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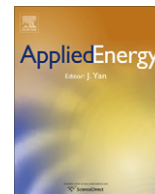
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ABSTRACT

The Shell coal integrated gasification combined cycle (IGCC) based on the gas quench system is one of the most fuel flexible and energy efficient gasification processes because is dry feed and employs high temperature syngas coolers capable of rising high pressure steam. Indeed the efficiency of a Shell IGCC with the best available technologies is calculated to be 47–48%. However the system loses many percentage points of efficiency (up to 10) when introducing carbon capture. To overcome this penalty, two approaches have been proposed. In the first, the expensive 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 novel water–gas shift (WGS) configurations that requires substantially less steam to obtain the same degree of CO conversion to CO₂, and thus increases the overall plant efficiency. We simulate and optimize these novel configurations, provide a detailed thermodynamic and economic analysis and investigate how these innovations alter the plant’s efficiency, cost and complexity.

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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 CO₂ capture for sequestration in deep geologic formations away from the atmosphere.

Numerous studies indicate that bituminous coal-based electric power with CO₂ 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 (about 900 °C) synthesis gas by generating high pressure steam prior to syngas cleaning and chemical processing. In plants with CO₂ 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 CO₂ 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 CO₂ capture. In this regard, dry feed gasifiers are at a disadvantage relative to coal–water slurry fed gasifiers (e.g. GE Energy and Conoco-Phillips 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.

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Nomenclature

Abbreviations and symbols

AGR	acid gas removal	O&M	operation and maintenance
AR	as received	p	absolute pressure, bar
ASU	(stand-alone, cryogenic) air separation unit	PC	pulverized coal (steam electric power plant)
BOP	balance of plant	pp	percentage points
COMPR	compressor	q	mass flow rate, kg/s
COMPR	compressor	QC	a particular plant configuration, defined below
COND	condenser	QW	quench water
CCR	annual capital charge rate	RH	reheat
CCS	CO ₂ capture and storage	SC	syngas cooler
dp	infinitesimal pressure drop, bar	SC	a Shell IGCC equipped with a conventional two-stage sour WGS unit for CCS, as defined in Section 2
ECON	economizer	SE	a Shell IGCC equipped with the novel ECN WGS unit for CCS, as defined in Section 2
ECN	Energy Research Centre of the Netherlands	SN	a Shell IGCC equipped with an innovative high efficiency WGS unit for CCS, as defined in Section 2
GE	general electric	SV	a Shell IGCC without CCS, as defined in Section 2
GT	gas turbine	S/CO	steam-to-CO mole ratio
HHV	high (or gross) heating value	SCOT	Shell Claus offgas treating
HP	high pressure (~140–170 bar)	SG	synthesis gas, or syngas
HT	high temperature	SH	superheater
HRSC	heat recovery steam cycle	ST	steam turbine
HRSG	heat recovery steam generator	TIT	turbine inlet temperature, °C
IDC	interest during construction	TOT	turbine outlet temperature, °C
IGCC	integrated gasification combined cycle	TPC	total overnight plant cost
L/G	liquid-to-gas mass ratio	TPI	total plant investment (TPC + IDC)
LEAP	Laboratorio Energia Ambiente Piacenza	T_{sat}	saturation temperature, °C
LCOE	levelized cost of electricity	v	specific volume, m ³ /kg
LHV	lower (or net) heating value	VGV	variable (air inlet) guide vanes
LLT	extra low temperature	WGS	water–gas shift
LP	low pressure (3–15 bar)	x_A	volume fraction of component A
LT	low temperature	ΔT	temperature difference, K
MDEA	N-methyl-diethanolamine	ΔT_{TIT}	TIT de-rating, K
MP	medium pressure (20–50 bar)	η	polytropic efficiency
NETL	National Energy Technology Laboratory		
NG	natural gas		

Researchers at the Energy Research Centre of the Netherlands (ECN) have recently developed an advanced WGS design that significantly reduces the flow of steam required for conversion of CO and H₂O to CO₂ and H₂ [5]. This system, which is modeled and optimized in this study, has recently been implemented at pilot scale at NUON's Buggenum IGCC plant in the Netherlands. In addition, we investigate an alternative advanced WGS layout specifically designed and optimized to further minimize the steam consumption and thus the carbon capture penalty.

This study compares the thermodynamic and economic performance of a bituminous coal-based IGCC plant using Shell gasification technology – with and without CO₂ 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 one of the two advanced WGS designs. The plants are designed, modeled in detail and optimized to maximize the net electric efficiency, using both exergy analysis and numerical optimization algorithms. Our goal is to understand what the preferred IGCC design is for dry feed, entrained flow gasifiers with relatively high levels of carbon capture (>90%).

2. Methodology

We model four cases, three with CO₂ capture:

SV – a Standard (i.e. with syngas coolers) Shell coal gasifier-based IGCC plant with syngas coolers and CO₂ Venting,

SC – a Standard Shell IGCC plant with CO₂ capture that uses a Conventional two-stage WGS unit,

SE – a Standard Shell IGCC plant with CO₂ capture that uses the advanced ECN WGS design,

SN – a Standard Shell IGCC plant with CO₂ capture that uses an optimized New WGS configuration derived from the ECN design,

QC – a partial water Quench Shell IGCC plant with CO₂ capture, using a Conventional 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 CO₂ 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 H₂-rich syngas) using the “Gas/Steam” (GS) simulation code developed at Politecnico di Milano and presented in [10,11,15,18], (6) configuring and optimizing the layout of the heat recovery steam cycle (HRSC) for each plant using a new method developed by Martelli [12,13] that maximizes the power output of the steam cycle, and (7) adding the cost framework required for a full techno-economic comparison between cases. Given that the first generation of IGCC power plants with CO₂ capture are likely to be operated as base-load plants, analyses of load-following performance and off-design calculations were not performed in this study. Indeed, commercially available gasification processes and

air separation units (ASU) have only a limited ability to operate at part-loads.

3. System design overview

In order to make “fair” comparisons between the four cases studied here and thus highlight their intrinsic differences, the basic plant configuration, key design parameters and modeling assumptions are held fixed for all cases. All plants share a common design: a Shell coal gasifier with a stand-alone cryogenic ASU, two GE 9FB gas turbines, and a three pressure level HRSG with reheat. Plants with CO₂ capture have the same extent (98%) of CO conversion in the water–gas shift (WGS) unit, and employ a two-stage Selexol™-based acid gas removal unit (AGR), CO₂ compressor and syngas saturator upstream of the gas turbines. As a result, thermodynamic (and economic) differences observed between cases **SC**, **SE** and **SN** should reflect only the design of the WGS unit, i.e., the number of reactors, their layout and downstream heat recovery. Cases **QC** and **SC** differ only in the method of syngas cooling and humidification upstream of identical WGS units.

