Comparison of CO₂ Capture Approaches for Fossil-Based Power Generation: Review and Meta-Study

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Abstract: This work is a meta-study of CO_2 capture processes for coal and natural gas power generation, including technologies such as post-combustion solvent-based carbon capture, the integrated gasification combined cycle process, oxyfuel combustion, membrane-based carbon capture processes, and solid oxide fuel cells. A literature survey of recent techno-economic studies was conducted, compiling relevant data on costs, efficiencies, and other performance metrics. The data were then converted in a consistent fashion to a common standard (such as a consistent net power output, country of construction, currency, base year of operation, and captured CO_2 pressure) such that a meaningful and direct comparison of technologies can be made. The processes were compared against a standard status quo power plant without carbon capture to compute metrics such as cost of CO_2 emissions avoided to identify the most promising designs and technologies to use for CO_2 emissions abatement.

1. Introduction

In the US, greenhouse gas (GHG) emissions from electric power generation constitutes about 29% of all GHG emissions over the course of a year [1], mostly arising from the combustion of fossil fuels such as coal and natural gas. However, there are many approaches available for reducing these emissions in meaningful quantities, such as replacing fossil-based power plants with renewable-based plants such as wind, solar, hydroelectric, waste, or biomass-powered electricity generation systems. Other possibilities include focusing on efficiency increases of traditional plants as a way of reducing fuel consumption (and thus GHG emissions across the entire supply chain) for the same power output, or using nuclear-based power generation, which has a very low carbon footprint. Another approach is to continue to use fossil-based power plants, but instead focus on using carbon dioxide capture and sequestration (CCS) as a way of avoiding most of the CO_2 emissions while still using fossil fuels. Each of these options has its own strengths and weaknesses in terms of cost, GHG reduction potential, feasibility, geological limitations, socio-political and cultural factors, logistics, resource availability, technical maturity, environmental impacts, and risks. As such, all of these options are likely to continue to play a role in the global electricity generation mix (for a perspective on why and an overview of these technologies for the North American market, see [2]).

This work focuses specifically on CO_2 capture technologies and processes. Solvent-based CO_2 capture has been in commercial use for decades as a way of either producing CO_2 for sale as a chemical or a necessary cleaning step in which CO_2 is a contaminant that must be removed, such as in removing the CO_2 from raw natural gas, from combustion flue gas, or from synthesis gas in a Fisher-Tropsch

process. The basic principle is to contact a gas stream in an absorber with a solvent that is selective for CO_2 over the other chemicals in the gas. The CO_2-rich solvent can then be regenerated using distillation (usually a stripper-type configuration) to release the CO_2 gas from the solvent such that the lean solvent can be reused again. Although the technology itself is reasonably well understood, its application in the power generation sector for the explicit purpose of avoiding GHG emissions for environmental purposes is only in the beginning commercial stages. This is because the amount of carbon dioxide that must be captured from a flue gas is so massive that the sheer size of the project makes it very expensive.

For example, the largest such commercial installment, the SaskPower Boundary Dam Power Station in Saskatchewan, Canada, was completed in 2014. It uses an amine solvent-based post-combustion capture process. From May 2016 to April 2017, the facility averaged about 109 MW of net power output with about 83% uptime and about 58% capture of generated CO_2 [3]. This is a significant achievement, especially considering that the captured CO_2 is actively being transported via a pipeline and sequestered underground. Some of the captured CO_2 is used for enhanced oil recovery, a process in which supercritical CO_2 is injected into a spent oil field that enables additional recovery of oil (for a review, see [4]). This is one of the main drivers of the project, because at this location, captured CO_2 has value as a product, as opposed to a negative value as a waste. Furthermore, the facility continues to improve its performance and gain valuable experiential knowledge. However, the technology is still not quite mature, since in order to be applied to power plants at the global scale, it must be implemented with a net power output in the 500 to 1000 MW range with CO_2 capture rates of up to 90%. It is an open question as to whether 90% is the optimal amount, since there are diminishing returns on increasing this number (the cost and energy penalties per CO_2 captured increases as the capture rate increases). However, it is the standard used by most studies and benchmarks, such as by the International Panel on Climate Change reports [5] or the US Department of Energy's "Baseline" studies [6].

1.1. Post-Combustion Solvent-Based Technologies

A schematic of a supercritical pulverized coal (SCPC) power plant using post-combustion CO_2 capture is shown in Figure 1. First, coal is pulverized and combusted with air in a furnace (or boiler). The heat of combustion is used to make steam at various pressure levels, with the highest pressure at about 240 bar (above the critical point of water). The steam produces mechanical power in steam turbines, which are attached by a shaft. The shaft is attached to a generator, which converts the mechanical power to electric power. A condenser is used to produce liquid water from the turbine exhaust, and then a pump is used to recompress the water to supercritical pressures. The combustion exhaust leaving the furnace typically goes through an ash removal, a NO_x removal, and a sulfur removal process, in which there are various options for each stage. When CO_2 capture is used, the gas leaving the final cleaning stage is sent to a solvent-based CO_2 capture process, which removes CO_2 for later compression, transport, and storage. The remainder of the gases are exhausted to the atmosphere. Typical solvents for this purpose include monoethanolamine (MEA), diglycolamine (DGA) [7], and methyldiethanolamine (MDEA) possibly mixed with piperazine [8,9].

A typical natural gas combined cycle (NGCC) process with post-combustion CO_2 capture is shown in Figure 2. In this process, natural gas is combusted with compressed air at high pressure in a gas combustion turbine, producing mechanical power. Again, a generator is typically attached to convert the mechanical power to electric power. The combustion exhaust leaves at high temperature, and so a heat exchanger network is used to capture this heat by making high-pressure steam for steam turbines, which in turn produce additional electric power. The cooled combustion gases are then subjected to a solvent-based absorption system for CO_2 removal. The solvent-based system typically uses an absorber column to scrub the CO_2 from the gases, with the cleaned gases exhausted to the flue. The loaded solvent is then purified in a stripper, which recovers lean solvent in the bottom and the CO_2 in the distillate for compression and transport.
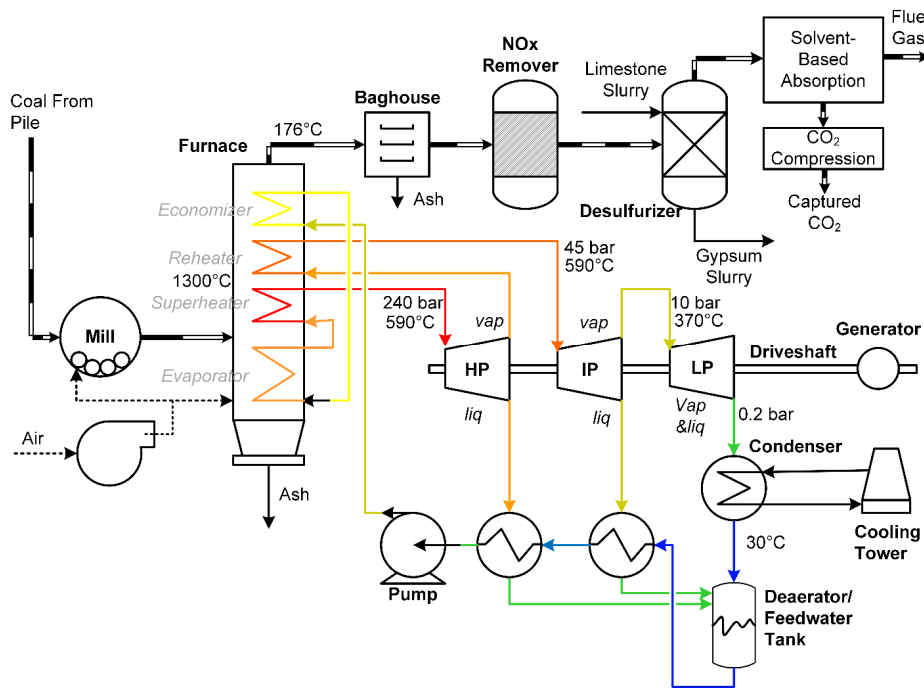
Figure 1. A typical schematic of a supercritical pulverized coal (SCPC) process with post-combustion carbon capture. Adapted from a combination of [6,10].
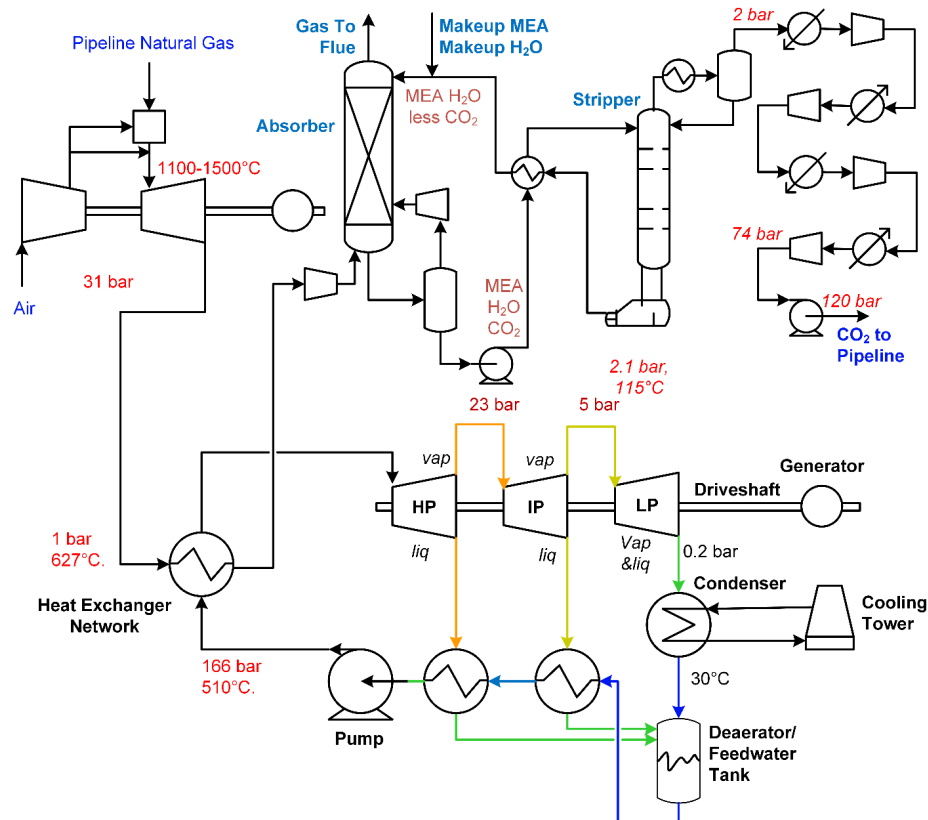
Figure 2. A typical natural gas combined cycle (NGCC) process with post-combustion carbon capture for an example system using monoethanolamine (MEA) as a solvent. Adapted from a combination of [6,11].

An alternative to amine-based solvents is ammonia. Although the process is similar in structure to amine-based processes, in the ammonia-based process, aqueous ammonia reacts with CO_2 in the flue gas in an absorption tower at atmospheric pressure and low temperatures (either at room temperature [12] or at chilled temperatures of 0–20 °C [13,14]) to form ammonium bicarbonate salts. The flue gas is exhausted after a washing step, and the ammonia bicarbonate slurry leaving the bottom of the absorber is sent to a regenerator at higher temperatures (100 °C) and pressures (20–40 bar) [14] in which the CO_2 is desorbed and released as a high pressure gas. The CO_2 is washed with water to separate the residual ammonia contained in the gases, and then the CO_2 is cooled and compressed to supercritical pressures for pipeline transport. The resulting lean ammonia solution leaving the bottom of the regenerator is recycled to the absorber, and the aqueous ammonia leaving the bottom of the water wash is recovered via a stripper [15].

Alternatively, cryogenic distillation is potentially competitive with solvent-based approaches. In this process, the post-combustion flue gas is first cooled, such that the water is condensed out. Then, the dried flue gas is chilled to between −100 °C [16] and −140 °C [17] depending on the design and the amount of CO_2 captured [18]. Under these conditions, the CO_2 precipitates out as a solid (for example, on the surface of heat exchangers [19]) thus separating the CO_2 from the N_2. The CO_2 can then be reheated/desublimated to a gas for treatment, compression, and storage. The key weakness is the energy intensity of generating the cooling loads necessary to condense the water and freeze the CO_2. Nevertheless, the temperatures are not as cold as the cryogenic air separation [18] necessary for other competing processes (see Sections 1.2 and 1.3), there are opportunities for heat integration with other processes, and no solvents are necessary. However, because of the lack of techno-economic analysis data available in the literature, cryogenic distillation is not included in the meta-study.

