Techno-economic Analysis of an Integrated Gasification Direct-Fired Supercritical CO_2 Power Cycle

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Abstract

This study describes recent systems analyses of coal-fueled, oxy-fired direct supercritical CO_2 (sCO_2) power cycles, which are inherently amenable to carbon capture and storage (CCS) processes. In this plant, coal is gasified in an entrained flow gasifier, cleaned of sulfur and particulate matter, and supplied to the sCO_2 cycle's oxy-combustor, where it burns with oxygen in a high pressure, heavily diluted sCO_2 environment. Following expansion of the combustion products through a turbine and recuperation of its thermal energy, water is condensed from the working fluid, which is recompressed for return to the combustor, with a portion of the CO_2 exhausted from the cycle for carbon capture and storage (CCS).

The conceptual designs for two coal-fueled direct-sCO_2 power plants are included in this study, one as a baseline sCO_2 plant, and another which includes improved thermal integration between the sCO_2 cycle and the gasifier. Parametric analyses are used to optimize the overall plant configuration and operating conditions, from which the capital cost and operating expenses of the plant are estimated, and the cost of electricity (COE) calculated. The study results yield plant thermal efficiencies of 37.7% (HHV) and 40.6% for the baseline and improved sCO_2 cases, including CCS. These efficiencies are a significant improvement on the 31.2% efficiency of the same gasifier in integrated gasification combined cycle (IGCC) plant configuration with CCS, and are also shown to be comparable to more advanced IGCC systems and other coal-fueled direct sCO_2 plants with CCS. Detailed economic analyses yield COE results of \$137.3/MWh and \$122.7/MWh for the baseline and thermally integrated sCO_2 plants, respectively, representing 10% and 20% improvements on the reference IGCC COE of \$152.6/MWh, with CCS. Recommendations are made for alternative gasification systems and modifications to the thermal integration between the gasifier and direct sCO_2 system, which should further improve plant performance relative to alternative systems.

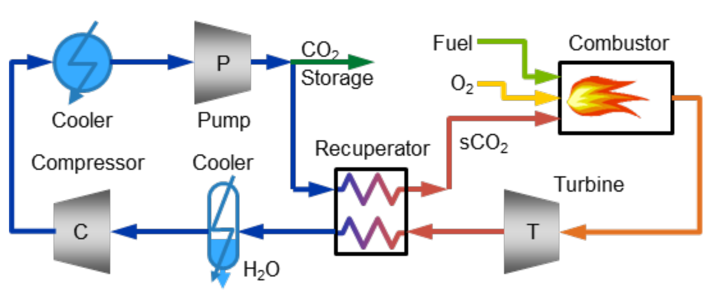
Keywords: supercritical CO_2 (sCO_2), direct sCO_2, coal gasification, integrated gasification combined cycle (IGCC), techno-economic analysis

1 Introduction

Supercritical carbon dioxide (sCO_2) power cycles have the potential for improved efficiency and cost relative to more conventional technologies, and are being explored as replacements for steam or gas turbine power cycles in many applications. sCO_2 cycles utilize high pressure CO_2 as the working fluid, typically above the CO_2 critical point of 31 °C and 7.37 MPa, which both reduces phase change heat transfer inefficiency and reduces compression power relative to gas-phase compression due to high sCO_2 density near the critical point. CO_2 is also inexpensive, abundant, non-toxic, and less corrosive than steam as a working fluid. Finally, turbomachinery sizes are smaller due to the high cycle pressures, and are potentially less expensive than their air- or steam-based counterparts.

While there are a variety of sCO_2 power cycles under consideration for a number of applications, the open, or direct sCO_2 cycle holds significant promise for combustion-based applications where carbon capture and storage (CCS) is desired. A simple schematic of the direct sCO_2 cycle is depicted in Figure 1. Oxygen and gaseous fuel are burned in a large CO_2 diluent stream, and the combustion products are expanded through a turbine to generate power. The turbine exhaust passes through a recuperator to transfer thermal energy to the recycle CO_2 stream returning to the combustor. After the recuperator, water is condensed from the working fluid, leaving a high purity CO_2 stream that is compressed and cooled for return to the cycle, with a high pressure exhaust stream available for CO_2 storage or utilization.

Figure 1 Direct sCO_2 Cycle Schematic



Work on the direct sCO_2 cycle was pioneered by Allam and colleagues, who have modeled both natural gas and syngas-fueled versions of the cycle [1] [2] [3]. Analysis of the natural gas-fueled plant has a reported net plant thermal efficiency of 53.2% higher heating value (HHV) [1], while the syngas-fired cycle, with an integral coal gasification train, yields a reported HHV efficiency of 49.7% [3], both with near 100% carbon capture. Allam and colleagues are commercializing this technology through 8 Rivers Capital, LLC, who is leading the construction of a 25 MWe natural gas-fueled demonstration plant in Laporte, Texas, slated to begin operation in 2017.

The promise of high plant efficiencies with an integral carbon capture process has led to additional studies of this cycle. Under slightly different assumptions from 8 Rivers Capital, Foster Wheeler's modeling of a natural gas-fueled direct sCO_2 plant yields a net thermal efficiency of 49.9% (HHV), with 90% carbon capture [4], and is shown to outperform other oxy-turbine power cycles under similar conditions. An Electric Power Research Institute (EPRI)-sponsored derivative analysis under different design basis assumptions yields a thermal efficiency of 48.4% [5]. Southwest Research Institute has reported modeled plant thermal efficiencies of 48.1% (HHV) with alternative natural gas-fired direct sCO_2 cycles [6]. EPRI has also studied a syngas-fired direct sCO_2 plant, with upstream coal gasification via a slagging, entrained flow gasifier [7]. This study yields a 39.6% plant thermal efficiency, with 99.2% CO_2 capture and an estimated cost of electricity (COE) at $133/MWh, though significant uncertainty was noted in the sCO_2 unit capital costs.

The following sections describe the technoeconomic analyses undertaken by the U.S. Department of Energy's (DOE) National Energy Technology Laboratory (NETL) on a coal-fueled direct sCO_2 plant integrated with a coal gasifier and CCS. The main objective of this study is to provide an impartial assessment of the coal-fueled, direct sCO_2 power plant concept based on well-documented and well-vetted methodology that allows more accurate and reliable comparisons of this concept with both conventional technology and other advanced technology options. In accomplishing these objectives, this study validates the claims that sCO_2 power plants can be more efficient and have more favorable economics than power plants based on conventional technology, and does so with a level of detail and transparency that has not been reported in the direct sCO_2 power cycle literature to date, effectively providing a realistic expectation of the eventual cost and efficiency of direct sCO_2 power plants.

Preliminary results of this study have been presented in other venues [8] [9], and have culminated in a baseline gasifier/direct sCO_2 plant design detailed in Reference [10], with a thermal efficiency of 37.7% (HHV) and COE of $137.3/MWh, excluding CO_2 transport and storage (T&S) costs. This paper discusses improvements to this plant design though innovation in both the approach to thermal integration, and the cost-based assessment of whether a candidate heat integration option is economically viable. These improvements have resulted in a "Case 2" plant design with a significant increase in thermal efficiency (40.6% HHV) and economic performance ($122.7/MWh), with 99.4% carbon capture. Except for integrated gasification fuel cell plants, this is among the highest efficiencies reported in NETL's many studies of bituminous coal-fueled power plants with CCS, with significant potential for further efficiency improvements identified in Section 4.3.

2 Techno-economic Modeling Methods

2.1 Design Basis Assumptions

The design bases from NETL's Bituminous Baseline Study [11] and Quality Guidelines for Energy System Studies (QGESS) series [12] [13] [14] [15] [16] [17] [18] were adopted so that the results from this study are consistent with established results and comparable to existing studies carried out by NETL. Additional details regarding the design basis assumptions can be found in Reference [10].

All plants in this study are assumed to be located at a generic plant site in the midwestern United States at sea level with an ambient dry bulb temperature of 15 °C and 60% relative humidity [17].

Coal properties from NETL's QGESS are shown in Table 1 for the bituminous coal used in this study. [13]

Table 1 Bituminous design coal analysis

Rank	**Bituminous**		
Seam	Illinois No. 6 (Herrin)		
Sample location	Franklin Co., Illinois		
Proximate analysis	**Dry basis, wt%**	**As fed, wt%**	**As received, wt%**
Moisture	0.0	5.00	11.12
Ash	10.91	10.36	9.70
Volatile matter	39.37	37.41	34.99
Fixed carbon	49.72	47.23	44.19
Ultimate analysis	**Dry basis, wt%**	**As fed, wt%**	**As received, wt%**
Carbon	71.72	68.14	63.75
Hydrogen	5.06	4.81	4.50
Nitrogen	1.41	1.34	1.25
Sulfur	2.82	2.68	2.51
Chlorine	0.33	0.31	0.29
Ash	10.91	10.36	9.70
Moisture	0.00	5.00	11.12
Oxygen	7.75	7.36	6.88
Heating value	**Dry basis**	**As fed**	**As received**
HHV, kJ/kg	30,506	28,981	27,113
LHV, kJ/kg	29,544	28,019	26,151

The direct sCO_2 power cycle produces a high pressure exhaust stream that is typically greater than 98 mol% CO_2. In most direct sCO_2 systems studies, this stream is "capture-ready" and sent to carbon storage as-is. [7] [3] In the present study, this stream is first sent through a CO_2 purification unit (CPU) to eliminate impurities and produce CO_2 at pipeline purity specifications as set forth in the NETL QGESS. [16] This increases the plant cost and reduces efficiency somewhat, but maintains consistency with other NETL studies which meet this CO_2 purity specification.

2.2 Modeling Methodology

The modeling methodology includes performing steady-state simulations of the various technologies using the ASPEN Plus® (Aspen) modeling program. For the gasification island, air separation units (ASU), syngas cleaning, and CPU sections, the Peng-Robinson equation of state (EOS) is used [11], while the

Lee-Kessler-Plöcker EOS, is used for the sCO_2 power cycle. [19] Component performance and process limitations are considered based upon published reports, information obtained from vendors and users of the technology, performance data from design/build utility projects, and/or best engineering judgment. The resulting mass and energy balance data from the Aspen model is used to size major pieces of equipment, which form the basis for component cost estimates. Since utility scale equipment for the sCO_2 power cycle has never been built, more reliance was placed on vendor estimates, including order of magnitude estimates, for these components.