3.1. Gasifier Island and ASU (common units)

The basic design of the gasification island is illustrated in Fig. 1 (which represents the overall plant **SC**); calculation details and assumptions are given in Table A1. East Australian bituminous coal (Table 1) is milled, dried to a moisture level of 2% (weight basis), and fed into the gasifier via lock-hopper pressurization using N₂ as a transport gas.

Table 1

Composition (in%, weight basis) and heating value of as received East Australian bituminous coal used here [9].

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

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 bar 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 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 (reported in 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].

3.2. 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

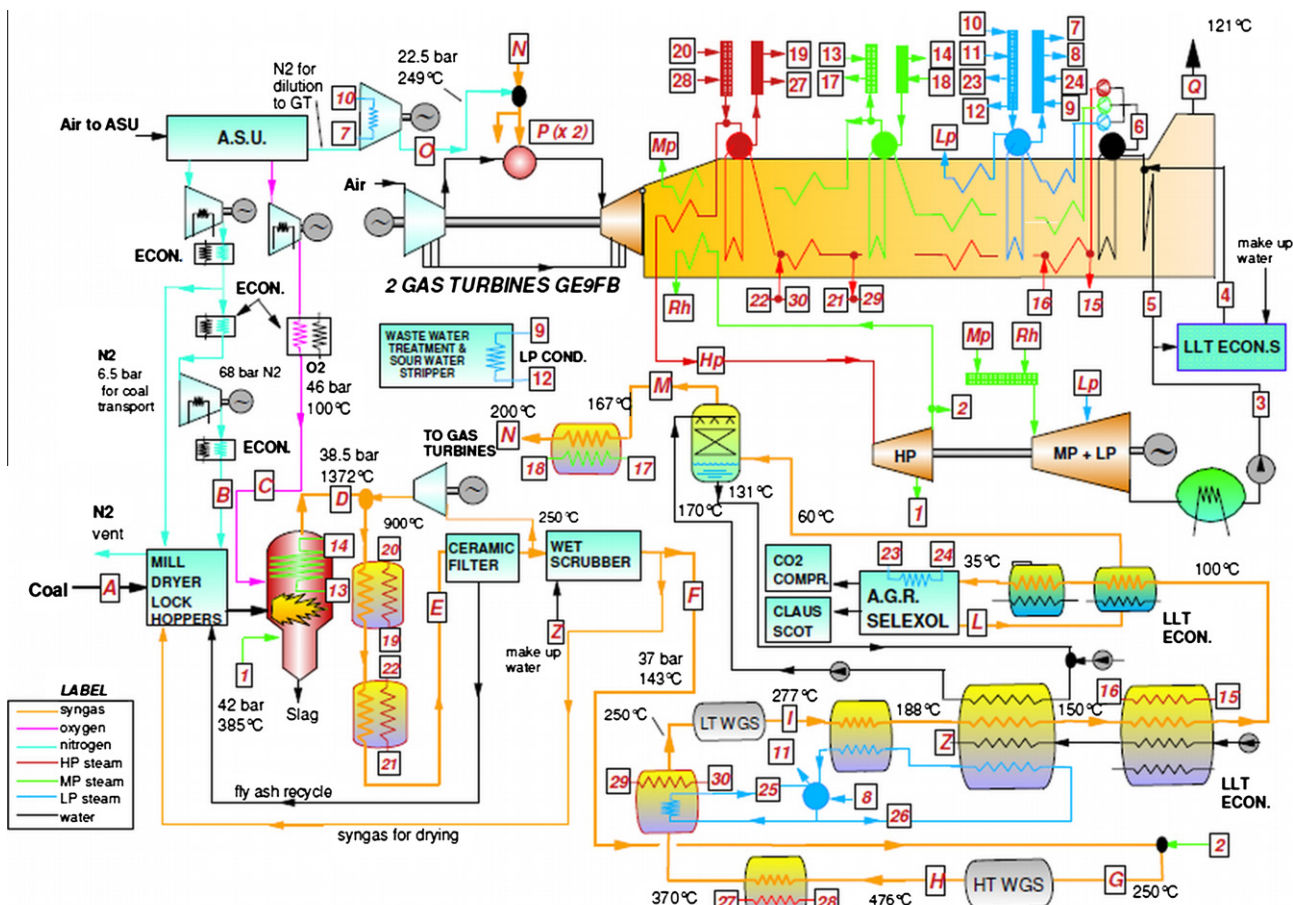


Fig. 1. Plant schematic for case **SC**, the standard Shell IGCC + CO₂ capture with a conventional WGS.

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 H₂S, and HCN to NH₃. The syngas is cooled to 40 °C and sent to a Sulfinol MTM acid gas removal (AGR) system that strips out virtually all of the H₂S (and 16% of the CO₂) which is sent to an O₂-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% O₂) by diluting the syngas with all the available N₂ and some steam in order to lower the stoichiometric flame temperature to 2027 °C [15].¹ Heat is efficiently recovered from the turbine exhaust in a triple 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 [12,13].

3.3. Case SC

Our design for the standard Shell IGCC with CO₂ capture (represented in Fig. 1) mirrors that of Ref. [1] 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 [14] and also to achieve an “overall carbon capture fraction”² of 93.1%. The scrubbed syngas is pre-heated, 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 CO₂ and H₂. 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 assumptions adopted to model the WGS reactors are given in Table A1. The shifted syngas is cooled to 38 °C, sent to the SelexolTM process for H₂S and CO₂ selective removal, saturated with water, diluted with some N₂ supplied by a dedicated compressor, heated to 200 °C, and burned in the gas turbines. According to [1], the two-stage selective Selexol process is capable of capturing 96.54% of the inlet CO₂, and co-capturing 0.53% of H₂ and 0.44% of CO. Thus, the overall carbon capture level (defined as mole flow rate of captured carbon divided by mole flow rate of carbon in fuel at inlet) is 93.0%. The captured CO₂ stream is dehydrated and compressed from 1.8 to 150 bar for pipeline transport and storage in geologic formations; the H₂S rich stream is treated in a conventional Claus unit followed by a Shell Claus Offgas Treating (SCOT) process.