The post-combustion capture approach is attractive in part because it can be retrofitted to existing chemical plants by being "added" at the back end. The ability to use the existing power plant, supporting infrastructure, regulations, contracts, and technical know-how is very beneficial. As such, the cost of a retrofit to an existing plant may be low enough to make sense. However, post-combustion CO_2 capture is very expensive and energy intensive. For example, one study found that adding post-combustion solvent-based CO_2 capture to a classic pulverized coal power plant caused the average levelized cost of electricity (LCOE) to increase by 50–90% [20]. This means that, although the capture system may recover about 90% of the CO_2 actually generated by the facility, the average lifecycle GHG emissions reduction of the plants surveyed in this work is only in the range of 65–71%, which compares well to reported ranges of 50–75% in other studies [21–23].

Moreover, because the fuel use is so much higher, adding CO_2 capture to a combustion-based power plant makes the environmental impacts as much as 10–67% worse in many other kinds of environmental impact categories [21,22]. For example, one life cycle analysis study concluded that adding post-combustion capture to a coal combustion power plant causes 38% more fossil fuel consumption, 46% more toxicity impacts in humans, 58–59% more toxicity in freshwater, marine, and land-based ecosystems, and 67% more freshwater depletion [22]. The decision as to whether the extra environmental damage in these areas is worth the reduction in climate change potential is at least partially a matter of emotional and political priorities. Life cycle impact assessment methodologies such as the ReCiPe method can be used to help make this comparison quantitatively, but recent studies looking at this issue found opposing results. One group of studies found that post-combustion capture was still better for the environment on the whole despite all of the negative trade-offs for both coal [22] and natural gas [23], while another found that it was better for the environment not to use post-combustion capture at all for either fuel, instead preferring business-as-usual [24]. Nevertheless, Saskatchewan has clearly decided that the trade-offs are worth it, at least for now.

Even so, the significant cost and efficiency penalties of post-combustion capture make it very unattractive for any non-retrofit scenario. For example, if an existing power plant is approaching the end of its lifetime, retrofitting a post-combustion CO_2 capture section may not make sense. As the results of this meta-study will show, for new builds, post-combustion capture makes even less sense as

a long-term solution. As such, there are a variety of technologies that are currently under development or study that use a completely different approach to generating electricity from fossil fuels specifically to avoid having to use a solvent to recover CO_2 from combustion flue gases, and some are not even at the pre-commercial stages yet. These processes include the integrated gasification combined cycle (IGCC), oxyfuel combustion, membrane-based CO_2 capture, and solid oxide fuel cell (SOFC) power generation.

1.2. Pre-Combustion Solvent-Based Technologies (Integrated Gasification Combined Cycle)

The next-closest in technical maturity to post-combustion capture is the IGCC process, shown in Figure 3. In this process, a cryogenic air separation plant is used to separate O_2 from air. The O_2 is fed to a gasifier in which coal is gasified in an oxygen-deficient environment at high pressure (about 50 bar). The oxygen partially combusts some of the coal to produce a mixture of CO, CO_2, H_2, COS, and H_2O (known as syngas) at high temperatures (as high as 1350 °C). This syngas is then shifted using the water gas shift (WGS) reaction, in which the CO and COS are reacted with water to produce H_2 and H_2S, respectively. Then after a number of cleanup steps, including the removal of water, mercury, and sulfur, the syngas primarily contains just H_2 and CO_2.
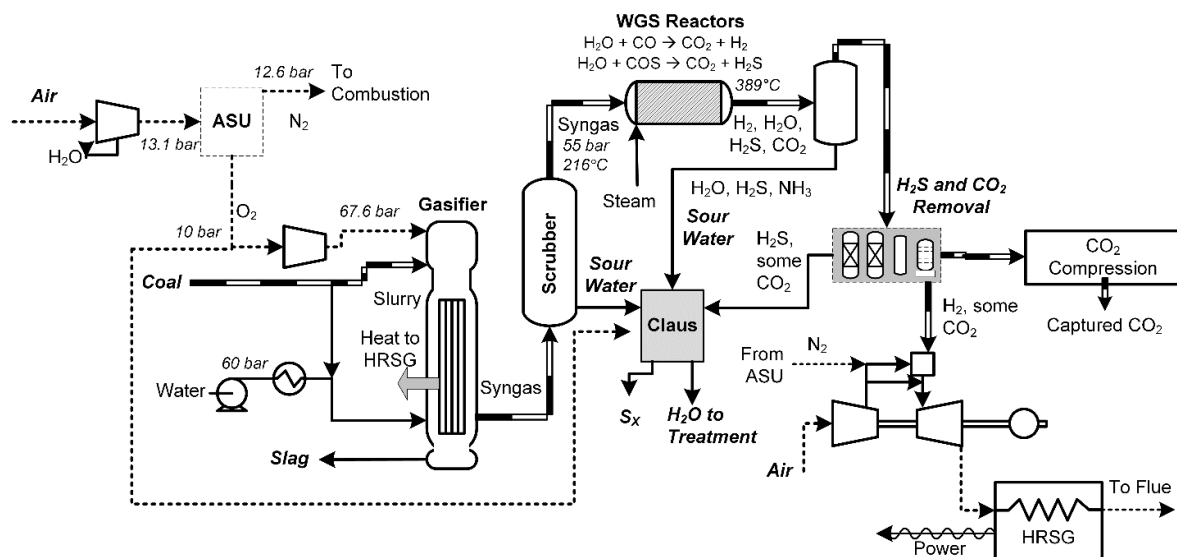
Figure 3. An example integrated gasification combined cycle (IGCC) process, based on a combination of [6,11,25,26].

A solvent-based process is used to remove most of the CO_2 from this stream for compression and storage purposes, as shown in Figure 4. Both H_2S and CO_2 need to be removed from the syngas, and the same solvent can be used to remove both. Typically, this is done in two stages, such that the H_2S and CO_2 can be removed separately using two separate absorbers in series. For the example using Selexol shown in Figure 3, the shifted syngas enters the first absorber, where it contacts a solvent mixture that absorbs the H_2S in the syngas (note that this solvent mixture is already loaded with CO_2 from the second absorber). The solvent leaving the first absorber (now loaded with both CO_2 and H_2S) enters a gas stripper, in which some nitrogen is added in order to remove some of the CO_2, such that the H_2S concentration in the loaded solvent increases. The loaded solvent is then purified in a stripper in which the recovered H_2S with some accompanying CO_2 is sent to the Claus process for conversion into elemental sulfur. The lean solvent is then used in the second absorber to recover CO_2. The larger portion of the loaded solvent leaving the CO_2 absorber is regenerated using a series of cascading flash drums (a stripper could be used, but this would be more expensive). Note that the pressure of the recovered CO_2 gas from the flash drums is moderately high, such that the costs of CO_2 compression are

much lower than for SCPC and NGCC post-combustion approaches that capture the CO_2 at flue gas pressures (near atmospheric). In addition to Selexol, other suitable solvents for various pre-combustion capture include MDEA [8], MDEA/Piperazine blends [9], MEA [27] (also called EconamineFG [6,28]), diglycolamine [29], and Rectisol [30].

After removal of the CO_2, a stream of primarily H_2 gas at high pressure remains, which is then combusted in a gas turbine. However, to avoid very high combustion temperatures, nitrogen from the air separation unit is used as a diluent. Like the NGCC case, the exhaust gases from the combustion turbine are sufficiently hot to power a second combined cycle using steam as a working fluid.
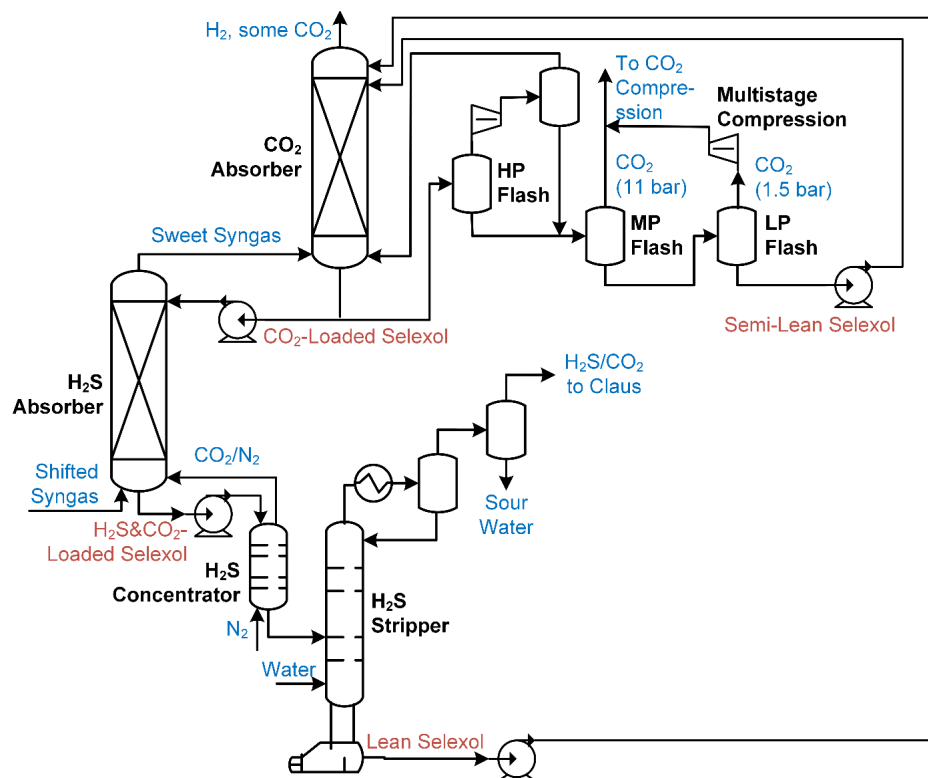

Figure 4. An example CO_2 removal process for IGCC, based on a combination of [6,11,26].

One such plant is currently running in Edwardsport, Indiana, USA, though not with the CO_2 capture and sequestration component [31]. Another IGCC plant in Kemper County, Mississippi, USA, is still under construction, with \$7.3 billion spent as of March 2017 [32], but this includes the CO_2 pipeline for transport of captured CO_2 (which has already been constructed), and the coal mine as well. IGCC uses a combination of coal gasification and the water gas shift reaction to produce a high-energy stream containing essentially H_2 and CO_2, such that all of the carbon in the coal ends up in CO_2, while most of the energy in the coal ends up in the H_2. The CO_2 can be separated from the H_2 using a solvent-based absorption process, with the remaining H_2 combusted in a gas turbine for power. In theory, this pre-combustion solvent-based capture approach carries a lower energy penalty than post-combustion capture, such that the overall GHG emissions and environmental impact is lower than for solvent-based post-combustion capture, although only by about 5–13% [5,22]. Curiously, there is conflicting evidence as to whether IGCC without carbon capture is any better for the environment than current state-of-the-art coal combustion power plants without capture [5,22]. This is also reflected in the survey results of this meta-study, as discussed in Section 3. Noting that the Edwardsport plant was far more expensive than expected [33], this indicates that the primary reason to choose IGCC as far as climate change is concerned is primarily for the slight potential environmental advantage over post-combustion capture when CO_2 capture is implemented.

One particular challenge with IGCC is the relatively poor availability of the gasifier, since gasifiers are notoriously difficult to operate. For example, Spliethoff [34] presented the availability of four gasifiers over time (Wabash, Tampa, Puertollano, and Buggenum) and showed that the most successful gasifiers operated at 70–80% uptime, compared to an average uptime for traditional coal and gas combustion of about 88%. The Puertollano gasifier took 5 years after commissioning to reach a peak availability of about 62%, and the Buggenum gasifier took 10 years of gradual improvements to reach 70%. Because of this, it is not uncommon to require the construction of two identical gasifiers, one of which is kept in a hot-standby state to be used in the event of the failure of the other gasifier, such as at the Eastman Chemical Company facility in Kingsport, Tennessee. This allows for an overall 98% system uptime [35], but adds considerably to the cost. If gasifier uptime can be improved significantly, this may help reduce costs and give IGCC a wider acceptance.

An alternative configuration uses an air-blown gasifier instead of high purity oxygen, and uses a pre-combustion capture approach. This simplifies the upstream portions of the process by eliminating the need for high-purity oxygen (thus eliminating the energy-intensive air separation unit). However, because of the large amounts of nitrogen in the syngas, the CO_2 capture problem is more difficult, both because of the significantly decreased partial pressure of CO_2, and because the CO_2 capture problem is a more difficult CO_2/N_2 separation problem, as opposed to the less difficult CO_2/H_2 separation problem of oxygen-blown IGCC. This requires an amine (such as MDEA [36,37]). It is also possible to use a post-combustion approach with air-blown IGCC as well, using an amine such as MEA [38] or with chilled-ammonia [39]. However, in both cases, there is little evidence that this would be a substantial improvement (or an improvement at all) over oxygen-blown gasification approaches from a cost perspective. We could not include the air-blown approaches in the meta-study because of the lack of techno-economic data in the literature.

1.3. Post-Combustion Solvent-Free Water Condensation-Based Technologies (Oxyfuels, Solid Oxide Fuel Cells, Chemical Looping Combustion, and Calcium-Oxide Looping).

