressure drops - gas side (%)	0.05
Heat exchanger heat losses (%)	0.5
Quench water pump hydraulic efficiency	0.85
Quench water pump elec. × mechanical efficiency	0.9
Saturator pump hydraulic efficiency	0.8
Saturator pump electrical × mechanical efficiency	0.9
ACID GAS REMOVAL (AGR) UNIT	
Selexol for selective removal of CO_2 & H_2S:	
Gas temperature at AGR inlet (°C)	38
H_2 co-absorbed (%)	0.533
Sulfinol-M for removal of H_2S:	
LP steam for stripping (MW_{th} per kg/s of stripped H_2S)	13.4
Specific electricity use: P_{elec} /SG mass flow (kJ/kg)	9.36
CO co-absorbed (%)	0.265
H_2 co-absorbed (%)	0.268

WATER GAS SHIFT (WGS) UNIT	
HT/LT WGS reactor approach temperatures (°C)	10/10
HT/LT WGS syngas input temperatures (°C)	250/250
CO_2 COMPRESSION AND DRYING	
Specific electricity use: P_{elec}/CO_2 mass flow (kJ/kg)	222.7
AIR SEPARATION UNIT (ASU)	
Air compressor (axial), polytropic efficiency	0.86
Pressure of O_2 and N_2 delivered by ASU (bara)	1.05
Excess air	0.06
Compressor electrical × mechanical efficiency	0.92
O_2 compressor (3 ICs), 4 radial stages, average η_{poly}	0.845
LP N_2 compressor (1 IC), 2 radial stages, average η_{poly}	0.858
HP N_2 compressor (2 ICs), 3 radial stages, average η_{poly}	0.79
Dilution N_2 compressor (axial machine,1 IC), avg. η_{poly}	0.887
Intercooler exit temperature (°C)	45
POWER ISLAND	
(2 GE 9FB GTs + 2 HRSGs + 1 steam turbine)	
HRSC, 3 pressure levels; condenser pressure: 0.05 bara	
Approach temperature: Steam SH – Hot flue gas (°C)	25
Min. Delta T pinch points and sub cooling in HRSG (°C)	8
Pumps: hydraulic efficiency	0.84
Pumps: electrical × organic efficiency	0.9
Maximum steam pressure (for the HP level), bara	140 (170*)
Minimum steam pressure (for the LP level), bara	5 (3*)
Steam turbine mechanical efficiency	0.98
Steam turbine generator electrical efficiency	0.99
Economizers: pressure losses (%)	16
Superheater and reheater pressure losses (%)	8
OTHER UNITS	
Sour water stripper, LP steam requirement, kJ/kg coal	84.0
Electricity for cooling components (% of rejected heat)	0.5

* "Aggressive" HRSC parameter assumptions