3.4. Case SE

In contrast to the two-stage conventional WGS unit,³ the advanced ECN system [5], as implemented here (reported in 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, as represented in Fig. 2) or 160 °C water (reactors 2–4) needed to satisfy the requirement 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 H₂O per CO, the H₂O/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.

Configuring the advanced WGS system shown in Fig. 2 is relatively straightforward, with an equal number of free variables and constraints. Making the simplifying assumption that all water input streams have the same temperature, there are ten free variables 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. Therefore, a single solution of the above (nonlinear) design problem exists. To have a fair comparison, the same key assumptions adopted to model the WGS reactors of case SC are adopted for cases SE and SN (i.e., those assumptions reported in Table A1). 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.

3.5. Case SN

The water quench employed by the ECN WGS is much more irreversible than the steam injection employed by the conventional WGS layout, as confirmed by the exergy analysis presented in Section 5. It is instructive to quantify in the ECN design the potential gain in efficiency that can be achieved by replacing the water quench with steam injection, followed by syngas cooling. Due to the high temperature of medium pressure steam (via a turbine bleed at 330 °C), gas coolers are necessary to bring the syngas temperature to the desired inlet temperature of the downstream reactors (i.e. 250 °C). This “alternative” WGS design is shown in Fig. 3. The outlet temperatures of coolers 1, 3 and 5 are adjusted to maximize the production of higher pressure steam with a pinch point temperature difference of 15 °C, while the outlet temperatures of coolers 2, 4 and 6 are fixed at 250 °C. The addition of syngas coolers increases the degrees of freedom of the design in comparison with the original ECN WGS: steam mass flow rates can be varied without violating the constraints imposed by the inlet temperature of the reactors (these temperatures are set by the syngas coolers). Hence this WGS unit has nine variables (four syngas split fractions, four steam/water flows, and the steam temperature) and only seven constraints (S/CO ≥ 2.5 in each reactor, the steam temperature, $T = 250$ °C at inlet of the first reactor, the temperature of MP steam and the overall CO conversion efficiency). This provides two degrees of freedom (i.e. number of variables minus number of constraints) for optimizing the design. Further details about the optimization of this WGS layout are in Section 5.

3.6. Case QC

In the partial water quench system with CO₂ capture (represented in Fig. 4), the standard syngas cooling system (gas quench via syngas recycle, followed by syngas coolers) of case SC is

¹ In case SV, the most efficient method of syngas dilution involves using all of the available N₂ and a small amount of steam. In all CO₂ capture cases we first saturate the syngas using low temperature heat that is not otherwise well utilized in the bottoming cycle, and then add N₂ 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 CO₂ stored underground.

³ Novel WGS designs that include upstream saturators and downstream desaturators are not considered in this work.

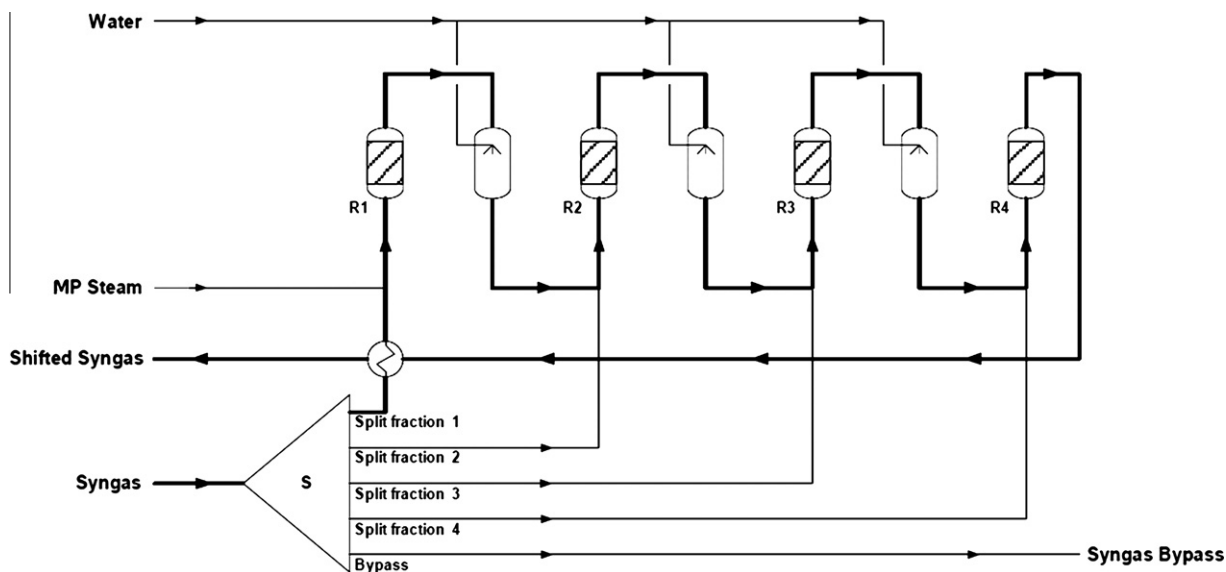


Fig. 2. Layout of the ECN advanced WGS.

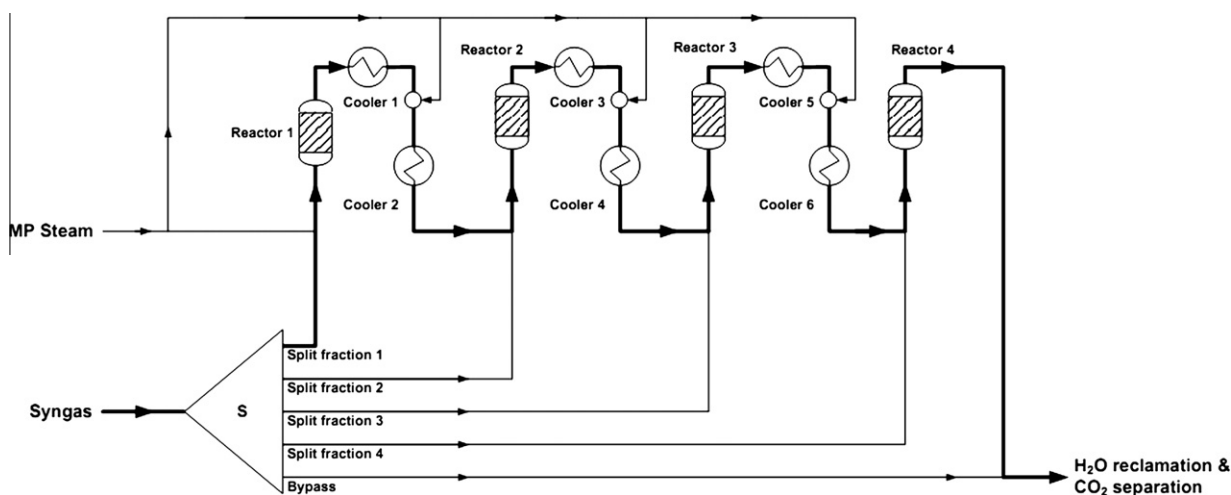


Fig. 3. Layout of the advanced WGS unit to be optimized (employed in plant SN).