The key principle of the oxyfuel process is to combust fuels in a N_2-lean environment. As shown in Figure 5, cryogenic air separation is used to separate O_2 from air, and use the recovered O_2 as the oxidant for combustion instead of air. Without N_2 present, once the ash, NO_x, and sulfur are removed, the combustion flue gases contain primarily a mixture of CO_2 and H_2O. These are relatively easy to separate via condensation of the water. However, without N_2, some other gas must be used as a diluent, and usually captured CO_2 downstream is recycled to the combustor for this purpose. Although there are many technical challenges with this approach, pilot plants up to 30 MW have been constructed [40], making them less mature than IGCC, but still feasible for large-scale implementation in the relatively near term. As will be shown in this work, oxyfuels have lower GHG emissions than post-combustion solvent-based capture, and are more cost effective on average, although there is high variability from study to study.
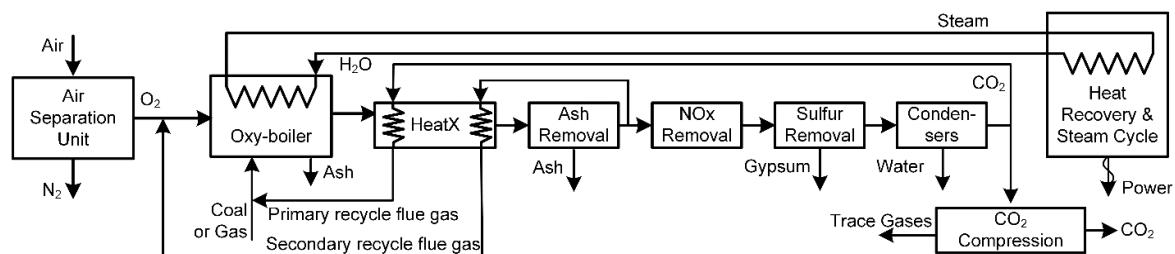
Figure 5. An example oxyfuel process, based on [41].

Solid oxide fuel cell (SOFC) power generation systems operate on much the same principle as oxyfuels, in that the fuel should be used without N_2 present, such that the exhaust products are an easily-separable CO_2 and water mixture. SOFC systems, however, do not use cryogenic air separation,

but instead feed ordinary air into the cathode of the SOFCs, as shown in Figure 6. Oxygen in the air, and not nitrogen, migrates through the solid barrier to the anode, where it reacts electrochemically with the fuel. As such, the SOFC acts as both electric power generator and air separator. It is a high-temperature (>800 °C) process with its own set of technical challenges, with low cell lifetime and difficulty of manufacture being just two of the major barriers to commercialization [42]. SOFCs are commercially available up to the 250 kW scale, but are largely used in niche markets, such as low-noise and low-vibration applications. As such, they are less mature than oxyfuels, but it is still within the realm of technical possibility to design large-scale processes which use existing materials and designs. However, much like IGCC and oxyfuels, various studies suggest that unless SOFC costs can be reduced significantly, SOFC power plants are not likely to be economically competitive with traditional power plants on a purely cost basis. Instead, their primary attractiveness is that SOFC systems combined with CCS tend to be better than post-combustion solvent-based alternatives in terms of cost, GHG reduction potential [22,42–44], or the balance between the two, as will be shown in Section 3.
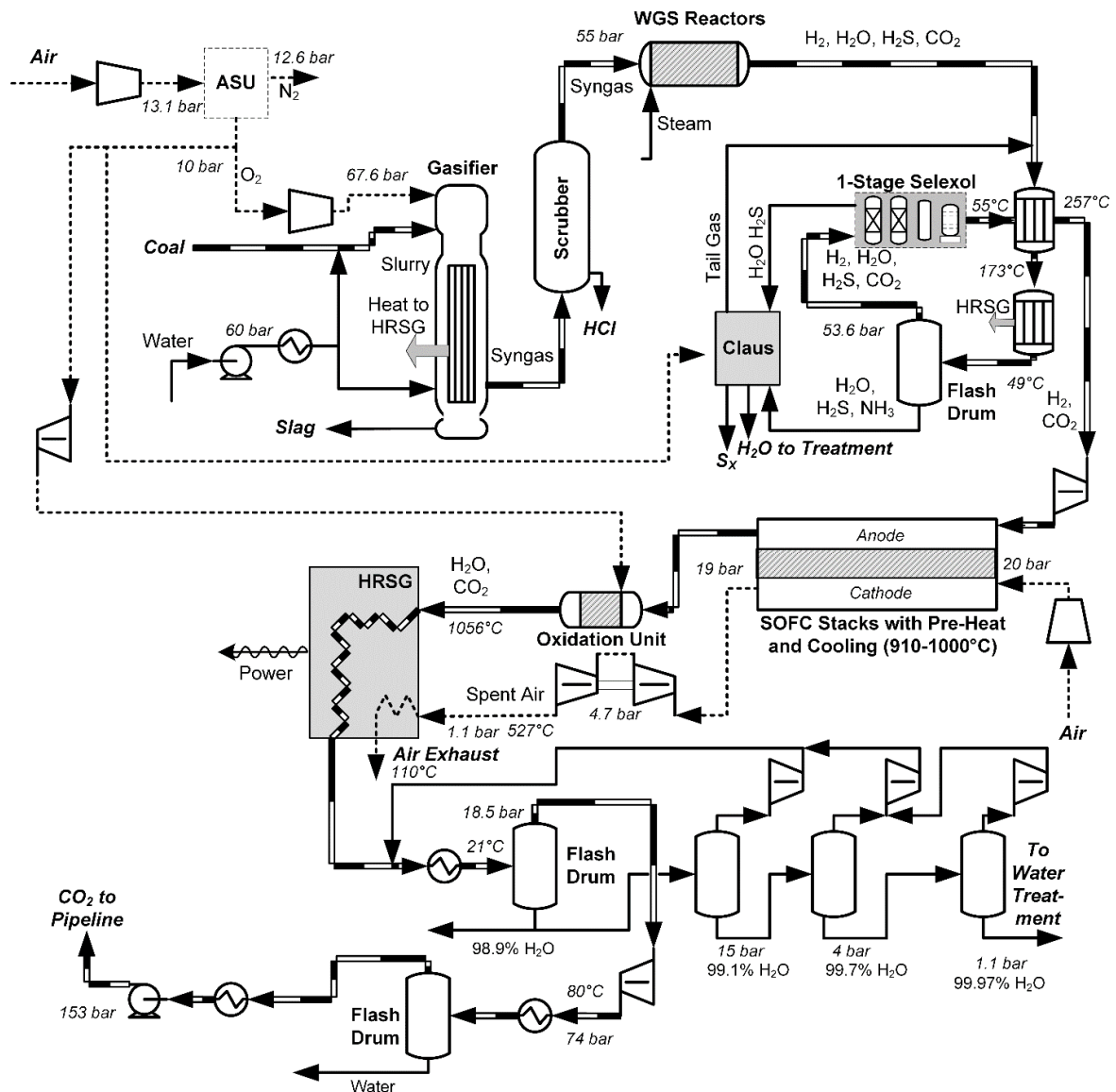

Figure 6. An example SOFC process from coal (also called integrated gasification fuel cell or IGFC), based on [25].

Like oxyfuels and SOFCs, chemical looping combustion (CLC) processes are designed to combust fuel without the presence of nitrogen by using a metal as an oxygen carrier. With this approach, oxygen is removed from air by reacting air at high pressure with reduced metal particles (such as nickel, iron, or calcium), thus producing a hot stream of metal-oxide particles entrained in hot air. The metal oxide particles are separated from the spent air via cyclone, and the cleaned, hot, high-pressure air is fed to a turbine to recover electricity. The recovered hot metal oxide particles are then reacted directly with a gaseous fuel (also at high pressure, such as syngas, natural gas, or hydrogen). The fuel is oxidized into CO_2 and H_2O, still at high pressure, by reducing the metal oxides. Heat is recovered from the CO_2/H_2O stream for steam generation for additional power, and the CO_2 can be captured by condensation of the water. The reduced metal is then returned to the air reactor, where it is oxidized again, thus completing the loop. A summary is shown in Figure 7. For extensive reviews on this topic, including how this process can be modified for the purposes of coal gasification or hydrogen generation, see [45,46].
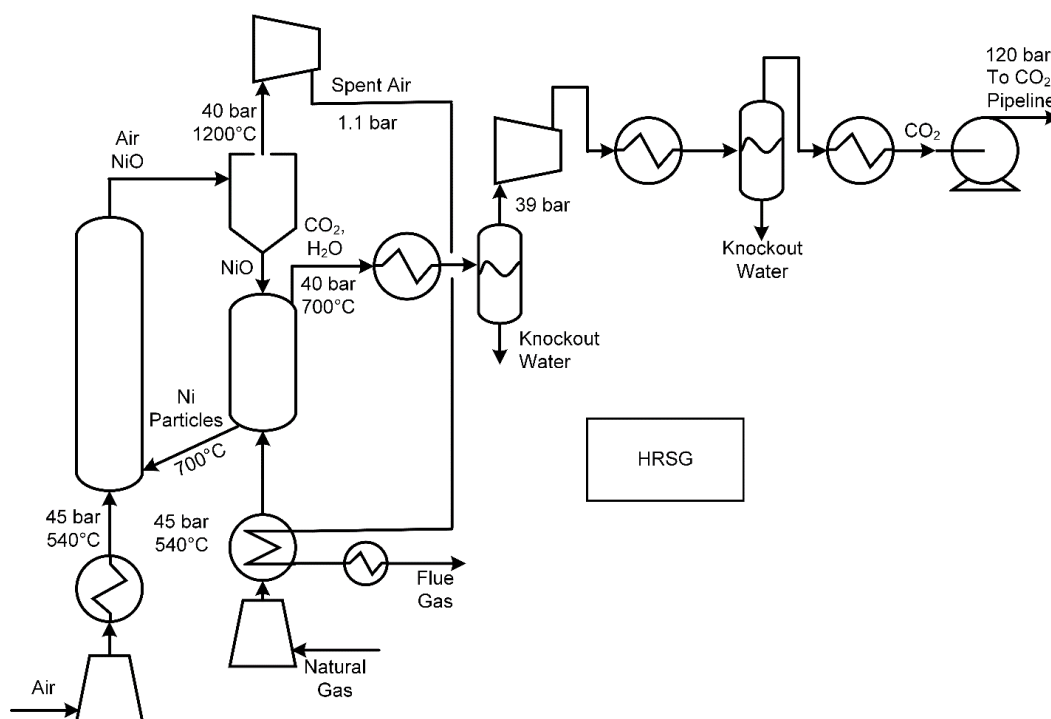
Figure 7. An example chemical looping combustion process using NiO, based on [46].

Calcium looping processes (CLP) take a different approach to CO_2 capture in which CO_2 is directly reacted with calcium oxide [47–49] or calcium hydroxide [50] to produce solid calcium carbonate, which can be separated easily from flue gas. The calcium oxide is regenerated in a calcination reactor by reaction with high-purity oxygen and recycled for more CO_2 capture. The calcination reactor is highly endothermic, and requires a large heat supply at high temperatures (above 800 °C), which is often met by in-situ oxy-combustion of coal or natural gas inside the calcination reactor. The CO_2 released from the calcination reactor is captured and compressed for storage after a heat recovery step. A spent sorbent purge stream is removed from the calcination reactor, in part because of the build-up of calcium sulfate and other solids that result from reactions with impurities in coal or other fuels. The key advantage is that there are many opportunities for heat integration within this process and with other processes. The key disadvantages are the large parasitic load of producing high purity oxygen, the relatively large amount of make-up sorbent (calcium carbonate) required, and the somewhat low CO_2 capture rate (around 85%). An example process is shown in Figure 8.