2.3 Cost Estimating Methodology

Plant capital costs in this study are estimated using NETL's quality guidelines, and are reported at a much finer level of detail than has been presented in the sCO_2 power cycle literature to date. [12] Capital costs are reported at two levels: bare erected cost (BEC) and total plant cost (TPC), which are overnight costs expressed in 2011 base-year dollars. Cost estimates for conventional unit operations have an expected accuracy range of -15/+30%. [11] Given that sCO_2 power cycle development is in its infancy relative to more commercial cycles, the cost estimate developed in this study is best described as an Association for the Advancement of Cost Engineering International (AACE) Class 4 Feasibility Study, with an estimated accuracy range of -15/+50% [20].

Capital and operation and maintenance (O&M) costs are estimated using cost algorithms consistent with a conceptual level scope definition, which incorporates the following: 1) NETL Bituminous Baseline Study costs (estimated by WorleyParsons), 2) literature and/or vendor supplied costs, and 3) R&D target costs for advanced technologies. Capital cost algorithms are reflective of "nth-of-a-kind" (NOAK) equipment and plant costs for mature technologies, and are free of component development costs embedded in "first-of-a-kind" (FOAK) plant costs. [12]

Process and project contingencies are included in cost estimates to account for expected but undefined costs that are omitted or unforeseen due to a lack of complete project definition and engineering. Process contingencies compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology. [14] Lower process contingency costs are used in this study for sCO_2 specific components, which is more reflective of an NOAK cost estimate.

Reported O&M costs are divided into two categories; fixed O&M costs that are independent of plant operation hours (e.g., labor, overhead, etc.), and variable O&M costs that are proportional to power generation (e.g., consumables, waste disposal, maintenance materials). The variable O&M and fuel costs are multiplied by an assumed capacity factor of 80%, consistent with gasification-based systems, to arrive at the actual expended annual cost.

The COE is reported on a $/MWh basis and consists of contributions from the O&M costs (fixed, variable, and fuel), CO_2 T&S costs, and the annualized capital cost over the assumed 30 year lifetime of the plant. Details on the estimated T&S costs are provided in References [15] and [18], while additional details on the cost estimating methodology and other economic assumptions are provided in References [10] and [14].

2.4 sCO_2 Component Cost Estimates

New cost scaling algorithms were derived for the components in the sCO_2 power cycle, which have not been reported in other direct sCO_2 system studies to date. Where possible, sCO_2 cycle component costs are estimated as NOAK costs to maintain a consistent basis for comparison to analyses of mature plants, such as NETL's Bituminous Baseline studies [11], though the estimates themselves have a high degree of uncertainty, as utility-scale direct sCO_2 plants have not yet been commercialized. The following sub-sections describe the approaches used for each of these components [10].

2.4.1 *CO_2 Oxy-Turbine*

The approach taken to costing the sCO_2 turbine is based on Toshiba's design, which employs inner and outer casings to contain the high system pressure, similar to the design of a high-pressure (HP) steam turbine [21]. In this approach, the outer casing can contain most of the pressure using conventional low temperature materials, with the more expensive high temperature and pressure materials reserved for the inner casing, which is actively cooled. The cost of the outer casing is based on known costs for a similarly-sized HP steam turbine, and amounts to about $4M in 2011 dollars [22].

For cost estimating purposes, the oxy-combustor, CO_2 expander, and inner casing, as well as the bearings, seals, and associated equipment, are considered analogous to a conventional gas turbine without the compressor. The sCO_2 unit size is based on the volumetric throughput of the working fluid, and assumed to be configured as a split-flow turbine, similar to HP steam turbines, to reduce or eliminate the thrust bearing requirements. A Trent 60 gas turbine uses roughly half of the total volumetric throughput of the sCO_2 turbine at a comparable firing temperatures, thus the cost of two Trent 60s, one for each flowpath of the split-flow turbine, is used as the base capital cost for the sCO_2 turbine. The BEC for these turbines is obtained from the Gas Turbine World Handbook [23], and an 18% deduction was applied for the cost of the unneeded compressors, based on gas turbine manufacturer feedback. The cost of the outer casing is added to this number to arrive at the total BEC cost of a mature sCO_2 oxy-turbine. In addition, consistent with the syngas-fired turbine in the reference plant, a process contingency of 5% of the BEC is applied to this sub-account due to the unique operating conditions of the combustor and turbine. For sensitivity analyses, the capital cost estimate is adjusted using a power law formula based on the net power output and an exponent of 0.7332.

2.4.2 sCO_2 Recuperators and Coolers

Equipment costs of the CO_2 recuperators are based on a report by Aerojet Rocketdyne, which details vendor-supplied recuperator estimates for a commercial-scale indirect sCO_2 plant. [24] The estimates employ modular recuperator designs, where low temperature modules are sized for 3 MW thermal duty each, and high temperature modules are sized for 8 MW each. Adjusting for variations in the log mean temperature differences and operating pressures relative to the reference recuperator cases, a baseline LTR cost of $0.294/(W/K) is derived. For the HTR, when the hot side fluid is at or below 600 °C, the HTR specific cost is $0.253/(W/K), and for temperatures above 600 °C, a material cost adjustment yields a HTR specific cost of $1.318/(W/K), with units assumed to be installed in series. The lower specific cost of the low temperature HTR modules, relative to the LTR modules, can be attributed to reduced distribution/interconnection piping requirements associated with the larger duty per module, as well as potential design differences required to meet the pressure drop specifications in the study. Cost estimates for the CO_2 pre-cooler and the CO_2 condenser are based on the LTR cost algorithm.

2.4.3 Compressors and Pumps

The cost estimates for the CO_2 pre-compressor and the CO_2 pump are scaled based on vendor data for a utility scale recompression Brayton sCO_2 cycle. The scaling algorithm divides the equipment cost into three factors dependent on required power (60%), inlet volumetric flow rate (20%), and inlet temperature (20%).

Cost estimates for the O_2 and syngas compressors used in the sCO_2 cycle are adjusted to 2011 dollars from a cost algorithm based on the power requirement of the compressor. [25]

2.4.4 CO_2 System Piping

The pipes that transport CO_2 working fluid can be very expensive due to the large flowrate of CO_2 and the elevated temperature and pressure of the fluid in some portions of the cycle. The four pipe lengths with the most severe service are the pipe from the CO_2 turbine to the HTR, the combined syngas and recycle CO_2 pipe to the oxy-combustor, the recycle CO_2 pipe from the HTR to the syngas cooler, and the syngas pipe from the syngas cooler. These pipe lengths are assumed to be 30.5 m (100 feet) except for the pipe run from the CO_2 turbine to the HTR, which is assumed to be 7.6 m (25 feet). The pipe inside diameter calculation is based on the actual working fluid volumetric flow rate and an assumed fluid velocity of 45.7 m/sec (150 ft/sec). Pipe thickness is calculated from ASME standards for a variety of materials, and the material capable of meeting the service temperature and pressure with the lowest estimated pipe cost is selected. [26] A 60% installation factor is assumed.

2.4.5 High Temperature Syngas Coolers

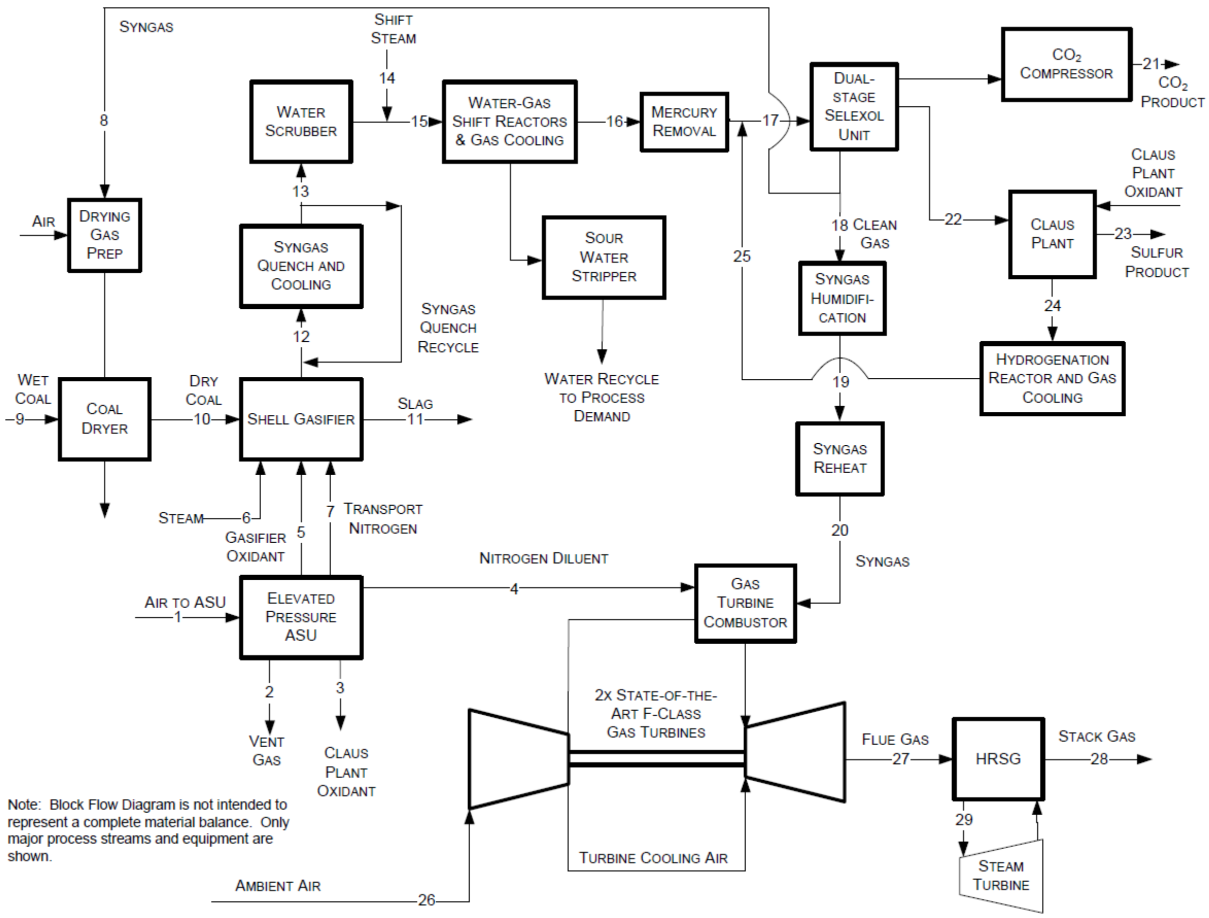
The high temperature syngas coolers are scaled from the estimate for the reference IGCC plant syngas coolers based on the difference in heat duties. In addition, an adjustment is made to the syngas cooler cost algorithm to account for differences in the cold side fluid between steam in the IGCC case and CO_2 in the present study. The portion of the syngas cooler used for heating sCO_2 accounts for 27% of the overall syngas cooler heat duty in the baseline sCO_2 plant design.