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 bar saturation temperature, $T_{\text{sat}} = 248.1$ °C.⁴

The wet scrubber was modeled as an adiabatic, countercurrent absorption column with 5 equilibrium stages and a fixed liquid-to-gas mass ratio⁵ (L/G) of 0.25 [16,17]. Thus constrained, the scrubber has only a limited capability to increase the S/CO ratio of the syngas (Fig. 5). 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 about 0.08 to only 0.3, requiring the addition of 382 MW of MP steam (this thermal power is obtainable by cooling the given steam mass flow rate to the environment temperature) to achieve the target S/CO ratio of 2.5. In contrast, raw syngas that is

⁴ 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.

⁵ 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.

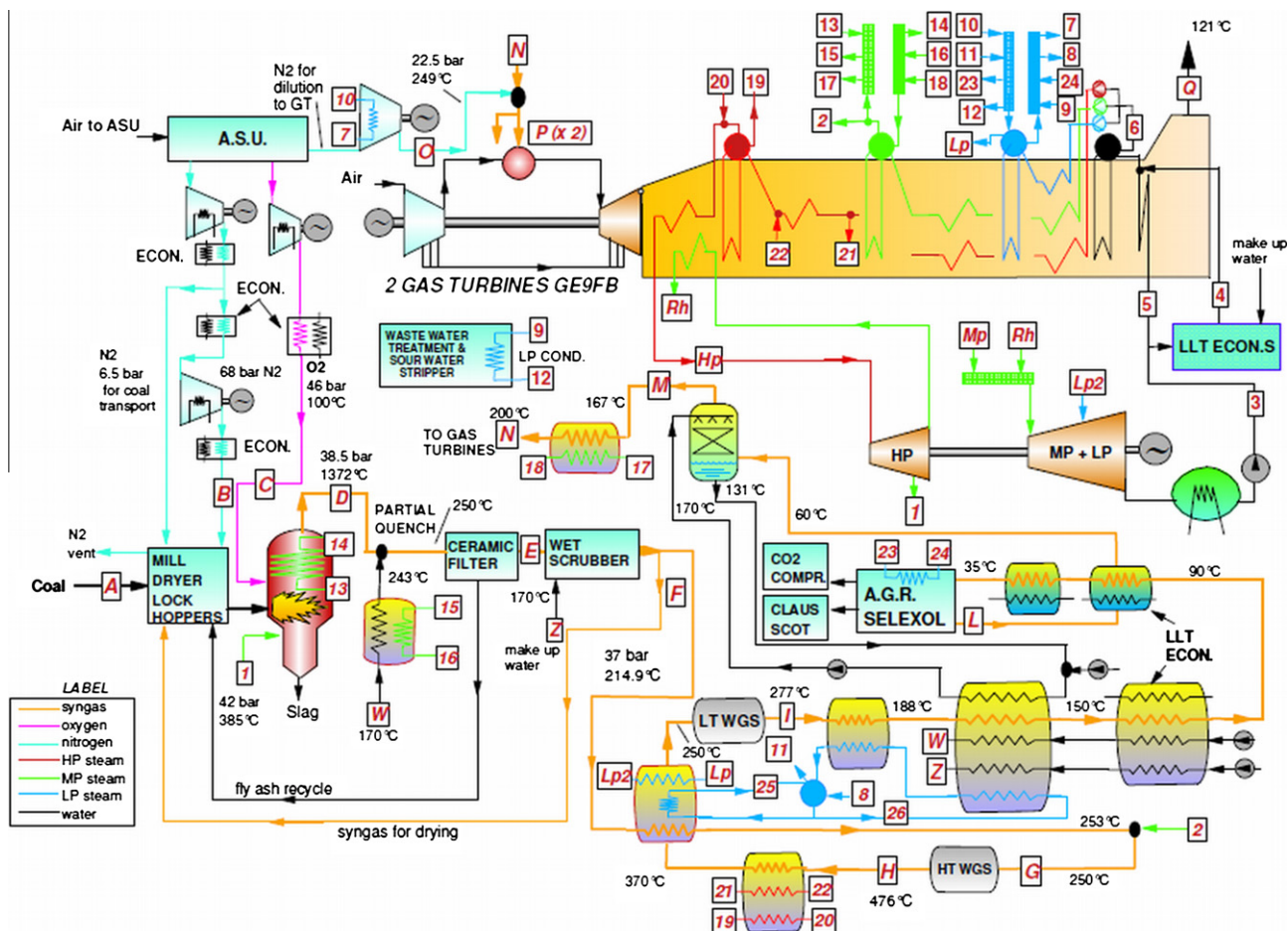


Fig. 4. Plant schematic for case QC, the partial water quench IGCC with the conventional WGS.

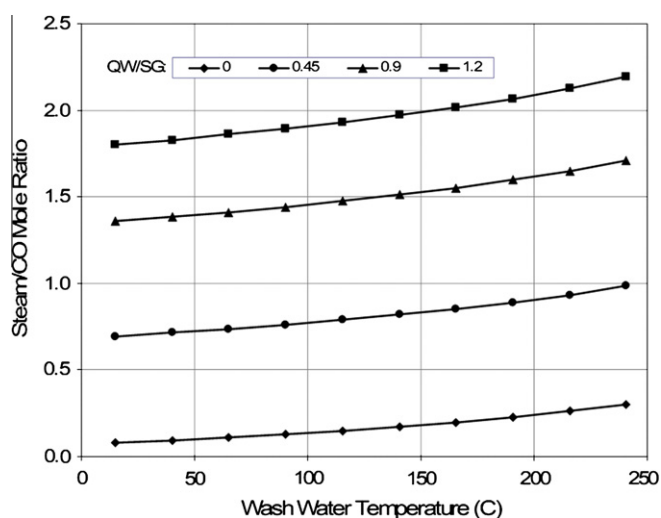


Fig. 5. Steam-to-CO ratio of scrubbed syngas as a function of wash water temperature and quench water-to-syngas mole ratio, QW/SG. The L/G ratio of the scrubber is fixed at 0.25.