Figure 8. An example calcium looping process for post-combustion CO_2 capture.
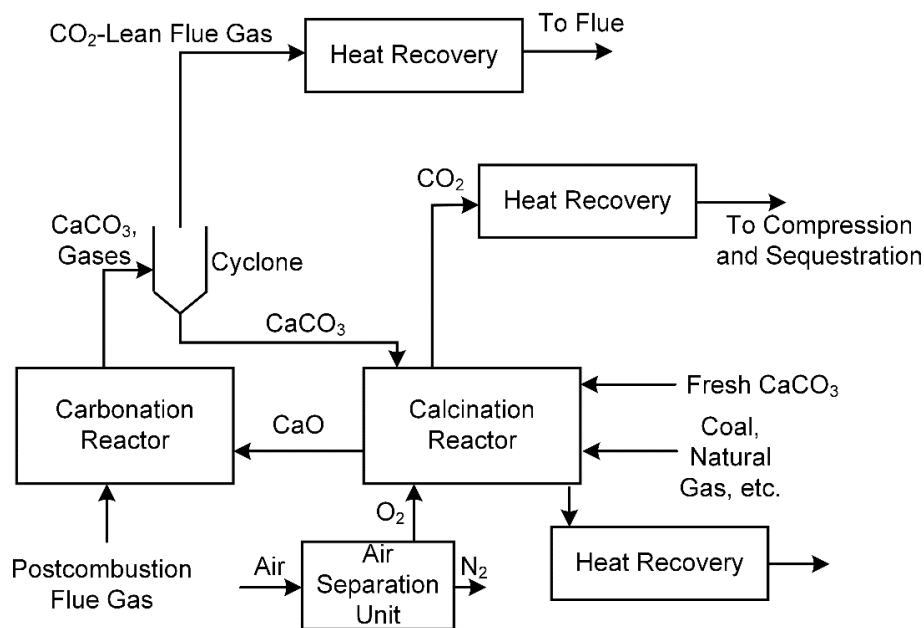
1.4. Pre-Combustion and Post-Combustion Membrane-Based Technologies

Membrane-based approaches use a variety of techniques that can be classified as either pre-combustion or post-combustion. Pre-combustion membranes use a gasification approach similar to IGCC, in which gasification and water gas shift reactors are used to produce a mixture of CO_2 and H_2, with a membrane (or multiple stages of them) used to separate CO_2 from H_2 instead of a solvent-based process. Either the CO_2 or the H_2 could be the permeating species, depending on the design of the system. An example superstructure of the processes surveyed in this meta-study is shown in Figure 9. In this case, the upstream portions of the process are very similar to the upstream portions of the IGCC process (Figure 3) or the coal-based SOFC process (Figure 6), in which desulfurized, shifted syngas is produced (primarily consisting of H_2 and CO_2). An H_2/CO_2 selective membrane is used to separate this into the component streams, with the captured CO_2 sent to a compression train for sequestration, and the remaining H_2 (with some CO_2 remaining) used for combustion in a gas turbine. It may also be possible to enhance the H_2/CO_2 membrane separation performance by using a membrane-enhanced water gas shift reactor, in which some of the H_2 is removed in the water gas shift reactor itself, as the reaction occurs. This has the dual benefit of shifting the reaction equilibrium toward a higher CO conversion (either resulting in more H_2 and CO_2 production or permitting operation at higher temperatures) and increasing the CO_2 concentration of the reactor effluent by removing the H_2 product. Nitrogen can be used as a sweep gas on the permeate side of the membrane-enhanced water gas shift reactor since the hydrogen gas needs to be diluted with nitrogen anyway before combustion in the gas turbine to prevent extremely high temperatures.
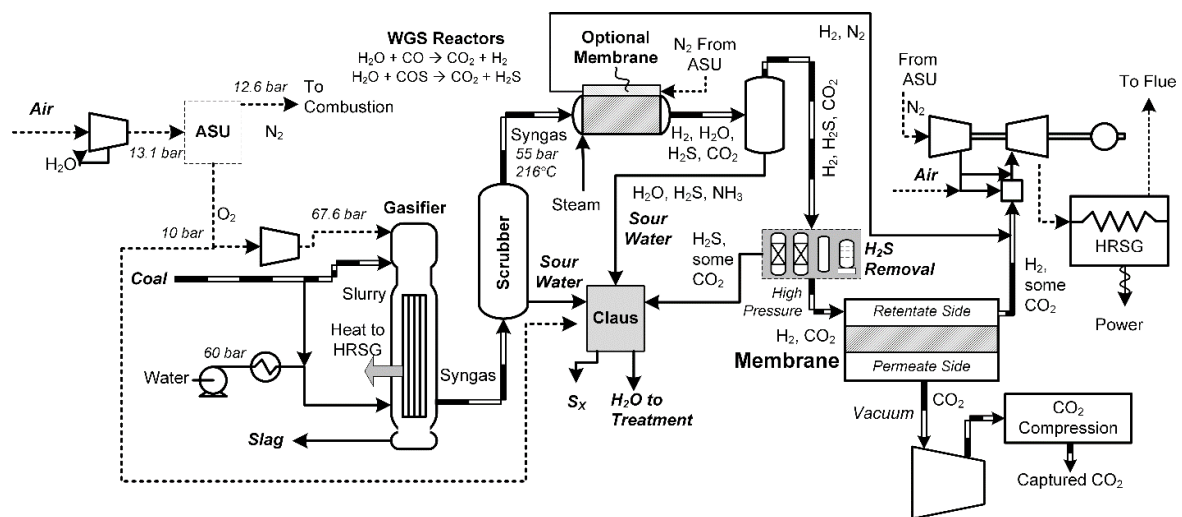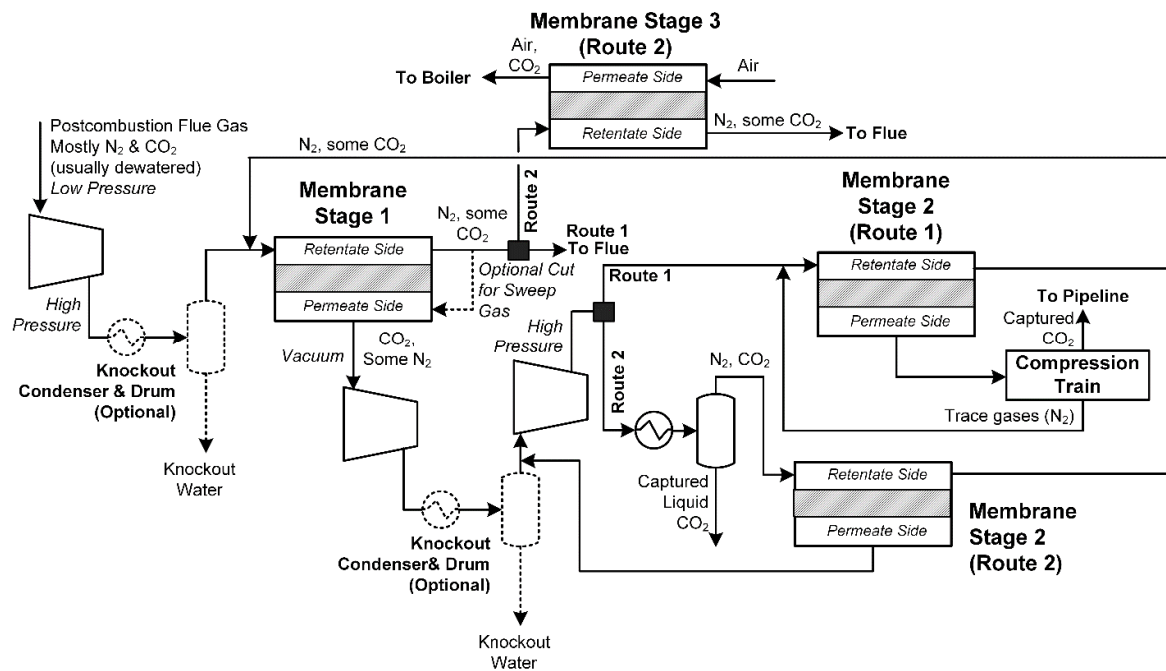
Figure 9. A superstructure of pre-combustion membrane separation processes for coal.

Alternatively, post-combustion membranes are used to remove CO_2 from combustion flue gases instead of post-combustion solvent-based capture. They may also be used to enhance solvent-based systems as well. A superstructure of common designs surveyed in this study is shown in Figure 10. Typically, the dewatered combustion flue gases (containing primarily N_2 and CO_2) are compressed from near-atmospheric pressure to a high pressure and sent to a membrane for CO_2/N_2 separation, which may or may not use a sweep gas in the permeate side. A vacuum is drawn on the permeate side to recover CO_2 at low pressure, and so this gas must be recompressed back up to high pressure again. Depending on the nitrogen content and process design, the permeate can either be sent to a second stage membrane for additional CO_2 purification (called Route 1 in Figure 10) or sent directly to CO_2 liquefaction (Route 2). In Route 2, an optional membrane could be used to partially separate the gases remaining from CO_2 liquefaction, with appropriate recycles. Additionally, it may be desirable to use an additional membrane to remove additional CO_2 from the retentate of the first stage membrane (Route 2) using boiler feed air as a sweep gas. This will remove additional CO_2 and increase the CO_2 content of the flue gas, resulting in a potentially greater CO_2 recovery rate and potentially better separation in the first stage membrane due to higher CO_2 concentrations. Although post-combustion membrane systems avoid solvent-based separation, one key disadvantage of these systems is the large compression load due to large pressure drops across the membranes.

Lastly, with the wide variety of process approaches and differences in implementation for all of the above technologies, it is difficult to gauge which technologies we should be pursuing now and in the future from the existing literature. Although many techno-economic studies of CO_2 capture processes often make comparisons with some base case to determine metrics such as LCOE or CO_2 emissions avoided, nearly all studies use different fuel costs, cost methods, base cases, and other assumptions which make a direct cross comparison of one study to another rather difficult. It is therefore the goal of this meta-study to examine a broad range of studies and processes in the literature and make informed comparisons between them using consistent metrics. Such a study has not been presented before in the literature to the best of our knowledge. For example, [5] has an excellent comparison of the CO_2 emissions of traditional post-combustion, IGCC, and oxyfuel plants, but it does not include SOFC or membrane technologies, nor does it have any indication of costs either. The present study, however, includes these technologies and not only examines life cycle CO_2 emissions, but the relative costs of avoiding them using a consistent basis for all technologies.

Figure 10. A superstructure of post-combustion membrane separation processes for coal or gas. Blocks with dashed lines are optional depending on the circumstances. Dark solid squares indicate points at which a design will typically use either Route 1 or Route 2, but not both.

2. Research Methods

In this work, we surveyed and examined techno-economic studies of power plant systems with CO_2 capture capabilities published since 2007 (for a review of CO_2 capture technologies prior to 2007, see [12,51]). The following CO_2 capture strategies were considered to be classified by these types: traditional combustion with solvent-based post-combustion CO_2 capture, IGCC with pre-combustion CO_2 capture, oxyfuels, membrane-based processes, and solid oxide fuel cells. Since none of these technologies has been fully commercialized, all of these studies are theoretical in nature, and are typically based on first-principles chemical process modelling and simulation. Although each study considered provided some information concerning basic economic and performance metrics (costs, efficiencies, CO_2 capture rates, etc.), the assumptions and methodologies used to compute these metrics differ from study to study. Moreover, the processes' objectives varied from study to study, such as the total power produced, the amount of CO_2 captured, and the quality and pressure of the captured CO_2. Therefore, in this work, we used a normalization strategy in order to predict the costs and performance of each process using a consistent basis of comparison.

2.1. The Standard "Status Quo" Plant

In this work, we use a single "status quo" plant design for the base case for comparison in the coal and natural gas categories. In the coal category, the base case plant is a supercritical pulverized coal power plant without CO_2 capture. For natural gas, the base case used is a natural gas combined cycle power plant. Both adequately represent commercially mature conventional power plant designs in use today. All studies assume baseline-only operation. The key characteristics for these plants are shown in Table 1. These plants were chosen as the basis of comparison because of its clarity of methods, availability of details, consistency in net power output at a reasonable scale (550 MW), as well as using a consistent approach for comparing both coal and gas fuels in the modern US market. Costs were converted to $US2016 using the methods described later in this section.

Table 1. Base cases for coal and gas, representing the current commercial state-of-the-art.

Property	Supercritical Pulverized Coal Base Case	NGCC Base Case	Units
Fuel Type	Bituminous Coal	Conventional Pipeline Natural Gas	
Location	United States	United States	
Project Year	1st Quarter 2016	1st Quarter 2016	
Levelized Cost of Electricity (As Reported)	82.3	65.0	$US2011 per MWh
Levelized Cost of Electricity (Adjusted to $2016)	81.1	48.4	$US2016 per MWh
Net Plant Power Output	550	550	MWe (AC, grid quality, at plant gate, prior to transmission)
Electrical Efficiency *	40.7	48.7	HHV %
Direct CO_2 Emissions	868.8	373.0	$kgCO_2e$/MWh of electricity
Indirect CO_2 Emissions [†]	75.6	67.3	$kgCO_2e$/MWh of electricity
Cradle-To-Plant-Exit CO_2 Emissions	944.4	440.3	$kgCO_2e$/MWh
CO_2 Captured	0%	0%	
Fuel Price	2.47	3.37	$US2016 per GJ
Fuel Price (as reported)	2.78	5.69	$US2011 per GJ
Non-Fuel Costs	59.3	23.6	$US2016 per MWh
Non-Fuel Costs (as reported)	57.7	22.9	$US2011 per MWh
CCA by switching from SCPC base case to NGCC base case (with no carbon capture)		−64.8	$US2016/$tCO_2e$
Reference	[6]	[52]	

* Electrical efficiency is defined as the net electricity produced (AC, grid quality) at the plant gate prior to transmission divided by the higher heating value of the fuel consumed, including any parasitic loads within the plant boundary. [†] These are the cradle-to-gate-entrance CO_2 emissions of producing, processing, and delivering the fuel to the plant. See Section 2.4 for details.