2.5 Reference Plant

A simplified block flow diagram for an IGCC process based on the Shell gasifier with carbon capture is shown in Figure 2. Since they use the same gasifier, this is used as a reference case for comparison to the sCO_2 plant, and is described in the Bituminous Baseline Study Rev 2b as Case B1B. [11]

The Shell gasifier is an entrained-flow, slagging gasifier that is dry-fed with coal. The coal feed system includes a pressurized lock hopper that requires dried coal at 5% moisture for proper operation. The IGCC plant utilizes an elevated pressure cryogenic ASU designed to produce 95% purity oxygen and high pressure nitrogen for use as a fuel diluent in the turbine. [11]

Figure 2 IGCC reference plant block flow diagram



In the IGCC plant, decarbonization requires water gas shift reactors, which are located downstream of the syngas scrubber. The acid gas removal is a two-stage Selexol process that removes both H_2S and CO_2 as separate product streams. Finally, the IGCC plant uses nitrogen as the transport gas for the dry feed lock hopper system, whereas the sCO_2 plant uses CO_2 to improve sCO_2 purity and, hence, cycle performance. [7]

The IGCC plant's power island uses a combined cycle with two F-frame gas turbines operated on hydrogen-rich syngas and a single sub-critical Rankine bottoming cycle. In the IGCC plant, there is further process integration with the use of byproduct nitrogen from the ASU as fuel gas diluent for power augmentation and NOx control in the gas turbine. [11]

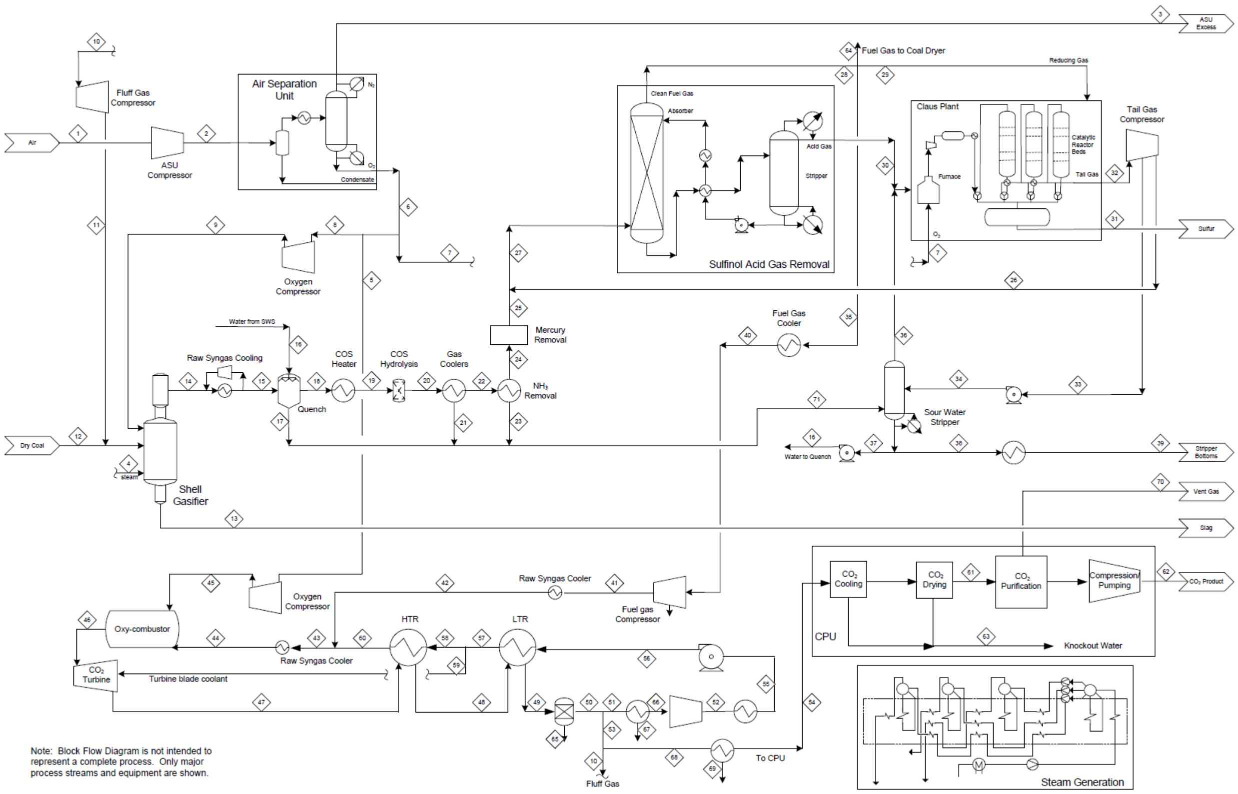
3 Integrated Gasification-Direct sCO_2 Plant Conceptual Designs

The Shell gasifier was selected for integration with the direct-sCO_2 cycle based on its commercial availability, syngas heat recovery capability, and dry coal feeding, which results in a high cold gas efficiency. Aspen models for the Shell gasifier and its auxiliary equipment are modified from NETL's Bituminous Baseline IGCC studies for use in this study [11]. The high pressure ASU is replaced with a low pressure ASU, since nitrogen at high pressure is not required, as it is for the gas turbine in an IGCC plant. Further, the ASU is designed to produce 99.5% purity oxygen to minimize argon and nitrogen contaminants in the sCO_2 cycle. Other systems studies have shown that the resulting reduction in CO_2 compression power due to higher sCO_2 purity more than offsets the increase in ASU power required to deliver higher purity oxygen to the cycle. [7] [4] In the same vein, a change to CO_2 as the coal carrier gas is implemented per the EPRI study's recommendations [7]. The IGCC syngas cleanup processes for ammonia, mercury, and sulfur removal are retained, but since CO_2 capture occurs in the sCO_2 cycle, the water-gas shift reactor is eliminated, and the double-stage Selexol process is replaced with a single-stage Sulfinol process. An alternative syngas cleanup method involving sulfur and nitrous oxide removal within the sCO_2 cycle could be considered to improve plant efficiency; however, this process is still under development, and a more conventional syngas cleanup method was chosen to improve prospects for near term deployment of the plant [1] [3]. Finally, the CPU is retained to ensure the production of pipeline-quality CO_2 for enhanced oil recovery or other purposes.

3.1 Baseline Plant Configuration

The baseline gasification-based sCO_2 power plant process flow diagram is shown in Figure 3. A detailed stream table accompanying Figure 3 for the overall sCO_2 baseline plant model can be found in the Supplemental Information.

Figure 3 Process flow diagram for coal-fired direct-fired sCO_2 power plant



The choice for the baseline direct sCO_2 power cycle configuration is similar to that in other studies [1] [4], with a few modifications. The syngas from the gasification train is compressed and preheated in a syngas cooler prior mixing with CO_2 exiting the sCO_2 cycles' high temperature recuperator. The mixture undergoes additional heating to 732 °C in another stage of the syngas cooler, and is then injected into the direct sCO_2 cycle for combustion. No oxygen preheating is considered. Steam is also raised in the gasification train to satisfy process steam requirements of the ASU, Sulfinol unit, and other processes, though unlike the EPRI study, additional steam power is not generated, opting instead to transfer that thermal energy to the sCO_2 cycle for power generation. Finally, the sCO_2 plant model includes an empirical turbine blade cooling model based on Reference [4].

Oxygen for the gasifier, sCO_2 oxy-combustor, and Claus plant comes from a low pressure cryogenic ASU. Oxygen for the sCO_2 oxy-combustor is compressed in a four-stage compressor with three intercoolers at an assumed isentropic efficiency of 85%. Intercooler stages assume an exit temperature 35 °C with water knock-out, and 69 kPa (10 psi) pressure drop per stage, consistent with intercooler heat recovery using gaseous process fluids.

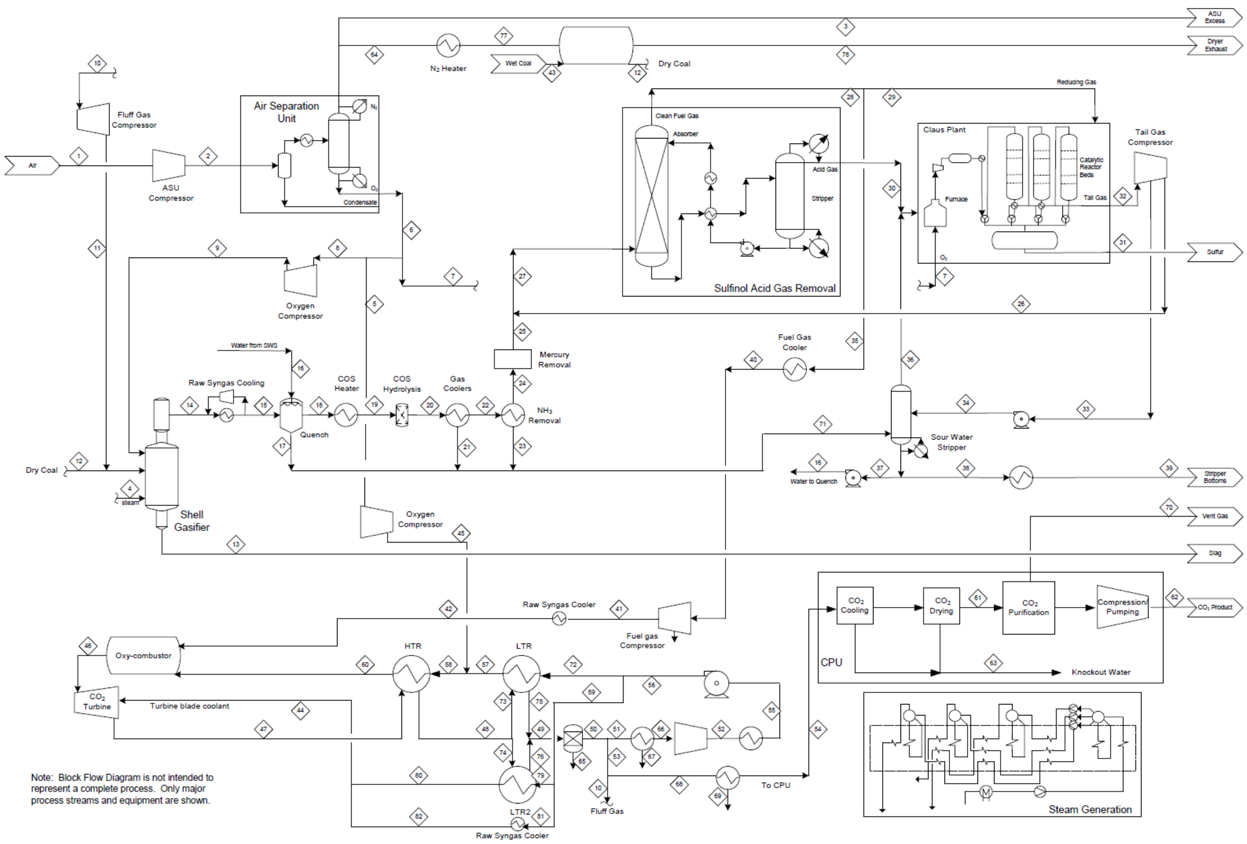
Desulfurization is done using a combination of Sulfinol and a Claus/Shell Claus Off-gas Treating unit. [11] The cleaned syngas leaving the Sulfinol unit is compressed to the combustor pressure in a two-stage compressor with a single intercooler and an assumed isentropic efficiency of 85%. The intercooler exit temperature is 35 °C (95 °F) with water knock-out, and assumes a 69 kPa (10 psi) pressure drop. Compressed syngas is preheated to 732 °C (1350 °F) in the syngas cooler, and enters the sCO_2 cycle combustor along with oxygen and recycled working fluid.