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 of 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 pre-cooler), 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 MW vs. 205 MW).

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).

4. Power island design and modeling

The power unit consists of two GE 9FB gas turbines, each with a triple 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

Table 2

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

GT parameter	GE 9FB reference data		Calibrated GT model (using “GS” software)		
	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 rate, kg/s	637.13	637.13	530.45	573.5	575.43
Fuel mass flow rate, 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 mass flow rate, 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 and TOT.^b TIT de-rated using Eq. (1).

the 9FB is a scaled-up version of the 7FB, it is expected to run on syngas without significant issues. The GT has been modeled using the “Gas/Steam” simulation code [10,11] and the performance estimation method of Chiesa [18].

The GT model was calibrated to reproduce the same output of a GE 9FB burning natural gas, and then adjusted for operating on syngas (see Table 2). We adopt the control strategy discussed in Ref. [15] 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 vs. non-dimensional mass flow rate), details of which are confidential and not readily predicted. In general, variations in pressure ratio can be neglected if the compressor air mass flow rate exceeds ~85% of its design point (NG) value.

We de-rated the TIT to offset the relatively high heat transfer between the H₂O-rich gases and the turbine blades, and thus insure standard blade lifetimes. Rather than using the method of Ref. [15] which adjusts the TIT until critical blade temperatures obtained during syngas operation match those during NG firing (using the same cooling flows), we employed instead the more conservative correlation of Oluyede [19]:

$$\Delta T_{\text{TIT}} = 13.312 * (x_{\text{H}_2})^{0.69} \quad (1)$$

which, for our shifted syngas, de-rates the TIT by 100 °C (compared to 35 °C using the method of Ref. [15]). Note that Eq. (1) expresses ΔT_{TIT} in °F. x_{H_2} denotes the volume fraction of H₂ in the diluted syngas fed into the combustor.

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 [20], but H₂-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 CO₂ capture cases.

Downstream of each gas turbine is a triple pressure level HRSG with reheat. The heat integration of each plant was individually optimized using a the methodology recently developed by Martelli [12,13] in which the arrangement of the heat recovery steam network and the design of the heat recovery steam generator (HRSG) are simultaneously optimized. By means of this optimization algorithm (called the “HRSC Optimizer”), steam mass flow rates raised in each tube bank of syngas coolers and HRSG, steam mass flow rates of extractions used by condensers, reboilers and steam users, pressure levels, pressure of turbine extractions, and pinch point temperature differences were numerically optimized to maximize

the net electric efficiency of the overall IGCC plants. Within this design methodology, capital costs are not explicitly considered in the objective function, but economic feasibility is obtained by adopting realistic design constraints and criteria derived from the industrial practice. For instance, all syngas coolers were constrained to be either evaporators or economizers with a minimum pinch point temperature difference of 15 K, and 50 K for the HT syngas cooler. Within the HRSG, pinch point temperature differences were set ≥ 8 K, sub-cooling temperature differences (i.e., temperature difference between saturation temperature and economizer exit temperature, adopted to avoid steaming in the economizers) = 3 K, and the superheat/reheat approach temperature difference (i.e., temperature difference between the flue gas and superheated steam at SH/RH exit) ≥ 25 K. The condenser pressure was assumed equal to 0.05 bar (i.e., we assumed the availability of either a cold river or sea water). These constraints were derived from the industrial design practice for large scale combined cycles. Regarding the intensive decision variables of the steam cycle (i.e., steam pressures and temperatures), we employed a “conservative” set of HRSC parameters in which $p_{\text{HP}} \leq 140$ bar, $p_{\text{MP}} \geq 39$ bar, $p_{\text{LP}} \geq 5$ bar, and T_{SH} (superheat/reheat temperature) = 540 °C. The limit on p_{HP} was fixed in order to adopt a standard drum type evaporator, the lower bound on p_{MP} was determined by the direct integration with the Shell gasifier membrane wall (which evaporates MP steam and whose steam pressure must be sufficiently higher than the syngas pressure), the lower bound on p_{LP} is commonly introduced to avoid expensive oversized pipes, tubes and headers, while T_{SH} was selected equal to 540 °C in order to use conventional steam turbines for combined cycles. 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} was limited to 170 bar (i.e., assuming to use Benson once-through evaporators, [21]), p_{MP} is free (i.e., neglecting the issues related to the gasifier membrane wall), $p_{\text{LP}} \geq 3$ bar, and $T_{\text{SH}} = 545$ °C (i.e., the gas turbine outlet temperature, about 565 °C for cases with CCS, minus 20 K of approach point temperature difference).

5. Plant performance

Plant performance is given in Tables 3 and 4.

5.1. SV: CO₂ venting reference case

The efficiency of the plant without capture, about 48% LHV, is higher than a similar case presented previously in [22] for three reasons: the GE 9FB is more efficient than the Siemens V94.3A,⁶ the clean syngas is highly pre-heated before entering the combustor, and the heat recovery steam cycle is fully optimized (pressure levels,

⁶ 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 HRSC for each case.

HRSC variable	SV	SC	SE	SN	QC
HP-MP-LP pressures, bar	140–39–5	140–39–5	140–39–5	140–39–5	140–42–5
HP steam produced, kg/s	248.27	271.24	237.22	261.37	161.64
MP steam produced, kg/s	25.99	60.48	38.19	62.52	0.00
LP steam produced, kg/s	7.79	35.74	15.63	19.16	55.66
HP steam at turbine inlet, kg/s	248.27	271.24	237.22	261.37	161.64
MP steam into gasifier, kg/s	7.24	7.73	7.71	7.73	7.73
MP steam bleed into WGS, kg/s	0.00	153.66	71.25	117.65	28.97
MP steam bleed into GT, kg/s	21.39	0.00	0.00	0.00	0.00
MP steam at turbine inlet, kg/s	243.17	171.31	195.75	195.95	152.31
LP steam at turbine inlet, kg/s	250.97	184.62	211.39	215.10	207.98
Net HRSC electric power, MW	359.84	281.08	300.72	310.85	254.34

Table 4

Plant performance, with breakdown of power consumption by unit.