2.2. Adjusting for Scale

All plants surveyed were converted to have the same net power output as the base cases shown in Table 1 (550 MW). The non-fuel portion of the costs primarily consist of capital costs, but also consist of other factors such as maintenance and overhead, labour, spare parts, working capital, and accounts receivable [53]. It is typical to estimate the costs of these other factors as percentages of the total equipment cost [54], and as such, they are all relatively proportional to each other. Because it is a power plant, raw material costs are insignificant compared to the capital or fuel costs. Therefore, a power law adjustment rule (commonly known as the "six-tenths rule") was used to estimate the costs of the non-fuel portion at the new scale according to the following formula:

$$C_{nonfuel,scaled} = C_{nonfuel,\ original} \left(\frac{NPO_{scaled}}{NPO_{original}} \right)^{p-1} \tag{1}$$

where $C_{nonfuel,scaled}$, is the cost of the non-fuel portion at the standard size shown in Table 1, $C_{nonfuel,\ original}$ is the cost of the non-fuel portion as reported in the original work (in the given currency per MWh of net power produced), NPO_{scaled} and $NPO_{original}$ are the net power output at the standard size and as reported in the original work (in MW), respectively, and p is the power law exponent. Commonly, $p = 0.6$ is assumed (hence "six-tenths" rule) when there is a lack of information about the economies of scale for a particular application [54], which derives from historical data for the industry in general. However, in this work, we have adopted $p = 0.9$ as our base case assumed value, based on the studies of Hamelinck et al. [55], which estimated p for a wide variety of chemical process equipment relevant to energy and electricity production to be in the range of 0.6 to 1.0, with a tendency on the higher side. For example, they suggested $p = 0.7$ for gasifiers, Selexol-based pre-combustion CO_2 capture systems, heat recovery and steam generators, gas turbines, and combined cycled power plants. Larger p was recommended for other relevant process operations such as solids handling systems (e.g., conveyors, dryers, and solid feeding systems), as well as air separation, oxygen compression systems, and dry gas cleaning. Moreover, when a plant is "scaled up" by a large amount, the sensitivity of key metrics like levelized costs of electricity (LCOE) to p is not insignificant. Therefore, we have conducted a sensitivity analysis on this assumption, as described in Section 3.4, but noting that the final conclusions were essentially unchanged with regard to p.

We assume that the efficiency of a given power plant is the same at the standard size as it is at the size of the plant originally reported, since all plants surveyed were within the same order of magnitude (approximately 200 to 1000 MW net power output). As such, the amount of fuel used scales linearly with size, and the amount of fuel required per MWh of electricity produced does not change with size. Expressed in equation form, this becomes simply:

$$C_{fuel,scaled} = C_{fuel,\ original} \tag{2}$$

where $C_{fuel,scaled}$ and $C_{fuel,\ original}$ are the costs of fuel in the given currency per MWh of net power produced at the standard size and at the size of the plant as originally reported, respectively.

For some studies, the fuel and non-fuel costs were not itemized or reported separately, but either the total costs or the LCOE were reported. In these cases, the fuel costs and non-fuel costs could be approximated using the reported efficiency of the power plant and the unit price of fuel. For example, for a case where the efficiency of the plant, LCOE, and fuel prices were reported, the fuel and non-fuel costs could be computed as follows:

$$C_{fuel,\ original} = \left(\frac{3600\ MJ/MWh}{\in_{original}} \right) U_{fuel,original} \tag{3}$$

$$C_{nonfuel,\ original} = LCOE_{original} - C_{fuel,\ original} \tag{4}$$

where $\in_{original}$ is the reported electrical efficiency (HHV), $U_{fuel,original}$ is the unit price of fuel as reported (in the reported currency per HHV of fuel in MJ), and $LCOE_{original}$ is the levelized cost of electricity reported for the original plant (in the reported currency per MWh of net power output). Note that the term in parenthesis in Equation (3) is commonly known as the heat rate, which is sometimes reported directly. In some cases, the reported efficiency was given in lower heating value (LHV) and so was converted to HHV for this purpose.

It is also important to note that LCOE definitions may vary from study to study, such as in using different assumed capacity factors, plant lifetimes, discount interest rates, or whether a constant-dollar or current-dollar approach was used. In most of the works included in the meta-study, these parameters were not provided and only LCOE was reported directly. Since there was insufficient data to convert reported LCOE forms to a standard LCOE form using a standard set of parameters, and the fact that the change in the values would be small anyway, we have taken all reported LCOE numbers at face value, assuming that the authors used reasonable values for these parameters.

Studies which did not have sufficient information to estimate both the fuel and non-fuel cost according to the above procedure were not included in this meta-study.

2.3. Adjusting for Currency Type, Location, and Year

The studies surveyed were conducted using different currency types, countries of construction, and currency years. Reported costs in the literature were converted as follows. First, when non-fuel costs were reported for locales outside the United States, the historical Purchasing Power Parity Index (PPPI) [56] was used to estimate the costs of an equivalent plant if it were located in the United States, instead. This includes both the currency exchange rates as well as relative differences in costs of materials, products, and services. Then, the non-fuel costs were scaled to \$US2016 by using the IHS North American Power Capital Costs Index, which is a quarterly capital cost index for the construction of new power plants within North America [57]. The resulting equation is as follows:

$$C_{nonfuel,scaled,standard} = C_{nonfuel,scaled} \left(\frac{PCCI_{2016Q1}}{PCCI_{originalyear}} \right) \left(PPPI_{country,originalyear} \right) \tag{5}$$

where $C_{nonfuel,scaled,standard}$ is the nonfuel cost in \$US2016 per MWh of electricity produced, $PCCI$ is the Power Capital Costs Index of either the first quarter of 2016 or the average index for the year of the original study, and $PPPI$ is the Purchasing Power Parity Index for the country of original work for the year of the original study with units of the local currency per \$US in the year of the original study. Note that studies that used a United States local use the PPPI for the United States, which is equal to 1 \$US/\$US by definition.

A similar approach was used for fuel costs, except that instead of using PPPI to convert prices to a US locale, a more specific energy index was used instead. This is because the PPPI and relative energy prices are often vastly different, despite the fact that energy prices do influence the PPPI to some degree. For example, the price of imported natural gas in Germany in 2012 (when converted to US Dollars) was nearly four times the US Henry Hub price [30], but the relative price of goods in general using the PPPI between Germany and the US in 2012 was very close to 1 [56]. When the fuel costs or unit price of fuel was provided in the original study, it was first localized to the United States in \$US in the base year using relative pricing information published by BP [58] and the currency exchange rate [59], and then scaled to the current year by using the ratio of the historical fuel price index for North America based on price history data from the US Energy Information Administration [60]. The equation is as follows:

$$C_{fuel,scaled,standard} = C_{fuel,scaled} \left(E_{currency,originalyear} \right) \left(BPR_{originalyear} \right) \left(\frac{ECI_{2016Q1}}{ECI_{originalyear}} \right) \tag{6}$$

where $C_{fuel,scaled,standard}$ is the fuel cost at the standard size (in \$US2016 per MWh of net electricity produced), $E_{currency,originalyear}$ is the historical currency average exchange rate for the project year of the original study in \$US per local currency unit, $BPR_{originalyear}$ is the relative price ratio of fuel (the average price of coal or gas in the US divided by the price of the same amount of fuel in the country of the study in the \$US for the year of that study), and *ECI* is the energy cost index for either bituminous coal or conventional pipeline natural gas for the North American market, for either the first quarter of 2016, or the average of the base year. In some cases, the project year of the study was not reported, and so the year of publication of the study was used as an approximation instead. The LCOE of each plant for the standard size and year ($LCOE_{scaled,standard}$) in units of \$2016 per MWh of net electricity produced becomes:

$$LCOE_{scaled,standard} = C_{fuel,scaled,standard} + C_{nonfuel,scaled,standard} \qquad (7)$$

Note that the LCOE computed does not consider distribution costs, profits/markups, or taxes.

In addition, we considered a second set of assumptions, in which the unit price of fuel for 1Q2016 in the North American market was used directly [60] and the originally reported fuel price was ignored. The reason for this was that fuel is a global commodity with readily tracked and recorded prices based on specific markets, and it would make sense to use the same price of fuel for each case when comparing many different kinds of power plants to each other on the most consistent basis possible, including the geographical location of construction. In some studies, the assumed fuel prices were far from both known historical market prices for their respective markets, and far from the assumptions made by other researchers (see the Section 3.2 discussion on outliers), and thus were not necessarily good values to include in the meta-analysis.

However, it is also the case that most techno-economic analysis studies are performed with some sort of optimization approach (even informally), in which the trade-offs between the fuel and non-fuel costs are balanced based on an economic objective function. For example, in cases where fuel prices were assumed to be high, researchers would be more likely to make design decisions that favour more expensive but more fuel-efficient designs, and vice versa. Therefore, the non-fuel portion of the costs as calculated in each study are in fact very likely related to the assumed fuel price. Therefore, it is not clear if the fuel prices used in the meta-analysis should be all the same in order to have a constant basis for comparison, or, based on the originally assumed prices in the cited studies in order to preserve the fuel vs. non-fuel trade-offs that factored into design decisions. Thus, both approaches were used in order to draw more general conclusions, and to serve as a sensitivity analysis on fuel price. Lastly, in a few cases, there was not enough information provided in a study to infer fuel price assumptions, and as such only the standard 1Q2016 North American energy price approach could be used.

2.4. Adjusting for Differences in CO_2 Capture Rates, Pressures, and Plant Gate Definitions

For this study, we defined the "plant gate exit" to include the compression of captured CO_2 up to at least 115 bar pressure, and its purification to minimum pipeline purity standards (at least 95 mol % pure CO_2 [61]). All of the studies included in the meta-analysis fell into this standard, although in some cases, the specific pressure was not reported but it was noted that the outlet pressure did meet non-specific pipeline specifications (which are typically at 115 bar or above). The CO_2 pressure of the studies surveyed ranged from 115 to 154 bar. Because CO_2 is supercritical in that pressure range, pressure increases from 115 up to 154 bar can be achieved using a pump, which is relatively inexpensive compared to the costs of a compressor. For example, the energy required to compress and pump CO_2 from 18.5 bar to 115 bar is over 16 times the energy requirement of pumping supercritical CO_2 from 115 to 154 bar [25]. Therefore, differences between processes due to differences in CO_2 compression pressure were neglected (since all studies cited considered capturing CO_2 at pipeline pressure) because the impact on the LCOE is negligible.

The percentage of CO_2 captured varies in each plant in the study. Typically, between 90 and 95 percent CO_2 created by fuel oxidation was captured for the solvent-based and membrane-based designs. The oxyfuel and SOFC designs typically had CO_2 capture rates approaching 100%. Because these differences in capture rates are case-specific and typically linked to a particular design or technology, we did not attempt to alter the cost estimates for each plant to meet a particular CO_2 capture percentage.

For standard comparison in this study, the LCOE used for cost comparisons do not include CO_2 transport and storage, since those are likely to depend heavily on geography and end-use. All studies considered in this work use reported costs that do not consider transportation and storage. In some cases, this required manual recalculation of either LCOE and/or non-fuel costs to make this adjustment.

2.5. Computing Costs of CO_2 Avoided

The total cradle-to-plant-exit lifecycle greenhouse gas emissions were computed as the sum of the direct emissions of the plant itself plus the indirect emissions associated with the production and delivery of the fuel. The total direct emissions of the plant itself were usually reported in each study surveyed in terms of global warming potential in units of CO_2 equivalents (CO_2e). However, because these studies are simulation based, other greenhouse gases such as NO_x are either ignored or are present in very small quantities, such that the total direct emissions of CO_2 (the chemical) is essentially equal to the total global warming potential in terms of CO_2e.

For studies in which the direct CO_2 emissions were not reported, they were estimated by using other information available. For example, given a known CO_2 capture percentage and a known efficiency, the CO_2 emissions rate could be computed by assuming that all fuel is completely oxidized such that all carbon atoms leave as CO_2. For some studies, the direct emissions were not reported in terms of CO_2e, but rather the emissions of the individual species of chemicals were reported instead. In this case, this information was used to compute the total global warming potential using the International Panel on Climate Change 5th Assessment 100-year metric (i.e., 1 kg of N_2O emissions = 298 kg of CO_2e) [62]. In some cases, CO_2 emissions were reported, but the effects of NO_x were not considered, resulting in an underestimate of the total global warming potential. In these cases, the NO_x emissions were estimated by assuming 0.318 kg NO_x/MWh produced for SCPC plants and 0.18 kg/MWh and IGCC plants [63]. For calcium looping studies, these numbers are applied to the upstream portions of the power plant (either SCPC or IGCC) only; it is assumed that no NO_x is produced in the calciner because of the very small presence of N_2.