The effluent from the combustor enters the CO_2 turbine and is expanded to a pressure that cools the working fluid sufficiently to enter the recuperator. An upper limit of 760 °C (1400 °F) was chosen for this location based on the high temperature and pressure limits of nickel-based alloys, which represent a major constraint on the system design. [2] The working fluid is cooled in the recuperator by heating the recycled CO_2. The cooled working fluid exiting the cold end of the recuperator undergoes further cooling to condense out water in the pre-cooler, and then undergoes recompression. Most of this recompressed stream is recycled to the recuperator and then to the oxy-combustor, though a portion of the flow is sent to the turbine for cooling duty after it is preheated to 400 °C in the LTR. Most of the purged portion of the recompressed working fluid enters a CO_2 purification unit and that product stream is sent for storage. A portion of the purge stream is recycled to the gasifier lock hoppers as a transport gas, which is required to maintain sCO_2 purity and low compression power requirements. [7]

3.2 Case 2 Plant Configuration

While the baseline sCO_2 system described above and in Reference [10] provides significant cost and efficiency improvements over the reference IGCC plant, several areas were identified for further enhancement of the gasification-sCO_2 plant design, culminating in an improved sCO_2 power plant designated as Case 2. Figure 4 shows the process flow diagram for Case 2 which is discussed below. A stream table accompanying Fig. 4 is available in the Supplementary Information.

Figure 4 Process flow diagram for the improved sCO_2 power plant



First, the baseline plant configuration burns about 1.2% of the generated syngas in a separate combustor for coal drying duty, as implemented in the IGCC reference plant [11]. A model change was implemented to spare the high-value syngas for sCO_2 oxy-combustion only, and to use the ASU byproduct nitrogen stream, preheated by low grade heat elsewhere in the cycle, for coal drying service.

Second, no preheating of the oxygen flow to the oxy-combustor is performed in the baseline sCO_2 system, though doing so would improve plant performance by reducing the fuel required to reach the desired turbine inlet temperature. This is implemented in the revised model by reducing the amount of oxygen compressor intercooling used to result in a maximum safe oxygen temperature limit of 300 °C at the exit of the compressor, which is then mixed with high pressure sCO_2 exiting the LTR at 400 °C prior to entering the HTR for final preheating.

Third, the final high temperature preheating of sCO_2 in the syngas cooler provides a significant efficiency benefit, but at a significant capital cost penalty. Eliminating this heat exchanger and the associated piping runs to and from the sCO_2 cycle results in TPC reduction of about $173 million, about 8.4% of the overall plant capital cost, as shown in Section 4.2.1. This also reduces the overall plant efficiency by about 2 percentage points, however, this can be recovered by improving the recuperation heat balance in the sCO_2 cycle such that higher sCO_2 temperatures are attained at the exit of the HTR. Thermal duty is thus transferred from the large and expensive syngas cooler, which must contend with particulate-laden syngas flows, to the compact sCO_2 recuperators, offering a more cost-effective sCO_2 preheating solution.

Fourth, to improve upon the sCO_2 recuperator balance, the sCO_2 cycle can be thermally integrated with other heat sources in the plant. Similar to thermal integration strategies proposed in other studies [1] [27] [4], in the revised configuration, a slip stream of cold sCO_2 is removed for heating elsewhere in the plant, in parallel with the LTR, to reduce the cold side thermal capacitance in the LTR. As will be seen below, this allows for reduced temperature differentials throughout the sCO_2 recuperator system, with the flow of the parallel slip stream adjusted to achieve a 10 °C approach temperature at the hot end of the HTR. Doing so effectively increases the sCO_2 preheating capability of the HTR, allows for the elimination of the cost of the final sCO_2 preheating in the syngas cooler, and negates the efficiency loss incurred by removing this heat exchanger alone. This has the overall effect of simultaneously increasing recuperator duty and reducing the temperature driving forces within the recuperators, resulting in larger and more expensive recuperators, partially offsetting the elimination of the high temperature syngas cooler's cost.

Finally, to increase the thermal energy available for integration with the revised sCO_2 cycle, an alternate ASU model is used to replace a compressor driven by medium pressure steam with an electrically-driven compressor, freeing up the process steam thermal duty for use in the sCO_2 cycle. The trade-off for the low steam duty is an 8% increase in the ASU specific power requirement.

3.3 sCO_2 Cycle Modeling

Table 2 shows the parameters that define the sCO_2 power cycle assumptions, which apply to both the baseline and improved Case 2 sCO_2 plant models. Most of the values were selected based on the results of many sensitivity analyses and represent a reasonable balance between the goal of maximizing efficiency and the goal of minimizing cost. [8] [10] [28]

Table 2: sCO_2 power cycle assumptions

Section	Parameter	Assumed Value
Combustor	Excess O_2	1%
	Pressure drop	689 kPa (100 psi)
Expander	Inlet temp	1204 °C (2200 °F)
	Inlet pressure	30.0 MPa
	Isentropic efficiency	0.927
	Blade Cooling	15.0%
Recuperator	Pressure Drop	68.9 kPa (10 psi) per side
	Max temp	750 °C (1382 °F)
	Min T_{app}	10 °C (18 °F)
CO_2 cooler	Cooler/condenser	26.7 °C (80 °F)
	Cooling source	Cooling tower
CO_2 pressurization	Compressor outlet pressure	7.4 MPa
	Pump outlet pressure	30.8 MPa
	Isentropic efficiency	0.85
	Intercooler temperature	26.7 °C (80 °F)
	Intercooler pressure drop	14 kPa (2 psia)

4 Results and Discussion

4.1 Performance Results

Table 3 shows the overall performance summary for the modeled cases in this study. The gross power output for the sCO_2 cycles includes the auxiliary power requirement for the CO_2 pre-compressor (70.5 MWe and 75.6 MWe for the baseline case and Case 2, respectively) and the CO_2 pump (67.3 MWe and 72.1 MWe for the baseline case and Case 2, respectively), and forms the basis for the sCO_2 power cycle efficiency of nearly 62% in each case.

Table 3 Overall performance summary

Parameter	**IGCC Reference Case B1B** [11]	**sCO_2 Baseline Case**	**sCO_2 Case 2**
Coal Flowrate (kg/hr)	211,040	198,059	198,059
Coal Thermal Input (MW_{th}, HHV)	1,591	1,493	1,493
Oxygen Flowrate (kg/hr)	160,514	391,227	394,234
Gross Power Output (MW_e)	673	777	828
Net Auxiliary Load (MW_e)	177	215	222
Net Plant Power Output (MW_e)	497	562	606
Net Plant Efficiency (HHV, %)	31.2	37.7	40.6
sCO_2 Flowrate (kg/hr)	--	7,243,859	7,734,832
sCO_2 Cycle Thermal Input (MW_{th}, HHV)	--	1260	1337
sCO_2 power cycle efficiency (%)	--	61.7	61.9
Carbon capture fraction (%)	90.0	97.6	99.4
Captured CO_2 purity (mol% CO_2)	99.4	99.8	99.8
Water Withdrawal ([m^3/min]/MW_{net})	0.043	0.036	0.033

The estimated plant efficiency of the baseline sCO_2 case is 37.7% (HHV basis), and that of the thermally-integrated Case 2 plant is 40.6%. These efficiencies are quite high for coal-fired power plants with CCS, and are 6.5 and 9.4 percentage points higher than the reference IGCC plant, respectively. The Case 2 plant generates 8% more net power than the baseline sCO_2 plant using the same coal feed due to increased thermal input to the sCO_2 cycle, and requires 3% more auxiliary power. The oxygen flow rate for Case 2 is 0.8% higher than for the baseline case because of the extra syngas available to the oxy-combustor from the elimination of the syngas coal dryer. Because of the improved heat recovery in the sCO_2 power cycle, as discussed in Section 4.1.4, the net sCO_2 cycle output increases 8% for Case 2.

Carbon capture in the sCO_2 baseline plant is a little less than 98%, compared to 90% in the reference IGCC plant. The major source of uncaptured carbon in the baseline sCO_2 plant is the syngas feed to the coal dryer. Eliminating this use of syngas allows the Case 2 plant to capture 99.4% of the input fuel carbon, with the remainder being exhausted in the CPU vent gas.

4.1.1 Auxiliary Power Requirements

Table 4 shows the breakdown of auxiliary power requirements in the modeled plants. The ASU's main air compressor draws significantly more power in the sCO_2 plants relative to the reference IGCC plant, due to the additional oxygen requirements (Table 3), while the sCO_2 CPU compression requirements are smaller due to the high pressure CO_2 at the CPU inlet for these cases. Overall compression power needs are higher in the sCO_2 plants, though their single-stage Sulfinol units (acid gas removal (AGR)) require less power than the two-stage process used in the reference plant to capture carbon.