Plant performance	SV	SC	SE	SN	QC
<i>Elec. power consumption, MW:</i>					
Coal handling, gasifier	17.30	18.47	18.42	18.42	18.47
ASU, O ₂ & N ₂ compr., pumps	144.55	123.03	128.69	128.69	123.65
AGR, Claus, SCOT units	1.04	19.69	19.75	19.75	19.69
CO ₂ drying & compression	0.00	35.08	35.18	35.18	35.08
Auxiliary elec. power, MW	162.89	196.27	202.04	202.04	196.89
<i>Elec. power production, MW:</i>					
GT net power (2 × GE 9FB)	627.60	592.92	593.92	593.92	592.92
HRSC net elec. power	359.76	279.57	300.58	310.86	253.91
Net electric power, MW	824.47	676.21	692.46	702.60	649.94
Coal input, MW (LHV basis)	1729.7	1846.7	1841.6	1841.6	1846.7
Net electric efficiency (LHV basis), %	47.66	36.62	37.60	38.15	35.19
Overall carbon capture level, %	0.00	93.09	93.09	93.09	93.09
Specific CO ₂ emissions, g/kWh	667.75	52.01	51.70	51.07	54.81

steam/gas temperatures and syngas cooler integration). Our results predict that such high efficiencies are possible using well optimized, commercial technologies.

5.2. **SC, SE, QC:** CO₂ capture cases

These plants share the same 98% CO conversion efficiency and AGR, and thus achieve the same overall degree of carbon capture, i.e., 93.1%. The difference between them are the syngas humidification process and the WGS layout.

5.3. **SV** vs. **SC:** CO₂ 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 CO₂ compression. (Note that the N₂ 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.

5.4. **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 rate of steam is only one of three important factors governing the steam turbine's mechanical power, equal to the integral of $q * \eta * v * dp$ along the expansion (where q is the mass flow rate, v is the steam specific volume, η is the polytropic efficiency, and dp the infinitesimal pressure drop). Despite the higher mass flow rate of HP steam in case **SC**, the product $v * \eta$ is lower because v is relatively small (HP section inlet: $v = 0.0243 \text{ m}^3/\text{kg}$, MP section inlet: $v = 0.0937 \text{ m}^3/\text{kg}$, LP section inlet: $v = 0.522 \text{ m}^3/\text{kg}$) 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.

5.5. **SC** vs. **SE:** ECN 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 2.7% (1 pp) higher than the conventional WGS case, i.e., **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 (7.5%) more power. This discrepancy is the most important factor underlying the difference in overall plant efficiency between the two cases.⁷

To investigate in detail why the bottoming cycle of **SE** generates more power than that of **SC**, we use a detailed exergy analysis. As observed by Kunze et al. [23], a structured exergy analysis is a powerful tool for identifying subsystems, components and processes with the highest potential for improvements in efficiency. For example, Kunze et al. compute unit-level exergy losses throughout an IGCC, and identify those with the largest exergy loss: the gas turbine, gasifier, AGR unit and WGS section. However,

⁷ The difference in overall net power is smaller than that of the HRSC power because of the difference in the electric consumption of the N₂ compressor for syngas dilution. Indeed 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 N₂ (thus more compression power) as a diluent to meet the stoichiometric flame temperature required for NO_x control.

Table 5

Comparison of exergy losses (expressed in 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 HRSCs.

Plant section	Exergy losses, MW			
	Process	SC	SE	Difference
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 in WGS unit	58.21	62.50	−4.28
LT syngas cooling section	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

a unit-level exergy analysis is not sufficiently detailed to explicate the fundamental differences between the WGS designs for cases **SC** and **SE**. Thus, we employ a more detailed analysis that computes exergy losses for individual physical/chemical phenomena (e.g., chemical reaction, heat transfer, mixing/separation process, pressure drop, etc.) and basic components (heat exchanger, mixer, reactor, turbine, compressor, etc.). The ExerCom routine for Aspen Plus [24] is used to calculate the physical exergy (i.e., the amount of reversible work which can be obtained by cooling and expanding a given stream to the environmental/reference pressure and temperature) and the chemical exergy of streams (i.e., the amount of reversible work which can be obtained by chemically converting, cooling and expanding a given stream up to the environmental pressure, temperature and composition) with reference to the environmental conditions defined by Szargut et al. in [25].

The primary differences between the HRSCs of cases **SC** and **SE** lie in the WGS unit and LT syngas cooling; the remaining units (i.e., HT syngas coolers, reboilers in the AGR, gas and steam turbines, and HRSGs) are essentially identical. Therefore, within this comparative approach, it is possible to limit the detailed exergy analysis to the processes and components of the WGS unit and LT syngas cooling section (see Table 5). Table 5 indicates that the process of mixing syngas with water in the ECN WGS is about 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 syngas cooling section where less latent heat of condensing water is wasted (exergically).⁸ As a proof of accuracy and correctness of our comparative approach, it is worth noting that the comparative exergy analysis accounts almost exactly for the difference in HRSC power between the two cases.

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 H₂O to the syngas, and efficient recovery of this heat is critical to high overall efficiency. The LP pressure level is particularly important⁹ in this regard; at $p_{LP} = 3.7$ bar, 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, 190–170 °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 [12,13] 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.

5.6. **SN** vs. **SE**: comparison between advanced WGS layouts

The exergy analysis of the WGS unit shows that reducing the moisture content of the shifted syngas lessens the carbon capture penalty by lowering the exergy loss in the LT cooling section. This observation is fundamental to optimize the WGS unit in case **SN**: the free variables of the design can be adjusted to minimize the water content at the exit of the last WGS reactor (i.e. molar flow of H₂O in the shifted syngas), or equivalently, the steam/water consumption (i.e. molar flow of H₂O injected in the WGS unit). The optimization has been carried out via the Sequential Quadratic Programming algorithm of Lang and Biegler [26] which is available in Aspen Plus. The resulting optimized layout is surprisingly simple: one only steam injection and two only feed gas split flows are necessary to achieve optimum performance (see Fig. 6). Compared with ECN case **SE**, the total water consumption is 9.4% lower and the LHV electric efficiency is 0.55 pp higher). A sensitivity analysis around the optimum configuration reveals that the **SN** plant efficiency decreases by only 0.08 pp if employing three reactors instead of four (due to the absence of one reactor, this layout needs 6% more steam but keeps the same layout shown in Fig. 6). Further reducing the number of reactors from three to two leads to a more significant loss of efficiency; indeed the optimized **SN** configuration with two reactors is exactly equal to the conventional WGS layout (i.e. all the feed gas enters the first reactor), case **SC**.