The indirect emissions associated with the cradle-to-plant-entrance production and delivery of coal was 30.8 kg CO_2e per MWh of coal delivered [64]. This number includes the emissions associated with mining and transporting bituminous coal 100 km to the plant entrance. Interestingly, almost two-thirds of the greenhouse gas potential is due to methane emissions (such as from fugitive coal-bed methane), not CO_2 itself [22]. Similarly, the indirect emissions for natural gas was 32.8 kgCO_2e per MWh of gas delivered [65] by assuming conventional pipeline gas made from the North American average mixture of gas sources (conventional, associated gas, coal bed methane, shale gas, tight gas, off-shore gas, and imported liquefied natural gas). The indirect emissions of electricity transportation (e.g., high-tension transmission lines) or CO_2 transport and storage were not included, as they are downstream of the plant in question and are specific to each unique implementation. Emissions associated with the construction of the power plants themselves were neglected, as prior studies found that the construction contributions for SCPC, IGCC, and SOFC-based plants were a very small percentage of the total emissions [22,23].

The cost of CO_2 avoided (CCA) is a useful metric for comparing unlike processes in terms of finding which process is the most effective way to spend money for the explicit purposes of reducing CO_2 emissions. The CCA is generally defined in reference to a base case plant used as the status quo or business-as-usual case, and is essentially the extra costs that one spends on a process or activity for

the specific purposes of avoiding CO_2 emissions caused by the base case divided by the amount of CO_2 emissions you avoid as a result. This results in the following equation:

$$CCA = \frac{LCOE_{scaled,standard} - LCOE_{basecase}}{GWP_{basecase} - GWP_{scaled,standard}} \tag{8}$$

where $LCOE_{basecase}$ and $GWP_{basecase}$ are the LCOE (in \$US2016 per MWh) and cradle-to-plant-exit global warming potential (in tCO_2e per MWh of net power output) of the base case (in Table 1), and $GWP_{scaled,standard}$ is the cradle-to-plant-exit global warming potential (in tCO_2e per MWh of net power output) of the plant. In this study, we have chosen to use two base cases, one for coal, and one for gas. As such, we report two CCAs, each in relation to the SCPC case (without CO_2 capture) or the NGCC base case (without CO_2 capture) in Table 1. Note that Table 1 also contains a calculation of the CCA when building the NGCC base case (without CO_2 capture) instead of the SCPC base case. This value is negative, meaning that NGCC has both a lower LCOE than SCPC and lower lifecycle CO_2 emissions than SCPC. In other words, with current fuel prices and construction costs, there is little incentive to choose an SCPC plant over an NGCC plant from either an economic or an environmental perspective.

The above CCA definition does not determine the CCA of retrofitting a CO_2 capture system onto an existing power plant. It also does not compare the CCA by building a power plant of an advanced technology such as SOFCs with CO_2 capture compared to the advanced technology without CO_2 capture. This is important to note because many studies present CCA numbers using the latter definition. Although that definition of CCA is interesting for that specific study, it cannot be used for cross-comparison across studies.

Lastly, it is important to note that although CCA is a useful metric for understanding the cost trade-offs in terms of a reduction in the global warming potential, it does not consider any other environmental effects such as acid rain, eutrophication, smog, ozone depletion, etc., and so it is not a substitute for a holistic picture of environmental impact. However, the CCA metric is immediately relevant to policy decisions with regard to carbon management and carbon taxes.

3. Results

3.1. Raw Data

The coal cases are summarized in Table 2. The table includes a classification of the process into one of five categories (SCPC, IGCC, COXY = coal-based oxyfuel combustion, IGFC = integrated gasification solid oxide fuel cell, and CMEM = coal-based membrane approaches), with the primary CO_2 capture approach described briefly. The base year is the project year as reported in the study. The LCOE is given both as reported and as converted to the standard plant conditions by the methods described in Section 2. The CCA is given using both SCPC and NGCC baselines (in Table 1) as the base plant for comparison, at the standard plant conditions as calculated according to the methods of Section 2. The lifecycle CO_2 emissions are for the cradle-to-plant-exit and calculated as described in Section 2.

Table 2. Coal-fired power plants surveyed in the meta-study, with key metrics shown. For other metrics and intermediate calculations, please see the Microsoft Excel file included with the supplementary material. Notes: tCO$_2$e is in metric tonnes. LCU = local currency unit. Converted to Standard means that the plant has been converted to the standard size, project year, and locale according to the methods described in Section 2.

Tag	Ref.	Project Year, LCU	Technology	Reported Capacity (MW)	Reported Efficiency (%HHV)	LCOE (Using As-Reported Fuel Prices) Reported (LCU/MWh)	LCOE (Using As-Reported Fuel Prices) Converted to Standard ($/MWh)	CO$_2$ Captured (%)	Lifecycle CO$_2$ Emissions (tCO$_2$e/MWh)	CCA (Converted to Standard Form Using As-Reported Fuel Prices, $US2016/tCO$_2$e) Compared to SCPC Baseline	CCA Compared to NGCC Baseline
SCPC-1	[9]	$2011	Solvent Based—EconFG+	550	32.5	133.2	132.6	90	0.290	78.7	561.1
SCPC-2	[66]	$2013	Solvent Based	525	27.4	129.5	125.3	90	0.327	71.6	678.7
SCPC-3	[67]	€2010	Solvent Based—Advanced amine	837	36.1	73	91.8	90	0.275	16.0	262.6
SCPC-4	[68]	€2013	Solvent Based—Cansolv	822	33.8	86.3	107.9	90	0.279	40.2	368.0
SCPC-5	[69]	$2010	Solvent Based—Amine	546	27.2	122	123.8	90	0.323	68.7	643.5
SCPC-6	[70]	€2010	Solvent Based—Advanced Amine	616	36.5	70.3	81.8	90	0.271	1.1	197.4
SCPC-7	[71]	$2006	Solvent based-Chilled Ammonia	548.7	28.4	80.8	95.0	90	0.314	22.0	367.4
SCPC-8	[71]	$2006	Solvent based-Chilled Ammonia	549.3	31.5	74.7	87.9	90	0.292	10.4	266.5
SCPC-9	[14]	$2007	Solvent based-Chilled Ammonia	558.7	27.9	100.2	102.0	90	0.325	33.7	464.2
SCPC-10	[12]	€2008	Solvent based-Chilled Ammonia	614	36.6	82.3	104.3	88.4	0.276	34.7	339.9
SCPC-11	[72]	£2011	Solvent Based-Amine	519	25.6	102	135.9	94	0.247	16.2	227.1
IGCC-1	[73]	$2011	Solvent Based-Selexol	543	32.6	135.4	134.7	90	0.291	83.8	585.2
IGCC-2	[73]	$2011	Solvent Based-Selexol	514	31	141.9	140.5	90	0.241	76.2	432.7
IGCC-3	[73]	$2011	Solvent Based-Selexol	497	31.2	141.9	140.2	90	0.252	85.8	488.7
IGCC-4	[66]	$2013	Data not available	500	29.9	152.6	147.7	86	0.250	85.1	482.6
IGCC-5	[69]	$2010	Solvent Based-Selexol NS	694	32	90	92.4	90	0.307	104.4	742.1
IGCC-6	[72]	£2011	Solvent Based—Shift + Selexol	455	35.1	79.7	105.6	94	0.197	32.8	235.4
CMEM-1	[74]	$2010	Membrane—2 stage	550	25.7	112	113.7	90	0.336	53.5	628.1
CMEM-2	[75]	€2008	Membrane—3 stage	754	33	94	122.0	90	0.293	62.7	498.3
CMEM-3	[76]	$2010	Membrane—Two stage two step with air sweep	550	25.7	94.1	94.3	90	0.339	21.8	451.2
CMEM-4	[77,78]	$2007	H$_2$-CO$_2$ selective membrane with sweep gas 2 stages membrane with argon as sweep gas	556	31	90.48	92.3	90	0.254	16.2	236.1
CMEM-5	[79]	$2007	(Amine-containing—membranes synthesized and tested)	550	31	100	101.4	90	0.254	29.5	285.2
CMEM-6	[80]	€2015	Cascade Membrane system-polymeric	439	23.38	85.9	95.3	90	0.352	23.8	528.0
CMEM-7	[80]	€2015	3 end membrane integration-ceramic	492	33.06	77.3	88.6	90	0.271	11.2	237.7
CMEM-8	[80]	€2015	Microporous ceramic membrane—H$_2$ selective	316	30.37	82	91.4	90	0.251	14.8	226.6
CMEM-9	[80]	€2015	Microporous ceramic membrane—soar WGS membrane	316	33.01	80.6	90.3	90	0.231	12.9	200.3
CMEM-10	[81]	€2015	Integrated WGS membrane reactor	286	32	86	104.6	81	0.331	38.3	513.1
CMEM-11	[82]	€2015	Two stages membrane	350	24.9	107.2	123.3	90	0.351	71.1	841.2
CMEM-12	[83]	€2014	Membrane-cryogenic	427	33.82	65.5	80.0	90	0.284	−1.8	201.7
CCRY-1	[15]	€2012	Cryogenic	699	33.06	99.7	124.3	90	0.286	65.6	491.2
COXY-1	[84]	€2015	Oxy-fuel	475	34.31	70.06	82.3	91	0.160	1.5	120.9
COXY-2	[84]	€2015	Oxy-fuel	475	32.2	88.54	105.1	90	0.178	31.4	216.3

Table 2. *Cont.*

| Tag | Ref. | Project Year, LCU | Technology | Reported Capacity (MW) | Reported Efficiency (%HHV) | LCOE (Using As-Reported Fuel Prices) | | CO$_2$ Captured (%) | Lifecycle CO$_2$ Emissions (tCO$_2$e/MWh) | CCA (Converted to Standard Form Using As-Reported Fuel Prices, $US2016/tCO$_2$e) | |
						Reported (LCU/MWh)	Converted to Standard ($/MWh)			Compared to SCPC Baseline	Compared to NGCC Baseline
COXY-3	[85]	€2012	Oxy-fuel	826	34.72	71.7	88.1	94	0.136	8.6	130.3
COXY-4	[85]	€2012	Oxy fuel with 10% air added	826	34.67	71.4	87.6	94	0.137	8.1	129.0
COXY-5	[85]	€2012	Oxy fuel with 20% air added	826	34.51	71.3	87.5	94	0.137	7.9	128.8
COXY-6	[85]	€2012	Oxy fuel with 30% air added	826	34.25	71.4	87.6	94	0.138	8.0	129.5
COXY-7	[85]	€2012	Oxy fuel with 40% air added	826	33.88	71.7	87.9	94	0.140	8.4	131.2
COXY-8	[85]	€2012	Oxy fuel with 50% air added	826	33.41	72.5	88.8	94	0.142	9.6	135.3
COXY-9	[86]	€2007	Oxy-fuel	357	32.29	57.5	69.8	90	0.231	−15.8	102.2
COXY-10	[87]	$2009	Oxy-fuel	601	32.74	82.3	85.4	93	0.147	5.4	126.1
COXY-11	[88]	€2014	Oxy-fuel with compression and dehydration	400	33.81	116.3	140.0	100	0.091	69.0	262.1
COXY-12	[88]	€2014	Oxy-fuel with double flash purification	400	29.51	133.5	160.7	92	0.189	105.3	446.2
COXY-13	[88]	€2014	Oxy-fuel with distillation purification	400	28.99	139.8	168.4	90	0.214	119.4	529.0
COXY-14	[89]	€2009	Oxy-fuel	630	36.6	67.2	84.3	90	0.203	4.3	151.1
COXY-15	[68]	$2007	Oxy-fuel (Case: S12F)	550	31.2	96	94.7	90.8	0.196	18.2	189.6
COXY-16	[68]	$2007	Oxy-fuel (Case: S22F)	550	30.8	106.9	105.3	90.9	0.199	32.5	235.5
COXY-17	[66]	$2013	Oxy-fuel (Boiler type & Coal rank: USC-subbit)	509	31.5	119.8	116.3	90	0.203	47.4	285.5
COXY-18	[66]	$2013	Oxy-fuel (Boiler type & Coal rank: SCPC-bit)	510	31	121.4	117.8	98	0.120	44.6	216.9
COXY-19	[66]	$2013	Oxy-fuel (Boiler type & Coal rank: SCPC-bit)	501	31.5	118.7	115.0	90	0.203	45.7	280.3
COXY-20	[68]	€2013	Oxy-fuel (Boiler type & Coal rank: USC-subbit)	833	34.1	91.6	115.1	90	0.182	44.6	258.4
COXY-21	[90]	$2007	Oxy-combustion Supercritical PC with CO$_2$ Capture	550	29.3	100.7	102.2	99.5	0.105	25.1	160.4
COXY-22	[90]	$2007	Oxy-combustion Supercritical PC with CO$_2$ Capture	555	29.5	99.8	101.4	99.4	0.104	24.1	157.5
COXY-23	[90]	$2007	Oxy-combustion Supercritical PC with CO$_2$ Capture	549	29.3	101.8	103.3	96.9	0.132	27.2	177.7
COXY-24	[90]	$2007	Oxy-combustion Supercritical PC with CO$_2$ Capture	549	29.2	102.6	104.0	85.5	0.182	30.1	214.9
COXY-25	[90]	$2007	Oxy-combustion Ultra-supercritical PC with CO$_2$ Capture	550	33	95.9	97.0	99.4	0.093	18.7	140.0
COXY-26	[90]	$2007	Oxy-combustion Ultra-supercritical PC with CO$_2$ Capture	545	32.7	98.6	99.6	93.2	0.154	23.4	178.7
IGFC-1	[44]	$2011	Uses WGS to minimize CO content in SOFC feed	719	42	78	79.9	100	0.073	−1.3	85.8
IGFC-2	[44]	$2011	IGFC-1 but with seasonal SOFC shutdowns to preserve life during low-demand deasons	719	42	78.8	80.8	100	0.073	−0.4	88.1
IGFC-3	[44]	$2011	IGFC-1 with integrated energy storage	719	41.5	81.5	83.6	100	0.074	2.9	96.1
IGFC-4	[44]	$2011	IGFC-2 with integrated energy storage	719	40.6	82.4	84.5	100	0.076	3.9	99.0
IGFC-5	[44]	$2011	Uses coal syngas directly in SOFC with no WGS	719	38.4	80.8	82.7	100	0.080	1.8	95.1
IGFC-6	[44]	$2011	IGFC-5 with seasonal SOFC shutdowns	719	38.41	81.4	83.3	100	0.080	2.5	96.8
IGFC-7	[44]	$2011	IGFC-5 with integrated energy storage	719	38	84.3	86.3	100	0.081	6.1	105.5
IGFC-8	[44]	$2011	IGFC-5 with integrated energy storage	719	37.4	83.6	85.6	100	0.082	5.2	103.7
IGFC-9	[91]	$2007	Atm-Pressure IGFC Plant	253	49.4	88	82.5	99	0.064	1.6	90.4
IGFC-10	[91]	$2007	Pressurized IGFC	253	56.2	79	74.0	99	0.056	−8.0	66.5