Table 4 Auxiliary power requirements, kWe

Auxiliary Load, *kWe*	**IGCC Reference Case B1B** [11]	**sCO_2 Baseline Case**	**sCO_2 Case 2**
Coal Handling	-460	-447	-447
Coal Milling	-2,170	-2,037	-2,037
Coal Dryer Air Blower	N/A	-140	-1,468
Slag Handling	-550	-492	-492
Air Separation Unit Auxiliaries	-1,000	-1,000	-1,000
Air Separation Unit Main Air Compressor	-59,740	-79,001	-88,234
Gasifier Oxygen Compressor	-9,460	-20,057	-18,345
sCO_2 Oxygen Compressor	N/A	-44,120	-41,783
Nitrogen Compressors	-32,910	N/A	N/A
Fuel Gas Compressor	N/A	-35,240	-35,674
CO_2 Fluff Gas Compressor	N/A	-413	-413
Feedwater Pumps	-3,500	-92	-93
Syngas Recycle Compressor	-790	-869	-869
CPU CO_2 Compression	-30,210	-15,934	-16,136

Acid Gas Removal	-18,650	-457	-457
Combustion Turbine Auxiliaries	-1,000	-1,000	-1,000
Steam Turbine Auxiliaries	-100	N/A	N/A
Circulating Water Pumps	-4,380	-3,984	-3,954
Miscellaneous Water Pumps	-1,760	-10	-9
Cooling Tower Fans	-2,270	-2,579	-2,559
Claus Plant/TGTU Auxiliaries	-250	-250	-250
Claus Plant TG Recycle Compressor	-1,830	-595	-594
Miscellaneous Balance of Plant	-3,000	-3,000	-3,000
Transformer Losses	-2,630	-2,943	-3,115
Total Auxiliaries	**-176,660**	**-214,659**	**-221,928**

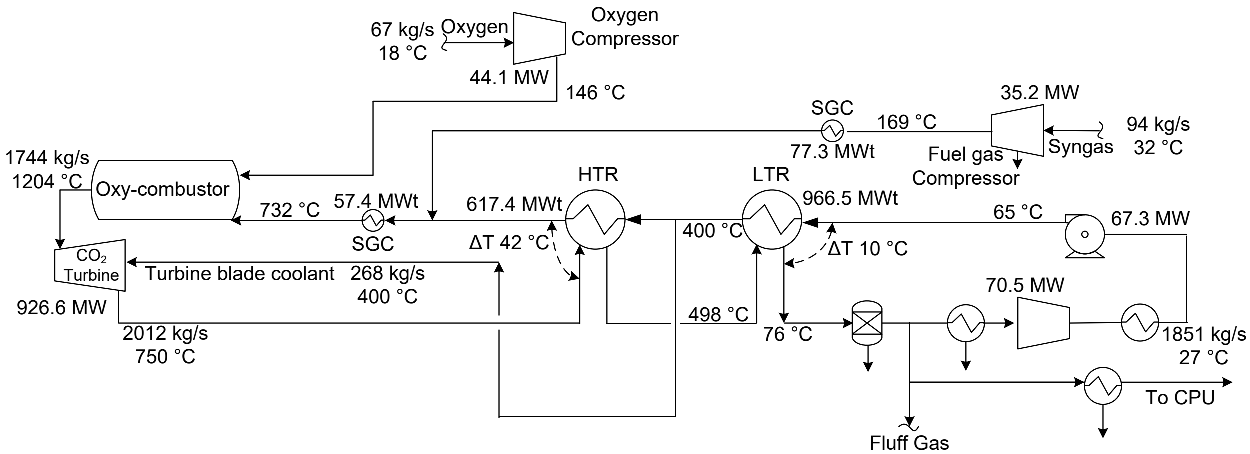
The most significant differences in the auxiliary power requirements for Case 2 compared to the baseline case are the higher coal dryer air blower power (due to the change from a syngas-fired coal dryer to a hot nitrogen dryer), the higher ASU main air compressor power (due to the higher specific power requirement in the Case 2 ASU and the increased oxygen demand in the oxy-combustor from the elimination of the syngas-fired coal dryer), the lower gasifier oxygen and sCO_2 oxygen compressor power (due to the higher oxygen product pressure in the Case 2 ASU), and the higher CPU compressor power (due to the increased carbon capture fraction in Case 2).

4.1.2 sCO_2 Cycle Performance

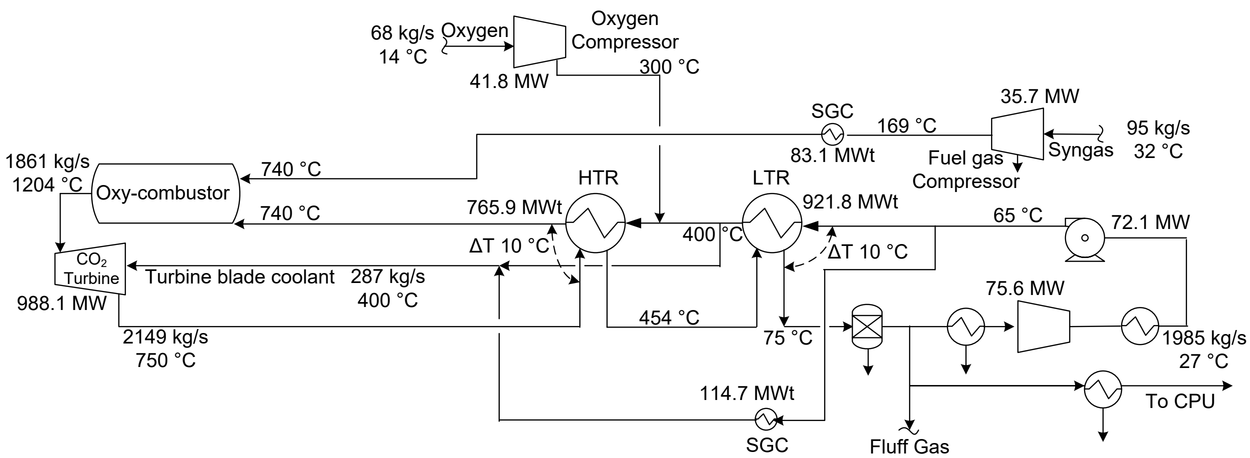
Key performance data for the sCO_2 power cycles are shown in Fig. 5a for the sCO_2 baseline, and in Fig. 5b for the Case 2 plant. Compared to the baseline case, the Case 2 sCO_2 power cycle has a larger CO_2 flow rate, larger net turbine output, and larger total recuperator duties reflecting the higher net power cycle output. Note the shift in recuperator heat duty from the LTR to the HTR that occurs in Case 2, and the large thermal input to the sCO_2 turbine cooling stream that results from thermal integration with the gasifier train, which are discussed in more detail in the following sections.

Figure 5 sCO_2 power cycle performance summary for (a) the baseline case, and (b) Case 2

(a)



(b)

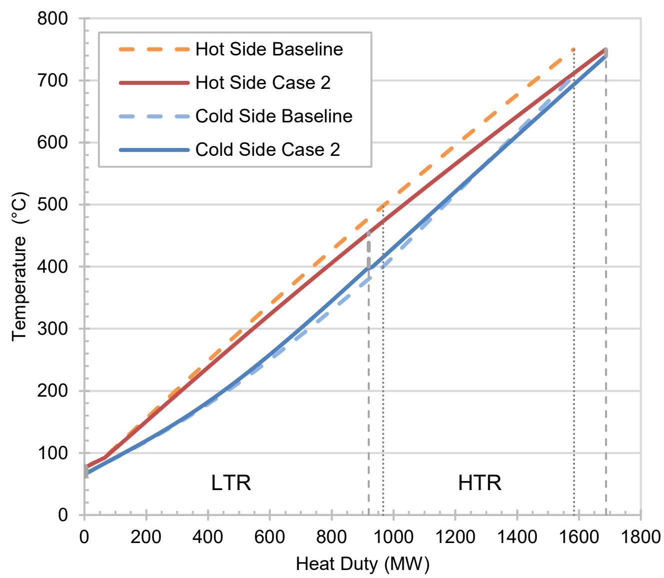


4.1.3 Recuperator Performance

Figure 6 is a temperature-heat duty (T-Q) diagram for the recuperators in each of the sCO_2 cycles, showing both the hot sides (red line) and cold sides (blue line). The dashed vertical lines denote the break points between the LTR and HTR for each case.

At the cold end of the LTR for each cycle, the temperature approach is 10.3 °C, which is slightly higher than the minimum design specification of 10 °C in Table 2. It remains almost constant, rising slightly through the first 6-7% by duty of the LTR and then begins to noticeably increase. The temperature at which the change in slope occurs on the hot side of the recuperator corresponds to the onset of water condensation as the working fluid is cooled through the low pressure side of the LTR. As thermal duty increases further, the maximum temperature approach of 98 °C occurs at the hot end of the LTR for the baseline sCO_2 recuperator system due to an excess thermal capacity of the sCO_2 fluid on the LTR's cold side. Once the turbine cooling flow is bled off after the LTR, the cold side thermal capacity is reduced, and the temperature approach reduces to 44 °C at the HTR's hot end.

Figure 6 Temperature-heat transfer diagrams for the sCO_2 cycle recuperators

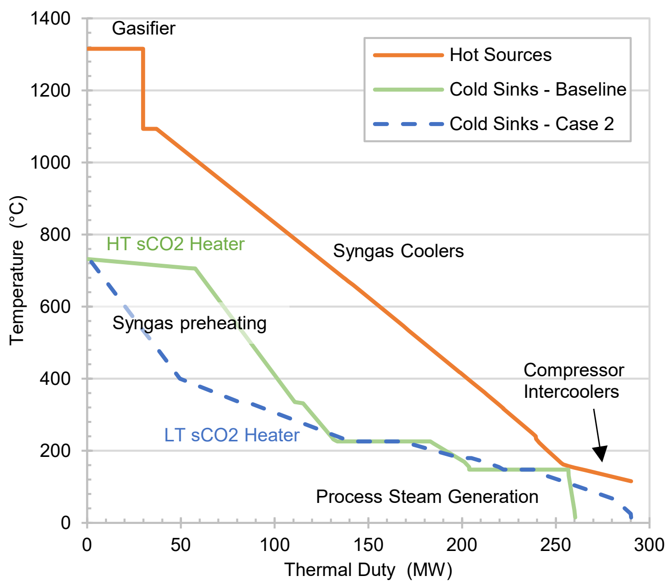


Compared to the baseline sCO_2 case, the HTR duty in Case 2 increases, the LTR duty decreases, and the average temperature difference between the hot side and cold side is reduced. The temperature approach at the hot end of the LTR has been reduced from 98 °C in the baseline case to 54 °C in Case 2. The temperature approach at the hot end of the Case 2 HTR is 10 °C compared to 43 °C in the baseline case, and increases the recycle sCO_2 delivery temperature to the oxy-combustor from 732 °C to 740 °C, including preheating of the oxygen stream, all of which improves the efficiency of Case 2. The overall increase in recuperation duty for Case 2 is consistent with the 6.7% increase in sCO_2 cycle mass flow relative to the baseline system, per Table 3.