Replacing the deleterious ECN water quench with the steam injection has multiple effects: 1. it reduces the water–syngas mixing irreversibility, 2. it increases the number of degrees of freedom leaving room for process optimization, 3. it reduces the total water consumption, and 4. it raises the plant efficiency by 0.55 pp. However this gain of efficiency is achieved by employing an optimized heat recovery steam cycle that is fully integrated with the power island and (additional) syngas coolers. Hence, while an economic comparison between cases **SN** and **SE** has not been attempted here, we suspect that the gain in efficiency is insufficient to justify the increased capital costs (of the required syngas coolers and the larger and more complex HRSC) and operating difficulties (start-up, shut-down and off-design operation) due to the integration between the steam generator and the WGS coolers. For these reasons, although the **SN** configuration is more efficient, the ECN WGS layout appears to achieve a better balance between capital cost, plant complexity and energy efficiency.

6. Comparative plant economics

Economic parameters used to estimate the costs of producing electricity are given in Table 6. Our costing analysis uses the EPRI TAG methodology [27], and is kept simple and transparent to facilitate revision by readers wishing to use different economic assumptions. The assumed coal price, 1.71 \$/GJ (LHV basis), was adopted from Ref. [1]. Interest during construction (IDC) is taken to be 12.3% of overnight capital (TCP), based on a 4-year construction schedule with equal, annual payments, and a real discount rate of 7.8%/yr. The 30 year levelized capital charge rate (CCR), applied to the total plant investment (TPI = TPC + IDC), is assumed to

⁸ 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.

⁹ Reducing the lower limit of p_{LP} appears to be significantly more beneficial than raising the upper limit on p_{HP} to 170 bar.

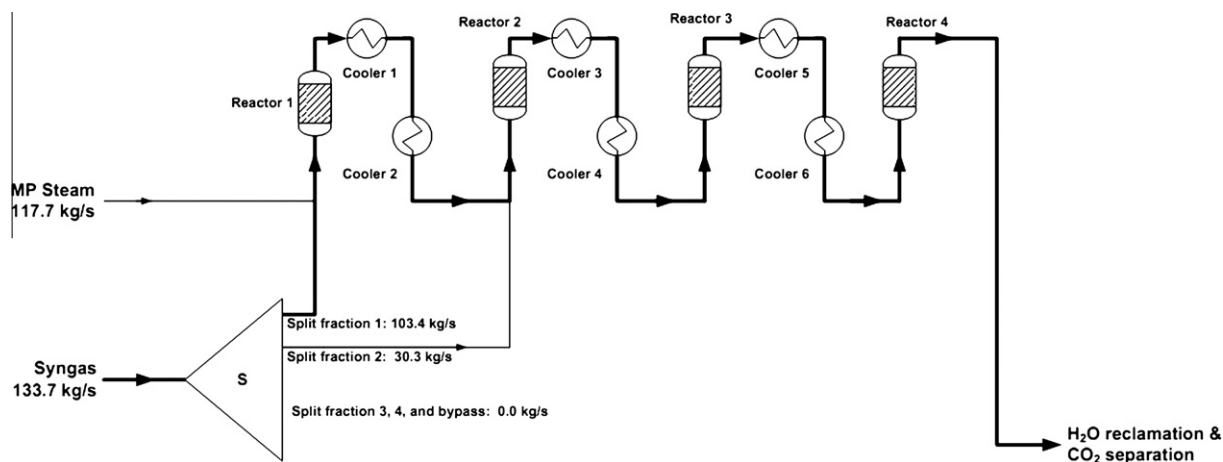


Fig. 6. Optimized layout of the proposed advanced WGS unit employed in option SN.

Table 6

Economic assumptions employed here.

Coal price [1], \$/GJ HHV	1.71
Capacity factor, %	85.0
Capital charge rate (CCR) [12], %/year	15.0
Interest during construction (IDC) [12], % of overnight capital	12.3
Operation & maintenance (O&M), % of overnight capital/yr	4.0
CO ₂ transport + storage costs [13], \$/tonne of CO ₂	7.1
U.S. dollars valued in year	2008
	(mid-year)

be 15%/yr.¹⁰ Operation and maintenance (O&M) costs are fixed at 4% of TPC per year. The plant capacity factor is assumed to be 85%. At these plant sizes, CO₂ removal rates are high (566.1 tonnes/hr in case **SC**), and so transport and storage costs are potentially quite modest, only 7.1 \$/tonne CO₂ here.¹¹ 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 [29,30]. The total plant costs (TPC), or “overnight construction costs”, given in Table 7 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 CO₂ 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 (exchanged thermal power basis), in case **QC** is derived from NETL [1]. 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 less WGS heat exchanged, assuming

an installed capital cost of 100 \$/kW (exchanged thermal power basis). The cost of additional catalyst in the ECN WGS system (estimated in Ref. [5] to be about 1.5 times greater than that used in conventional WGS units) was neglected.

6.1. Cost of electricity

The levelized cost of electricity (LCOE) for each plant is given in Table 8 for two prices on CO₂ emissions: zero and 45.6 \$/tonne, the lowest “crossover” carbon price, i.e. the smallest CO₂ emissions price at which the LCOE for the most advantageous CO₂ capture plant (in this case **SE**) equals that of CO₂ venting case **SV**.

At a CO₂ emissions price of 45.6 \$/tonne, despite the lower efficiency caused by lack of syngas coolers, the LCOE for case **QC** is 0.9% less than that of **SC**. This is because the TPC is lower by 6.5% 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 about 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.

7. Conclusion

This work has investigated two methods of humidifying syngas in a dry feed coal IGCC in order to make it more efficient when employing WGS and a high level of CO₂ capture. Compared to the reference case with gas quench and high temperature syngas coolers, the partial water quench is simpler, less expensive, but less efficient (–1.4 pp LHV efficiency); however, it offers the potential of appreciable economic benefits (–6.5% total plant cost and –0.9% levelized cost of electricity). On the other hand, keeping the gas quench system with high temperature syngas coolers, it is thermodynamically advantageous to employ advanced WGS layouts with low steam and water consumption: the ECN configuration gains 1 pp while the optimized WGS layout gains 1.5 pp of LHV efficiency.