Table 2. *Cont.*

Tag	Ref.	Project Year, LCU	Technology	Reported Capacity (MW)	Reported Efficiency (%HHV)	LCOE (Using As-Reported Fuel Prices)		CO$_2$ Captured (%)	Lifecycle CO$_2$ Emissions (tCO$_2$e/MWh)	CCA (Converted to Standard Form Using As-Reported Fuel Prices, $US2016/tCO$_2$e)	
						Reported (LCU/MWh)	Converted to Standard ($/MWh)			Compared to SCPC Baseline	Compared to NGCC Baseline
IGFC-11	[92]	$2007	Partially methanated Syngas IGFC (TREMP)	847	48.32	96.3	100.1	99	0.066	21.6	138.1
IGFC-12	[92]	$2007	Partially methanated Syngas IGFC (HICOM)	925	52.81	84.8	88.9	99	0.065	8.8	107.6
IGFC-13	[92]	$2007	IGFC-DIRECT	865	49.36	90.6	94.4	99	0.069	15.2	123.9
IGFC-14	[93]	$2006	Liquid-tin anode SOFC	93.9	57.2	71	75.9	100	0.054	−5.9	70.9
CaL-1	[48]	€2005	CaO Looping	621	27.87	48.3	105.2	85	0.298	37.2	397.8
CaL-2	[49]	€2009	CaO Looping w/o heat recovery	446	18.59	102	142.8	85	0.476	131.8	−2650.7
CaL-3	[49]	€2009	CaO Looping with Combined Cycle Design 1	594.8	24.24	59.5	84.7	85	0.410	6.7	1179.7
CaL-4	[49]	€2009	CaO Looping with Combined Cycle Design 2	446	32.67	71	99.8	85	0.326	30.2	447.9
CaL-5	[49]	€2009	CaO Looping with Combined Cycle Design 3	594.8	33.54	55.5	79.6	85	0.323	−2.5	264.3
CaL-6	[49]	€2009	CaO Looping with Combined Cycle Design 4	846	35.26	54.3	80.2	85	0.337	−1.5	308.8
CaL-7	[49]	€2009	CaO Looping with Combined Cycle Design 5	835.3	34.78	54.5	80.4	85	0.340	−1.2	317.5
CaL-8	[50]	$2011	CaO Looping	932.9	33.10	90.68	94.9	99	0.129	16.9	149.5
CaL-9	[47]	€2013	Sub-critical pulverized coal with CaO Looping	558.9	32.06	73.77	88.9	92.25	0.258	11.4	222.2
CaL-10	[47]	€2013	SCPC with CaO Looping	544.22	35.91	68.41	82.5	92.66	0.238	2.0	168.6
CaL-11	[47]	€2013	IGCC with CaO Looping	558.08	35.97	76.48	93.0	92.95	0.199	16.0	184.7
CaL-12	[94]	$2012	SCPC with CaO Looping	1060	36.00	136.85	143.8	90	0.236	88.6	467.8
CLC-1	[95]	$2003	Chemical Looping Combustion	164.484	30.88	58.4	99.5	90–99	0.132	22.7	165.6
CLC-2	[95]	$2003	Chemical Looping Gasification	256.83	36.90	52.2	92.4	90–99	0.124	13.7	139.0
CLC-3	[96]	$2007	Syngas Chemical Looping	346.8	34.70	100.7	103.9	90	0.229	31.9	262.2
CLC-4	[97]	$2011	Coal Direct Chemical Looping (CDCL) Retrofit to Pulverized Coal Power Plants for In-Situ CO$_2$ Capture	548	35.20	78.4	78.3	97	0.115	−3.4	91.9
CLC-5	[98]	$2011	Alstom Limestone Chemical Looping plant using Fast CFB transport reactors	550	35.80	96.7	95.4	97	0.104	17	139.7

Table 3. Gas-fired power plants surveyed in the meta-study, with key metrics shown. For other metrics and intermediate calculations, please see the Microsoft Excel file included with the supplementary material. Notes: tCO_2e is in metric tonnes. LCU = local currency unit (given in third column). Converted to Standard means that the plant has been converted to the standard size, project year, and locale according to the methods described in Section 2.

| Tag | Ref. | Project Year, LCU | Technology | Reported Capacity (MW) | Reported Efficiency (%HHV) | LCOE (Using As-Reported Fuel Prices) | | CO_2 Captured (%) | Lifecycle CO_2 Emissions (tCO_2e/MWh) | CCA (Converted to Standard form Using As-Reported Fuel Prices, $US2016/tCO_2e$) | |
						Reported (LCU/MWh)	Converted to Standard ($/MWh)			Compared to SCPC Baseline	Compared to NGCC Baseline
NGCC-1	[9]	$2011	Solvent Based—Econamine FG+	559	45.7	83.3	65.7	90	0.115	−18.6	53.0
NGCC-2	[99]	€2011	Solvent Based—MEA	789	46.1	74.8	54.8	90	0.114	−31.6	19.6
NGCC-3	[99]	€2011	Solvent Based—Advanced Amine	804	47	68.9	47.8	90	0.112	−40.0	−1.9
NGCC-4	[100]	$2007	Solvent Based—Econamine FG+	448.9	42.6	84.2	50.6	90	0.123	−37.1	7.0
NGCC-5	[100]	$2013	Solvent Based—Amine	485	42.4	112	94.8	90	0.122	16.7	145.9
NGCC-6	[72]	£2011	Solvent Based—Amine	389	40.8	108.41292	80.3	94	0.110	−1.0	96.6
NMEM-1	[101]	$2012	Solvent supported CO_2 membrane	376.9	48	63.8	57.4	90	0.112	−28.5	27.3
NOXY-1	[102]	$2011	Oxy-fuel	227.6	31.22	123.7	104.3	90	0.167	29.8	204.4
NGFC-1	[103]	$2007	SOFC base case ($1000/kW installed)	693	74	64.5	45.2	100	0.044	−39.9	−8.1
NGFC-2	[103]	$2007	SOFC ($500/kW installed half-capital)	693	74	57.4	35.6	100	0.044	−50.5	−32.3
NGFC-3	[103]	$2007	SOFC with 100% increase in operating	693	74	67.3	54.1	100	0.044	−30.0	14.4
NGFC-4	[103]	$2007	SOFC with 25% increase in capital cost	693	74	70.4	55.9	100	0.044	−28.0	18.9
NGFC-5	[103]	$2007	SOFC with 50% increase in capital cost	693	74	76.3	63.1	100	0.044	−20.0	37.0
NGFC-6	[104]	$2007	Low pressure autothermal reforming (ATR)-six parallel SOFC sections.	550	56.3	77.9	60.4	NA	0.059	−23.4	31.3
NGFC-7	[104]	$2007	Low pressure ATR-eight parallel SOFC sections. ASU oxidant capacity is 58% smaller than the previous case	550	61.8	62.6	46.9	NA	0.053	−38.4	−3.9
NGFC-8	[104]	$2007	High pressure ATR	550	64.8	67.8	52.6	NA	0.055	−32.1	10.8
NGFC-9	[104]	$2007	High pressure ATR-ASU oxidant capacity is 58% smaller than the previous case	550	64.8	65.3	50.2	NA	0.055	−34.8	4.5
NGFC-10	[104]	$2007	No ATR	550	65.9	61.2	46.4	NA	0.050	−38.8	−5.1
NCLC-1	[105]	$2002	NG Chemical Looping Combustion	341	46.9	41.2	52.7	98	0.077	−32.8	11.7

The natural gas cases are summarized in Table 3. The four categories included were NGCC, NOXY = natural gas based oxyfuels, NMEM = natural gas based membranes, and NGFC = natural gas solid oxide fuel cells. There were far fewer cases for natural gas available in the recent literature that contained the necessary information required for the meta-study. This is primarily because most studies on CO_2 capture focus on coal since it has a much larger CO_2 intensity compared to gas, and there is less incentive to use CO_2 capture in natural gas power plants. For example, in the United States, federal limits on the CO_2 emissions of new power plants were set by the US Environmental Protection Agency in 2015 at 1400 lb CO_2e per MWh for SCPC (635 kg CO_2e/MWh) and 1000 lb CO_2e per MWh for NGCC (454 kg CO_2e/MWh) [106], which is specifically chosen such that coal-based power plants would require CO_2 capture, but state-of-the-art natural gas plants would not. It is possible that these standards for new plants will be modified or removed under the new administration, since the rule is now under official review [107], and the proposed rule for existing plants (which was never formally adopted) has been officially dropped [108]. In addition, a recent set of studies of cradle-to-grave life cycle impacts for coal and natural gas combustion plants with post-combustion capture showed that while adding CO_2 capture to a SCPC plant reduces cradle-to-grave lifecycle CO_2 emissions by about 74%, adding CO_2 capture to an NGCC plant reduces them by only about 67% [22,23].

3.2. Identifying Outliers

Due to the broad nature of assumptions and methodologies used in the studies surveyed, it is necessary to identify outliers that may result in potentially incorrect conclusions. Some can be identified visually. Figure 11 shows a plot of the unit fuel prices and non-fuel costs for coal-based power plants as reported by each study, scaled to the standard conditions according to the methods of Section 2. The dashed rectangle is the visually identified region of the set of normal results, with the points lying outside classified as outliers. Note that no formal method was used to identify outliers, except that these points had unit fuel prices more than ±35% from the mean fuel unit price of \$1.97/GJ \$US2016 (note the average 2016 coal price was very close at \$2.01/GJ [60]).

The three points on the low end are COXY-9, COXY-15 and COXY-16. However, out of these, only COXY-9 appears to be an outlier in terms of both low non-fuel costs and low assumed fuel prices. COXY-12 and COXY-13 have average assumptions for fuel price, but have unusually high non-fuel costs. This is because COXY-12 and COXY-13 use additional purification steps in order to achieve much higher than normal CO_2 purities (about 99%) than in most CO_2 capture applications (specifically double-flash purification for COXY-12 and direct distillation for COXY-13). Therefore, they are outliers because their design objectives are significantly more stringent than the other selected studies, and so it is not fair to compare them with the other studies directly. CMEM-6 through CMEM-9 are outliers in both non-fuel (disproportionally low) and fuel (disproportionally high), but they were not removed from the analysis because their key metrics, such as LCOE and CCA, were comparable with other coal-based membrane studies.

In general, however, there is surprisingly little correlation between calculated non-fuel costs and classification of power plant. The membrane, oxyfuels, IGCC, SCPC, calcium looping, and chemical looping combustion classes span most of the entire range of the data set, with oxyfuels having the widest range. The SOFC non-fuel costs are consistently on the low-end as predicted by reasonable diversity of authorship (four different groups). However, note that SOFCs still remain unproven for CCS systems or at the 500 MW scale without CCS.