4.1.4 *Plant Thermal Integration*

The heat integration schemes for the sCO_2 plants are derived based on a pinch analysis. The hot streams available in the process include the high and low temperature syngas coolers, the gasifier, the Claus unit, the sour water stripper bottoms cooler, the oxygen compressor intercooler, and the fuel gas intercooler. The primary cold sinks are the steam plant and the CO_2 and syngas feeds to the combustor. The hot and cold streams used to perform the pinch analysis are tabulated in the Supplemental Information. No attempts were made to match specific hot and cold streams in the analysis, although some potential pairings are evident in plotting the heat integration scheme on a T-Q diagram, shown in Figure 7.

Figure 7 Temperature-heat transfer diagrams for process heat integration in the sCO_2 cases



With a consistent coal flow rate between the two sCO_2 cases, the heat sources available for integration in the gasification train are also the same, with differences arising in the cold sinks requiring heat between the two cases. The heat integration scheme for the Case 2 plant is very similar to that used for the baseline case, except that the final high temperature (HT) sCO_2 preheating is replaced with low temperature sCO_2 preheating in parallel to the LTR (see Fig. 4), and the steam duty for Case 2 is lower due to the change from a steam-driven to electric driven ASU compressor, as discussed in Section 3.2.

This latter change also allows for nitrogen preheating with heat from the gasification train for use in coal drying in Case 2.

In the baseline sCO_2 case, 260 MW of heat is recovered from heat sources in the plant, though this is limited by the pinch point at 148 °C in the Low Pressure (LP) steam evaporator, as seen in Figure 7. An extra 30 MW of heat is recovered using the improved thermal integration scheme in Case 2, for a total of 290 MW. The revised scheme eliminates the pinch point problem in the baseline case, and allows for heat recovery in the compressor intercoolers down to 116 °C, compared to a minimum recoverable temperature of 151 °C in the baseline case. Per Table 3, this helps increase the thermal input to the Case 2 sCO_2 cycle, increasing its net power output, and hence its thermal efficiency.

4.2 Economic Results

4.2.1 Plant Capital Costs

Table 5 compares the account level TPC between the sCO_2 plants and the reference IGCC plant. Specific component costs for each account are detailed in cost tables found in the Supplemental Information. As an example, the Power Cycle & Accessories account is presented in Table 6 for the sCO_2 cases.

The relative differences in the capital cost accounts in Table 5 qualitatively follow the trend expected from the increased net power and reduced coal feed rate for the sCO_2 plants relative to the reference IGCC plant. The Coal Handling, Coal Prep, and Ash Handling accounts show a relative decrease in TPC corresponding to the decrease in coal feed rate. The increase in relative TPC for Accessory Electric Plant, Instrumentation & Control, and Improvements to Site accounts reflect the larger net power output of the sCO_2 plants compared to the reference IGCC plant. The TPC for the Gasifier account is 49% higher for the baseline sCO_2 plant than the IGCC plant; this is due to a combination of effects including the larger ASU needed to produce oxygen for the oxy-combustor and the 25% larger syngas cooler duty to preheat the sCO_2 feed to the oxy-combustor. The impact of the higher ASU TPC in the sCO_2 plant is more than offset by a reduction in the gas cleanup cost (due to not having to capture the CO_2 in the Sulfinol plant) and the CO_2 compression cost (since the CO_2 is delivered to the CPU at pressure). Further, the elimination of the high temperature sCO_2/syngas cooler in the baseline case appears as a significant TPC savings in the Case 2 Gasifier account.

Table 5 Comparison of Total Plant Costs (TPC)

Account	**Reference IGCC Plant Case B1B**		**Baseline sCO_2 Plant**		**sCO_2 Case 2**	
	TPC ($1000)	**TPC ($/kW)**	**TPC ($1000)**	**TPC ($/kW)**	**TPC ($1000)**	**TPC ($/kW)**
Coal Handling System	43,156	87	41,775	74	41,775	69
Coal Prep and Feed	218,724	440	199,571	355	199,571	329
Feedwater & Miscellaneous BOP	65,849	133	21,252	38	21,363	35
Gasifier and Accessories	681,168	1,371	1,014,116	1,803	889,417	1,468
Gas Cleanup & Piping	323,580	651	160,519	285	160,528	265
CO_2 Compression & Storage	81,688	164	60,601	108	61,460	101
Power Cycle & Accessories	160,049	322	261,793	465	290,387	479
HRSG Ductwork & Stack	56,527	114	0	0	0	0
Steam Plant	85,322	172	34,428	61	29,214	48
Cooling Water System	39,217	79	39,523	70	39,332	65
Ash/Spent Sorbent Handling Sys.	42,945	86	41,753	74	41,753	69
Accessory Electric Plant	103,353	208	113,231	201	115,201	190
Instrumentation and Control	32,252	65	33,972	60	34,829	57
Improvements to Site	22,691	46	23,966	43	24,655	41
Buildings and Structures	21,081	42	13,713	24	13,789	23
Total	**1,977,603**	**3,981**	**2,060,211**	**3,663**	**1,963,273**	**3,240**

The baseline sCO_2 Power Cycle & Accessories TPC is 64% higher than for the reference IGCC plant, though in the latter this only includes the combustion turbine generator. The aggregate power island TPC for the IGCC plant (including Heat Recovery Steam Generator (HRSG) and Steam Plant accounts), is 2% higher than the baseline sCO_2 plant, although it produces 13% less net power. Accounting for this aggregate power island cost, on a $/kW basis the sCO_2 capital cost accounts are all lower than the corresponding accounts for the IGCC plant except for the gasifier (28% higher). Overall, the respective $/kW TPCs for the baseline and Case 2 sCO_2 plants are 8% and 19% lower than the reference IGCC plant.

Compared to the baseline case, the Case 2 plant produces 8% more net power using the same amount of coal. The Case 2 plant has a slightly higher ASU capital cost ($349MM) compared to the baseline case ($347MM) commensurate with the 0.8% increase in oxygen production. The CO_2 Compression and Storage cost for Case 2 is 1.4% higher than for the baseline case due to the higher carbon capture fraction recovery in Case 2. Overall, the TPC for Case 2 is 5% lower than for the baseline case on an absolute basis and 12% lower on a $/kW basis.

Table 6 compares the sub-account level TPC between Case 2 and the baseline case for the Power Cycle and Accessories account. Overall, the sCO_2 cycle TPC for Case 2 is 11% higher than the Baseline Case but on a $/kW basis, there is only a 3% increase. The largest increases in the sCO_2 cycle TPC were from the HTR (~200% increase) and the LTR (64% increase). This was expected in removing the expensive high temperature sCO_2/syngas cooler from the baseline sCO_2 system, with a net savings in overall cost. The largest decrease in sCO_2 cycle TPC was due to Piping (27% decrease) due to elimination of high temperature piping runs between the sCO_2 cycle and syngas cooler.

Table 6 Power Cycle and Accessories sub-account component cost comparison for the sCO_2 plants

Sub-account	**sCO_2 Baseline Case**		**sCO_2 Case 2**	
	TPC ($1000)	**TPC ($/kW)**	**TPC ($1000)**	**TPC ($/kW)**
Oxy-turbine generator	51,515	92	53,987	89
High temperature recuperator	11,083	20	33,433	55
Low temperature recuperator	8,837	16	14,505	24
CO_2 pre-cooler	1,696	3	1,812	3
CO_2 pre-compressor	52,124	93	55,557	92
CO_2 cooler/condenser	7,058	13	7,566	12
CO_2 pump	23,636	42	25,006	41
O_2 compressor	43,681	78	41,140	68
Syngas compressor	34,848	62	35,211	58
Piping	21,585	38	15,781	26
Foundations	5,730	10	6,388	11
Total	**261,793**	**465**	**290,387**	**479**

4.2.2 Operating and Annualized Costs

Table 7 compares the annual O&M costs between the modeled plants. Most costs are comparable, with the exception of the water & chemicals account, where IGCC plants incur higher chemical costs due to the 2-stage Selexol process for CO_2 removal. The largest overall variations occur in fuel costs, which are directly related to plant thermal efficiency differences.

Table 7 Comparison of annual plant operating and maintenance (O&M) costs

	Reference IGCC Plant		**sCO_2 Baseline Case**		**sCO_2 Case 2**	
Account	**$1000/yr**	**$/MW-hr**	**$1000/yr**	**$/MW-hr**	**$1000/yr**	**$/MW-hr**
Fixed O&M						
Labor	31,837	9.1	35,673	9.1	34,243	8.1
Taxes & Insurance	39,552	11.4	41,204	10.5	39,265	9.2
Total Fixed O&M	**71,389**	**20.5**	**76,877**	**19.5**	**73,508**	**17.3**
Variable O&M						
Maintenance Materials	34,438	9.9	38,215	9.7	36,338	8.6
Water and Chemicals	6,543	1.9	3,400	0.9	3,370	0.8
Waste Disposal	4,166	1.2	3,864	1.0	3,864	0.9
Total Variable O&M	**45,146**	**13.0**	**45,479**	**11.5**	**43,573**	**10.3**
Fuel	**111,740**	**32.1**	**104,867**	**26.6**	**104,867**	**24.7**
Total O&M	***228,275***	***65.6***	***227,223***	***57.6***	***221,948***	***52.3***

Table 8 Comparison of Cost of Electricity (COE)

COE Component	**Reference IGCC Plant ($/MWh)**	**Baseline sCO_2 Plant ($/MWh)**	**Case 2 sCO_2 Plant ($/MWh)**
Capital	87.0	79.6	70.5
Fixed O&M	20.5	19.5	17.3
Variable O&M	13.0	11.5	10.3
Fuel	32.1	26.6	24.7
Total (Excluding T&S)	**152.6**	**137.3**	**122.7**
CO_2 T&S	9.8	8.8	8.3
Total (Including T&S)	**162.4**	**146.1**	**131.1**

Table 8 compares the Cost of Electricity between the modeled plants, including the plant O&M costs from Table 7. The COE for the baseline sCO_2 plant is 10% lower than the COE for the reference IGCC plant, both with and without CO_2 T&S costs. This decrease in COE is primarily due to the higher efficiency of the sCO_2 plant, and its reduced capital cost on a $/kW basis. Case 2 has a 10-11% lower COE compared to the baseline case, and a 19-20% lower COE than the reference IGCC plant, due to both reductions in TPC and improvements in efficiency. Improving thermal integration between the gasifier and direct sCO_2 cycle is thus shown to provide a significant benefit in terms of both plant thermal efficiency and cost of power production over the life of the plant.