Among these innovative configurations, the ECN WGS coupled with the Shell gas quench system appears to be the most attractive one because brings increased efficiency, a significantly reduced WGS steam cycle integration and also improved economics (–1%

¹⁰ CCR is calculated using EPRI TAG methodology [27] with the modified accelerated capital recovery system (MACRS) for tax depreciation, assuming an inflation rate of 2%/year, a real discount rate—assumed equal to the weighted, after-tax cost of capital—of 7.8%/year (based on real costs for debt and equity capital of 4.4 and 14.0%/year, respectively, and a capital structure consisting of 55% debt and 45% equity), a book life of 30 years with no plant salvage value, a tax life of 20 years, a federal + state income tax rate of 38.2%/year, and annual owners costs of 5.5% of TPI. Note that CCR includes property taxes and insurance in the amount of 2% of total cash expended (or 1.71% of TPI).

¹¹ Estimated costs for CO₂ T + S are based on a 100 km pipeline, aquifer depth of 2 km, CO₂ injectivity of 2500 tonne/day per well, and a 19%/yr capital charge rate [28].

Table 7

“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_0 , of a single train of a reference component of size S_0 by the relationship: $C = n^e C_0 [S/(n S_0)]^f$, 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.

Plant component	Scaling parameter	So	n	f	Co (M\$)	SV		SC		SE		QC	
						S	C (M\$)	S	C (M\$)	S	C (M\$)	S	C (M\$)
Coal + sorbent handling	AR coal, tonne/day	5447	1	0.67	40.4	5776	42.0	6167	43.9	6150	43.8	6167	43.9
Coal prep & feeding	AR coal, tonne/day	2464	2	0.67	101.6	5776	211.0	6167	220.4	6150	220.0	6167	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 compressor	Pure O ₂ , tonne/day	2035	2	0.50	106.7	4013	197.7	4284	204.3	4272	204.0	4284	204.3
Std. gasifier, SG coolers	AR coal, MW LHV	737.4	2	0.67	178.1	1730	369.8	1847	386.3	1842	385.6	–	–
Partial quench gasifier	AR coal, MW LHV	770.9	2	0.67	139.5	–	–	–	–	–	–	1847	293.8
LT heat recovery, saturator	AR coal, MW LHV	737.4	2	0.67	17.3	1730	36.0	1847	37.6	1842	37.5	1847	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	–	–	1847	6.1	1842	2.3*	1847	18.9
Gas cleanup BOP	AR coal, MW LHV	815.2	2	0.67	6.1	1730	11.8	1847	12.4	1842	12.3	1847	12.4
AGR (H ₂ S 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 compression & drying	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, auxiliaries	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)							1523		1747		1745		1647
Specific Total Plant Cost (\$/kW of electric power)							1847		2584		2520		2535

* Cost reduced by 16.5 M\$ because of reduction in 165 MW of heat exchangers.

Table 8

Levelized cost of electricity for each case.

Cost component, mid-2008 \$/MWh	SV	SC	SE	QC
Installed capital (at 15% of TPI)	41.7	58.8	56.9	57.2
O&M (at 4% of TPC per yr)	9.9	14.0	13.5	13.6
Coal (at 1.71 \$/GJ, HHV)	13.5	17.6	17.1	18.3
CO ₂ disposal (at 7.1 \$/tonne)	0.0	5.8	5.7	6.0
LCOE (no carbon price)	65.0	96.2	93.1	95.1
CO ₂ emissions (at 45.6 \$/tonne)	30.5	2.4	2.4	2.5
LCOE with CO ₂ price (46.4 \$/tonne)	95.5	98.5	95.5	97.6
Cost of avoided CO ₂ , \$/tonne of CO ₂	–	50.5	45.6	49.0

total plant cost and –3.1% levelized cost of electricity compared to the reference case).

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Appendix A

Table A1.

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 (bar)	38.5
H ₂ O/O ₂ molar ratio (steam as moderator)	0.266

Table A1 (continued)

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 rate (wt ratio)	0.3193
LP N ₂ for coal transport/coal mass flow rate (wt ratio)	0.1244
HP N ₂ into syngas/dried coal (wt ratio)	0.103
Maximum oxidant temperature (°C)	100
Syngas coolers: pinch point temp. diff. HT/LT coolers (K)	50/15
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 and 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 ₂ S, data from [1]:	
LT heat for stripping (MJ per kg of stripped H ₂ S)	44.7
H ₂ co-absorbed (%)	0.533
CO co-absorbed (%)	0.0044
Specific electricity consumption (kJ per kg of abs. CO ₂)	125
Sulfinol-M for removal of H ₂ S, data from [1]:	
LT heat for stripping (MJ per kg of stripped H ₂ S)	13.4
Specific electricity consumption (kJ per kg of syngas)	9.36
CO co-absorbed (%)	0.265
H ₂ co-absorbed (%)	0.268
Water–gas shift (WGS) unit	
HT/LT WGS reactor approach temperatures (K)	10/10
HT/LT WGS syngas input temperatures (°C)	250/250
CO ₂ compression and drying	
Specific electricity use (kJ per kg of CO ₂), data from [1]	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

Table A1 (continued)

Excess air	0.06
Compressor electrical \times mechanical efficiency	0.92
O ₂ compressor (3 ICs), 4 radial stages, average $\eta_{\text{polytropic}}$	0.845
LP N ₂ compressor (1 IC), 2 radial stages, average $\eta_{\text{polytropic}}$	0.858
HP N ₂ compressor (2 ICs), 3 radial stages, average $\eta_{\text{polytropic}}$	0.79
Dilution N ₂ compressor (axial machine, 1 IC), avg. $\eta_{\text{polytropic}}$	0.887
Intercooler exit temperature (°C)	45
<i>Power island</i>	
(2 GE 9FB GTs + 2 HRSGs + 1 steam turbine)	
HRSC, triple pressure levels with RH	
Approach temperature: Steam SH – Hot flue gas (K)	25
Min. pinch points/sub-cooling temp. diff. in HRSG (K)	8/3
Pumps: hydraulic efficiency	0.84
Pumps: electrical \times organic efficiency	0.9
Maximum steam pressure (for the HP level), bar	140 (170 [*])
Minimum steam pressure (for the LP level), bar	5 (3 [*])
Steam turbine mechanical efficiency	0.98
Steam turbine generator electrical efficiency	0.99
Economizers: pressure losses (%)	25
Superheater and reheater pressure losses (%)	8
Condenser pressure (bar)	0.05
<i>Other units</i>	
Sour water stripper, LP steam requirement, kJ/kg coal	84.0
Electricity for cooling components (% of rejected heat)	0.5

* “Aggressive” HRSC parameter assumptions.

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