4.2.3 Recuperator Approach Temperature Sensitivity

In the Case 2 configuration, the minimum temperature approach at the cold end of the LTR and at the hot end of the HTR are both 10 °C. A sensitivity analysis was performed in which this minimum temperature approach is increased to 20 °C. Table 9 compares the overall performance of these two cases.

Table 9 Performance summary for sensitivity analysis to minimum temperature approach

Parameter	10 °C min T_{app}	20 °C min T_{app}
Coal flow rate (kg/hr)	198,059	198,059
Oxygen flow rate (kg/hr)	394,234	394,231
sCO_2 flow rate (kg/hr)	7,734,832	7,575,265
Carbon capture fraction (%)	99.4	99.4
Captured CO_2 purity (mol% CO_2)	99.80	99.80
Raw water withdrawal (m^3/s)	0.337	0.346
Carbon conversion (%)	99.5	99.5
Power Summary		
Coal thermal input (HHV)	1,493	1,493
sCO_2 turbine power output (MW)	828	811
Total auxiliary power load (MW)	222	222
Net power output (MW)	606	589
Net plant efficiency (HHV %)	40.6	39.5
sCO_2 power cycle efficiency (%)	61.9	60.4
Economic Analysis Summary		
COE (w/o T&S) (2011$/MWh)	122.7	125.0

The overall result is that increasing the minimum temperature approach from 10 °C to 20 °C decreases the process efficiency 1.1 percentage points and increases the COE by 2.3 $/MWh or 2%. The expected recuperator capital cost benefit in increasing this approach temperature is not shown to offset the reduction in plant efficiency, on a COE basis.

4.3 Comparison to Other Studies

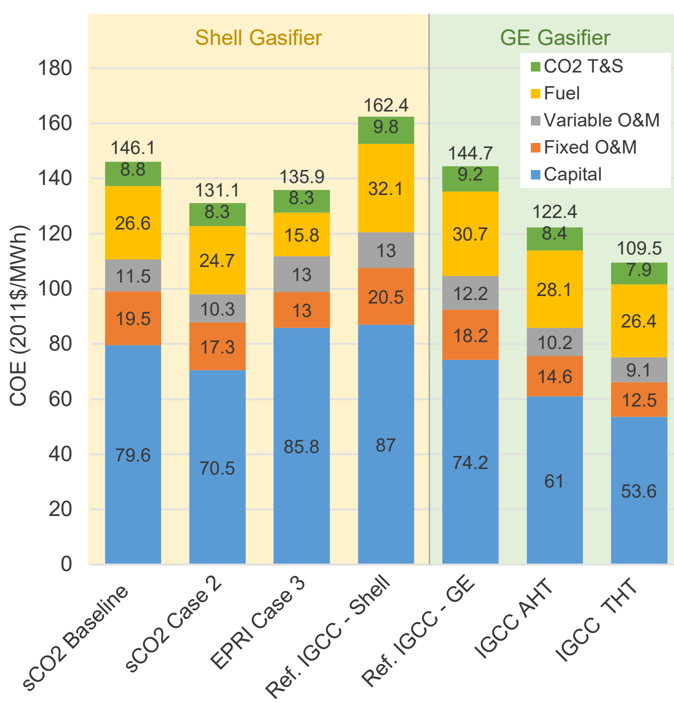
As noted above, the Case 2 sCO_2 plant design performs well compared to the baseline sCO_2 plant and the reference IGCC case. The performance results can also be compared against other gasification-based sCO_2 [3] [7] and IGCC [10] [29] plant design studies, as summarized in Table 10 and Figure 8. Economic results are compared in Figure 9 on a 2011 dollar-year basis.

Table 10 Plant Design and Performance Comparison

Item	sCO_2 Baseline	sCO_2 Case 2	EPRI Case 3 [7]	8 Rivers Case 1 [3]	Reference IGCC Shell Gasifier [10]	Reference IGCC GE Gasifier [10]	IGCC-AHT [29]	IGCC-THT [29]
Coal Type	Illinois #6	Illinois #6	PRB	Illinois #6	Illinois #6	Illinois #6	Illinois #6	Illinois #6
Coal Feed	Dry	Dry	Dry	Dry	Dry	Water Slurry	Water Slurry	Water Slurry
Gasifier Type	Shell	Shell	Shell	Shell	Shell	GE-RGC	GE-RGC	GE-RGC
Syngas Heat Recovery	Syngas cooler	Syngas cooler	Syngas cooler	Syngas cooler	Syngas cooler	Radiant Syngas cooler	Radiant Syngas cooler	Radiant Syngas cooler
Other Processes	Steam plant	Steam plant	Steam power cycle	None	Gas turbine steam cycle	Gas turbine steam cycle	AHT + steam cycle	THT + steam cycle
Sulfur Removal	AGR	AGR	AGR	DeSNOx	AGR	AGR	AGR	AGR
Turbine Cooling	Yes	Yes	No	?	Yes	Yes	Yes	Yes
Turbine Inlet Temp. (°C)	1204	1204	1123	1150	1337	1337	1450	1700
Net Plant Power (MWe)	562	606	583	~280	497	543	771	1057
Net Plant Efficiency (HHV, %)	37.7	40.6	39.6	49.6	31.2	32.6	35.7	38.0
Carbon Captured (%)	97.6	99.4	99.2	~100	90.0	90.0	90.0	90.0
Captured CO_2 Purity (%)	99.8	99.8	98.1	?	99.4	99.5	99.5	99.5

Figure 8 Net Plant Efficiency Comparison

Figure 9 Cost of Electricity Breakdown



The performance and economics of Case 3 from the EPRI study is very comparable to Case 2 of the present study, despite several important differences between the plant designs. The largest difference is in the fuel gasified, where EPRI uses a low-sulfur sub-bituminous Powder River Basin (PRB) coal, while the present study uses a more expensive bituminous coal (Illinois #6), as reflected in the relative fuel costs in Fig. 9. Overall, this study's inclusion of additional loss mechanisms not present in the EPRI study (e.g., turbine cooling streams, high combustor pressure drops, and a carbon purification system) are more than offset by process improvements in Case 2 (e.g., higher sCO_2 turbine inlet temperature, oxygen and syngas preheating, lower cold sCO_2 temperature, gasifier thermal integration), leading to a slight improvement in plant efficiency and COE relative to the EPRI study.

Among the cases studied by 8 Rivers Capital, Case 1 is the most similar to the current study, employing a high purity ASU to minimize nitrogen and argon diluent in the sCO_2 cycle, a Shell gasifier with dry feed of Illinois #6 coal, and a syngas cooler for heat recovery [3] [27]. The reported efficiency of this case is 49.7%, compared to 40.6% in Case 2 of this study [3]. The most significant difference is the use of in-cycle sulfur and NOx removal in the 8 Rivers plant, which allows for elimination of the AGR system in the present study, as well as all process steam generation, which is primarily used in the AGR and Claus sulfur recovery units.

The level of detail included in the Case 2 study allows for an evaluation of this approach as an improved syngas cleanup strategy, as well as a partial validation of 8 Rivers' reported plant efficiency. In the Case 2 plant design, 59.2 MW_{th} is used in the gasification train for steam generation, and unrecovered syngas heat from the minimum recoverable temperature of 115.6 °C to 35 °C amounts to an additional 23.6 MW_{th}. Accounting for the thermal duties of the Claus plant furnace (-7.1 MW_{th}), sulfur condenser (-3.9 MW_{th}), and the ASU's pre-purifier adsorbent regenerator (6.0 MW_{th}), a total of 65.8 MW_{th} might be available to the sCO_2 cycle in the absence of the AGR, Claus, and steam generation units, providing an additional 40.7 MW_e to the plant's electric power output at the sCO_2 cycle conversion efficiency of

61.9%. Elimination of auxiliary electric loads from these units could add another 2.1 MW_e to the plant's power output. The syngas and oxygen used in the Claus unit could add another 1.1 MW_e to the power output the sCO_2 plant, and inclusion of the chemical energy of the syngas sulfur content, as well as the additional oxygen needed to burn it, could add another 10.7 MW_e, for a total plant net power output of 660.5 MW_e. Assuming no additional heat or auxiliary power loads are required for in-cycle sulfur and NOx cleanup, this corresponds to a net thermal efficiency of 44.2% (HHV), 3.6 percentage points higher than that obtained in Case 2. Neglecting the additional low grade heat recovery, this validates the 3 percentage point efficiency benefit assigned to the elimination of the AGR in Reference [27]. Further, the capital cost impact of eliminating the steam, COS hydrolysis, AGR, and Claus units from the Case 2 results is estimated at $180 million, about 9.1% of the TPC, based on economic data in the Supplementary Information. These costs would be replaced by additional sCO_2/syngas heat exchangers and in-cycle cleanup component costs in a comparable plant, and may also incur additional costs for acid resistant materials of construction for components within the sCO_2 cycle.

If technically viable, this analysis shows that the in-cycle syngas cleanup approach pursued by 8 Rivers is a very promising approach to further improving the performance and economics of a coal-fueled direct sCO_2 power plant. To validate the reported plant efficiency of Case 1 in the 8 Rivers study, an adjustment must be made for the differences in the low sCO_2 temperatures used (20 °C used in the 8 Rivers study [27], versus 27 °C in Case 2), which improves thermal efficiency by about 1.2 percentage points due to increased sCO_2 density at the compressor and pump inlets, reducing their power requirements. This yields an adjusted Case 2 thermal efficiency of 45.4% at comparable conditions to 8 Rivers' Case 1 plant design, which is excellent for a coal-fueled power plant with carbon capture. The remaining thermal efficiency differences must be attributed to the unknown detailed process design, operating conditions and assumptions used by 8 Rivers Capital, which are not publicly available.

Finally, comparison to the Transformational Turbines Supplement to NETL's IGCC Pathway Study [30] [29] points to an additional improvement which should be considered in future work on the Case 2 plant design of the current study. This study considered the effects of advanced (AHT) and transformational (THT) hydrogen turbines with increased firing temperatures and pressure ratios on the performance and cost of an IGCC plant based on a GE gasifier with a radiant syngas cooler. Efficiency differences are shown in Figure 8, where the Shell gasifier cases on the left are contrasted with the GE gasifier cases on the right, including the reference IGCC case using a GE gasifier. Note that the GE gasifier results in a reference IGCC plant with CCS that is 1.4 percentage points more efficient, and has a COE that is $17.2/MWh lower (Fig. 9, w/o T&S) than the Shell gasifier plant [11]. Consideration of a GE gasifier with the direct sCO_2 cycle may therefore further improve the performance and cost-effectiveness of the Case 2 sCO_2 plant design, and may yield a COE that is lower than the AHT and THT IGCC cases.

5 Conclusions

The above study describes the integration of a coal gasification plant with a direct sCO_2 power cycle using a detailed and well-documented technoeconomic analysis methodology. This work demonstrates that an oxy-fired sCO_2 power plant can capture 99% of CO_2 and still have more favorable thermal efficiency and economics than most conventional or advanced power plant concepts that attain only 90% capture, without monetizing the CO_2 product.

While the baseline sCO_2 plant design is shown to yield significant cost and performance improvements compared to a Shell gasifier IGCC plant with CCS, an improved Case 2 sCO_2 plant design is pursued to further explore the performance potential of coal-fueled direct sCO_2 systems. Improvements to the baseline sCO_2 plant include: 1) elimination of the final stage of sCO_2 preheating from the syngas cooler, 2) premixing and preheating of the oxygen stream with sCO_2 prior to combustion, 3) replacement of the syngas-fired coal dryer with a hot nitrogen coal dryer, and 4) additional low quality heat recovery in the

sCO_2 power cycle in parallel with the LTR. Combined, these improvements result in a Case 2 plant thermal efficiency of 40.6%, and a COE of $122.7/MWh, with 99.4% carbon capture, among the highest reported efficiencies for coal-fueled plants with CCS. These performance results are derived from detailed system models that account for most major loss mechanisms, and include detailed plant component costs, albeit with a high degree of uncertainty in component costs for the direct sCO_2 cycle, and an overall estimated accuracy range of -15/+50%. Details on the baseline and Case 2 cycle conditions, thermal integration approaches, component costing, and operating costs are provided in Supplemental Information to provide transparency on the technoeconomic modeling methods used.

Further improvements on the Case 2 results are possible, as noted in Section 4.3. Improved efficiency or capital cost savings may result from investigation of other gasifier systems, alternative syngas cleanup strategies, and reductions in cold sCO_2 temperature via increased cooling capacity. Additional model refinements, such as detailed turbine cooling models, and consideration of the effects of incomplete combustion, could also be implemented to improve the fidelity of the study results, and may be pursued in future work. Beyond the steady-state analyses performed here, dynamic cycle modeling could also be performed in the future to assess the performance of this system at off-design and part load conditions, as would be required for a plant connected to a power grid that is subject to varying degrees of demand fluctuation due to renewable power integration. Additional technical detail on turbomachinery performance maps, recuperator geometries, and cycle control strategies, as needed for future dynamic modeling efforts, should become available as commercialization of direct sCO_2 cycles progresses.

The Case 2 results represent a significant improvement over most other coal gasification plant types with CCS, and provides an efficient, cost-effective route to utilization of coal for power production in an environmentally responsible manner. It is hoped that further developments in sCO_2 technology and near-term demonstrations of sCO_2 power cycle operation and benefits will help validate these results, and pave the way for commercial deployment of these systems.

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7 Disclaimer

8 References

[1] R. Allam, M. Palmer, G. J. Brown, J. Fetvedt, D. Freed, H. Nomoto, M. Itoh, N. Okita and C. Jones Jr., "High Efficiency and Low Cost of Electricity Generation from Fossil Fuels While Eliminating Atmospheric Emissions, Including Carbon Dioxide," *Energy Procedia,* vol. 37, pp. 1135-1149, 2013.

[2] R. Allam, J. Fetvedt, B. Forrest and D. Freed, "The Oxy-Fuel, Supercritical CO2 Allam Cycle: New Cycle Developments to Produce Even Lower-Cost Electricity from Fossil Fuels without Atmospheric Emissions," in *Proceedings of ASME Turbo Expo, ASME Paper: GT2014-26952*, Düsseldorf, Germany, 2014.

[3] X. Lu., B. Forrest, S. Martin, J. Fetvedt, M. McGroddy and D. Freed, "Integration and Optimization of Coal Gasification Systems with a Near Zero Emissions Supercritical Carbon Dioxide Power Cycle," in *Proceedings of ASME Turbo Expo, ASME Paper: GT2016-58066*, Seoul, South Korea, June 13-17, 2016.

[4] International Energy Agency Greenhouse Gas (IEAGHG), "Oxy-combustion Turbine Power Plants," Cheltenham, United Kingdom, August 2015.

[5] Electric Power Research Institute (EPRI), "Oxy-Fired Coal and Natural Gas Power Plants – 2016 Detailed Feasibility Study," 3002008148, Palo Alto, CA, 2017.

[6] A. McClung, K. Brun and J. Delimont, "Comparison of Supercritical Carbon Dioxide Cycles for Oxy-Combustion," in *Proceedings of ASME Turbo Expo, ASME Paper: GT2015-42523*, Montréal, Canada, 2015.

[7] Electric Power Research Institute (EPRI), "Performance and Economic Evaluation of Supercritical CO2 Power Cycle Coal Gasification Plant," (3002003734), Palo Alto, California, December, 2014.

[8] N. Weiland, W. Shelton, C. W. White and D. Gray, "Performance baseline for direct-fired sCO2 cycles," in *5th International Supercritical CO2 Power Cycles Symposium*, San Antonio, Texas, March 29-31, 2016.

[9] N. Weiland, W. Shelton and C. W. White, "Performance of an Integrated Gasification Direct-Fired Supercritical CO2 Power Cycle," in *IEA Clean Coal Centre's First Workshop on High Efficiency, Low Emissions Coal-Fired Plant (HELE2016)*, Tokyo, Japan, May 23-24, 2016.

[10] National Energy Technology Laboratory (NETL), "Performance and Cost Assessment of a Coal Gasification Power Plant Integrated with a Direct-Fired sCO2 Brayton Cycle," DOE/NETL-341/063016, Pittsburgh, PA, 2017.

[11] National Energy Technology Laboratory (NETL), "Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b – Year Dollar Update," Pittsburgh, Pennsylvania, 2015.

[12] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies -- Capital Cost Scaling Methodology," NETL, Pittsburgh, January 2013.

[13] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies -- Detailed Coal Specifications," NETL, Pittsburgh, January 2012.

[14] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies, Cost Estimation Methodology for NETL Assessments of Power Plant Performance," NETL, Pittsburgh, January 2013.

[15] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies: Estimating Carbon Dioxide Transport and Storage Costs," NETL, Pittsburgh, 2013.

[16] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies - CO2 Impurity Design Parameters," Pittsburgh, Pennsylvania, August, 2013.

[17] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies -- Process Modeling Design Parameters," Pittsburgh, Pennsylvania, May, 2014.

[18] National Energy Technology Laboratory (NETL), "Quality Guidelines for Energy System Studies: Carbon Dioxide Transport and Storage Costs in NETL Studies," Pittsburgh, September, 2013.

[19] C. W. White and N. T. Weiland, "Evaluation of Property Methods for Modeling Direct-Supercritical CO2 Power Cycles," in *Proceedings of ASME Turbo Expo 2017: Turbomachinery Technical Conference and Exposition: GT2017-64261*, Charlotte, North Carolina, June 26-30, 2017.

[20] AACE International Recommended Practice Number 16R-90, "Conducting Technical and Economic Evaluations -- As Applied for the Process and Utility Industries; TCM Framework 7.3 -- Cost Estimating and Budgeting," AACE, 2003.

[21] H. Maeda and others, "Turbine and Operating Method of the Same". USA Patent 2014/0023478 A1, 23 January 2014.

[22] M. T. Breslin, "Heavy Shell Casings Design/Operation and Repair Criteria," in *Proceedings of the 18th Turbomachinery Symposium*, 1989.

[23] Gas Turbine World, "GTW Handbook 2009," *Gas Turbine World,* pp. 72-91, 2009.

[24] Aerojet Rocketdyne, "Turbine Technology Program Phase III - Task 2(b): Supercritical CO2 Turbomachinery (SCOT) System Studies Recuperator Development," DOE Contract No. DE-FE0004002, August 13, 2014.

[25] R. Turton, R. C. Bailie, W. B. Whiting and J. A. Shaeiwitz, Analysis, Synthesis, and Design of Chemical Processes, 2nd Edition, Prentice Hall, 2003.

[26] National Energy Technology Laboratory (NETL), "Report on Newly Developed A-USC Materials," NETL, Pittsburgh, 2014.

[27] X. Lu, "Novel sCO2 Allam Cycle for High-Efficiency, Low Cost and Emission-free Power Generation," in *IEA Clean Coal Centre: High Efficiency, Low Emissions Coal-Fired Plant Workshop*, May 23-25, 2016.

[28] C. W. White, W. Shelton and R. Dennis, "An Assessment of Supercritical CO2 Power Cycles Integrated with Generic Heat Sources," in *4th International Supercritical CO2 Power Cycles Symposium*, Pittsburgh, PA, September 2014.

[29] W. W. Shelton, R. Ames, R. Dennis, C. W. White, J. Plunkett, J. Fisher and R. Dalton, "Advanced and Transformational Hydrogen Turbines for Integrated Gasification Combined Cycles," in *ASME Turbo Expo, ASME Paper: GT2016-57723*, Seoul, South Korea, June 13-17, 2016.

[30] National Energy Technology Laboratory (NETL), "A Pathway Study on IGCC R&D with Carbon Capture Using Bituminous Coal Supplemental Chapter: Transformational Turbines," DOE/NETL-2014/1639, Pittsburgh, PA, June 23, 2015.

9 Supplementary Information

Supplementary Information on this work can be found at: https://doi.org/10.1016/j.fuel.2017.10.022, including stream table and pinch analysis data for the sCO_2 Baseline Case and the improved Case 2, as well as detailed capital cost accounts and O&M accounts for each of the sCO_2 cases.