Natural gas outliers could not be identified, since the number of data points was low.
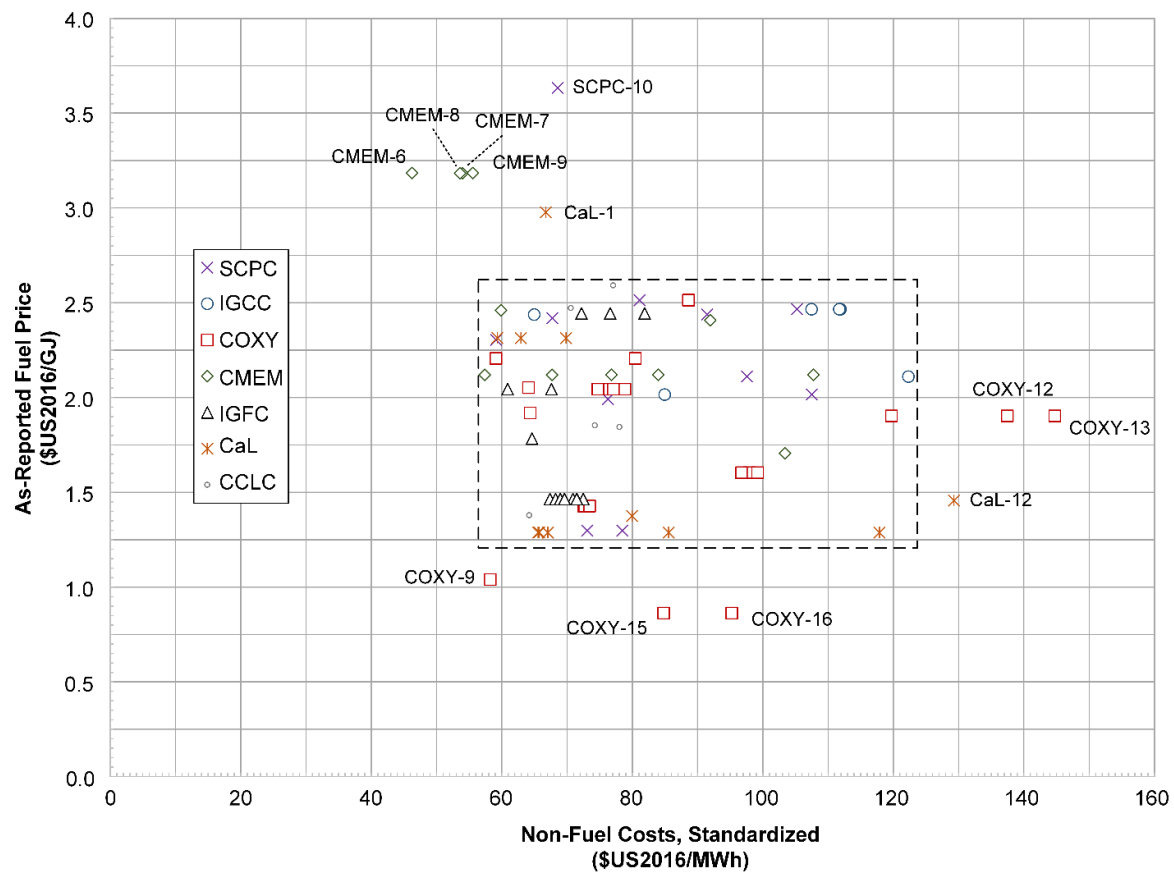
Figure 11. Scatter plot of fuel unit prices and non-fuel costs for coal-based systems as reported by each study and then converted to $US2016 according to the methods of Section 2. COXY = coal-based oxyfuels, CMEM = coal-based membranes, IGFC = integrated gasification (solid oxide) fuel cell, CaL = calcium looping capture, and CCLC = coal-based chemical looping combustion. The dashed line identifies the "typical" range of the data set as a visual aid only. Note that all plants in the above plot are CCS-enabled.

3.3. Trends in Key Metrics

Figure 12 shows the standardized LCOE of each plant using the as-reported fuel prices versus the life cycle CO_2 emissions (from the cradle to the plant exit). Some very clear trends emerge. First, the NG-based SOFC systems are clearly the best in terms of both LCOE and GHG emissions, with coal-based SOFCs coming in at a close second. Both classifications are also clearly outside the ranges of competing strategies such as NGCC and oxyfuels of either type, indicating a distinct difference due to the difference in technology paradigm. There is also sufficient agreement among the different independent research groups studying SOFC systems, such that, even if the SOFC studies that were published by members of our own research group were removed (for the sake of eliminating bias), the results remain essentially unchanged. The NGCC systems with CO_2 capture have a wider range in LCOE (with the most expensive being almost double the least expensive), but are very close in agreement on lifecycle GHG emissions. Generally speaking, the NGCC systems with CCS are preferable to oxyfuel technology.
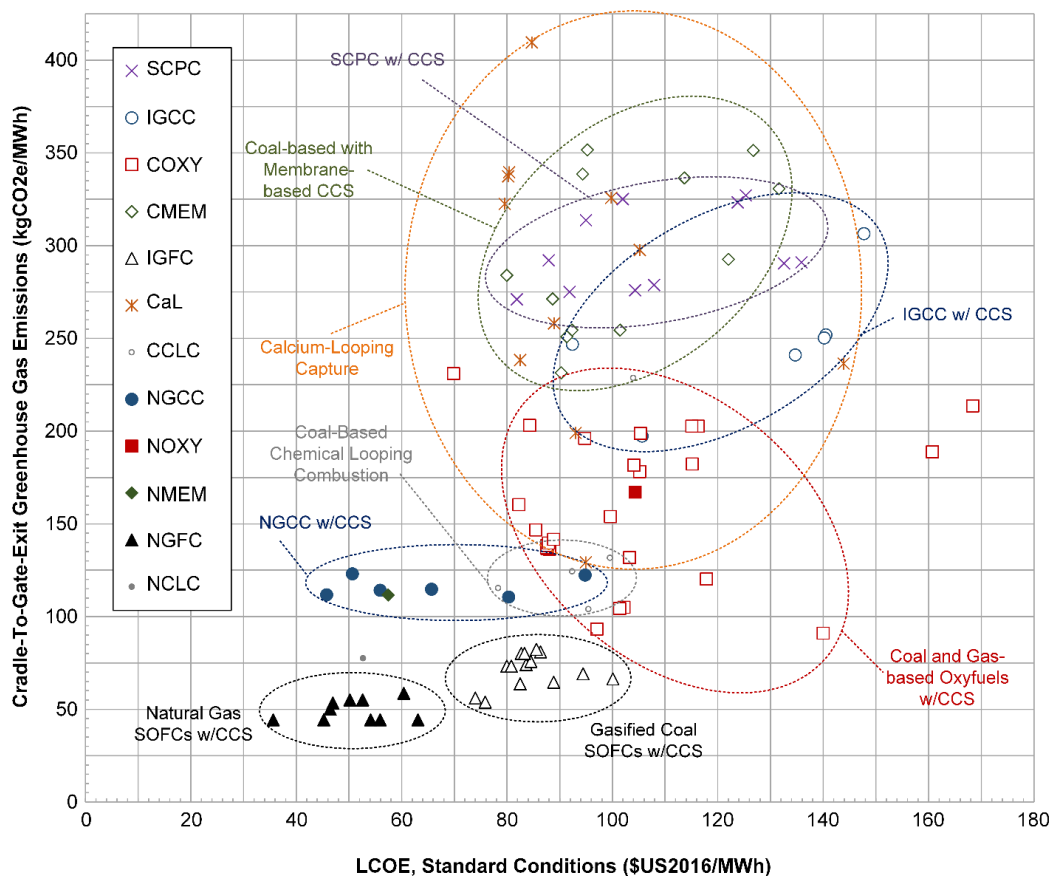
Figure 12. Scatter plot of the calculated LCOE at standard conditions of each plant vs. the calculated life cycle greenhouse gas emissions, using as-reported fuel price assumptions. NOXY = natural gas-based oxyfuels, NMEM = natural gas based membranes, NGFC = natural gas (solid oxide) fuel cells, and NCLC = natural gas-based chemical looping combustion.

The oxyfuel system and calcium-looping capture studies show the widest variation in GHG emissions and LCOE, but in general, oxyfuels have lower GHG emissions than the IGCC, SCPC, or coal-based membrane systems for approximately the same cost. The fuel type does not matter much, although the data on gas-based oxyfuels are few. Our ellipse drawn for the general oxyfuels region does not include the outliers COXY-9 (with very low LCOE) and COXY-12 and COXY-13 (with very high LCOE), for reasons explained in the previous subsection. The variability in the coal-based calcium-looping capture results is so large such that few conclusions can be drawn, except that it is generally comparable to solvent-based post-combustion capture, membranes, and IGCC. There is little evidence that calcium looping capture would outperform oxyfuels, chemical looping combustion, or SOFCs in terms of GHG emissions and or LCOE. Note that CaL-2 does not appear in Figure 12, because the efficiency of that system is so low as to be practically infeasible (resulting in very high LCOE and GHG emissions off the scale). CaL-8 has very low GHG emissions and LCOE compared to the other calcium-looping studies. This system uses IGCC as the primary power generation mechanism, is the only calcium looping study to predict greater than 99% CO_2 capture (the rest being in the 85–92% range), and predicts non-fuel costs lower than almost every other IGCC plant in the meta-study. This may be an outlier among the calcium looping studies. In addition, our methodology did not consider the GHG impacts of limestone production and transport, which could contribute to overly optimistic lifecycle emissions if the purge rate of sorbent is high.

The IGCC and SCPC systems are similar in LCOE and GHG emissions, except that IGCC has a tendency toward slightly higher LCOE and slightly lower GHG emissions. Since their overlap

is significant, the distinction between the two is not strong. The coal-based membrane systems, however, also overlap both IGCC and SCPC systems significantly. This is explained in part because, although the CO_2 capture portion of the system is different from their IGCC and SCPC counterparts, the upstream portions of the power plants using membranes for CO_2 capture are virtually the same (using either post-combustion capture with supercritical combustion or pre-combustion capture with gasification). This means that all of the potential performance improvements are limited to the CO_2 capture section of the plant. There is also no clear difference between pre-combustion and post-combustion membrane-based systems in terms of LCOE or GHG trade-offs. It is also not clear from Figure 12 whether there are any benefits to using membrane systems at all compared to IGCC or SCPC technologies. In addition, within the SCPC category, there is not enough information to suggest that a particular type of solvent (such as amines or chilled ammonia) is preferable to another in the general case, since all data points remained in a relatively small region.

With the exception of one study, the coal-based chemical looping combustion studies were tightly compacted in a small range (and performed by three different research groups). Although more studies are needed, based on the present results, the chemical looping combustion processes appear to be generally preferable to all of the other coal-based systems except for SOFCs in terms of these two metrics. Similar conclusions can be drawn concerning the lone natural gas-based chemical looping combustion point, although there is not enough data available in the literature to make this generalization against other natural gas systems.

NMEM-1 is the only natural-gas-based membrane study, and it lies inside the NGCC w/CCS ellipse. Although with only one data point there is little to conclude about gas-based membrane systems in general, this point does not hint at any major benefits over solvent-based systems when using natural gas.

Figure 13 shows the same analysis, except that the LCOE is computed using the same fuel prices for coal and for gas (the average US 2016 price for each fuel). COXY-9 now appears inside the oxyfuels ellipse (since it had assumed very low fuel prices originally). However, there is very little change in the general results, with IGCC, SCPC, and coal-based membranes still clouded together. Oxyfuels, NGCC, IGFC, and NGFC remain clearly clustered in distinct groups in the same positions.

Figure 14 shows a plot of the CCA (using as-reported fuel prices) compared to both the SCPC and NGCC without capture baselines of Table 1. It is clear that for a given baseline of comparison, all CO_2 capture-enabled systems are strongly correlated along a common, not-quite linear curve. Although CCA is a direct function of both LCOE and lifecycle GHG emissions, variations in LCOE explain most of the influence on CCA. Deviations from the curve are small, and due to either nuances in design differences or noise associated with differences in assumptions, estimates, and methodologies from study-to-study (likely a combination of both).

Negative CCA numbers for gas-based systems compared to coal are expected because the current price of natural gas in North America is relatively low. For example, [105] reported CCAs against SCPC without CCS of only 9.1 and 5.2 $US2007/tCO_2e for NGFC-7 and NGFC-10, respectively. However, in 2007, the relative price of gas per GJ was about 4 times higher than the price of coal per GJ, whereas in 2016 the ratio was only about 1.3 [60], and so it is expected that using 2016 prices could result in negative CCAs. However, some IGFC systems have negative CCA compared to SCPC without CCS, and similarly, some NGFC have a negative CCA compared to NGCC without CCS. This is at least partly explained by the significantly higher efficiencies of SOFC-based systems compared to their combustion counterparts, which have the dual effects of having significantly lower fuel costs and lower lifecycle CO_2 emissions per MWh of electricity produced, since higher efficiencies reduces emissions throughout the entire supply chain. Some contribution to negative CCAs could be that the power plants chosen for the baseline were simply more expensive than "normal".
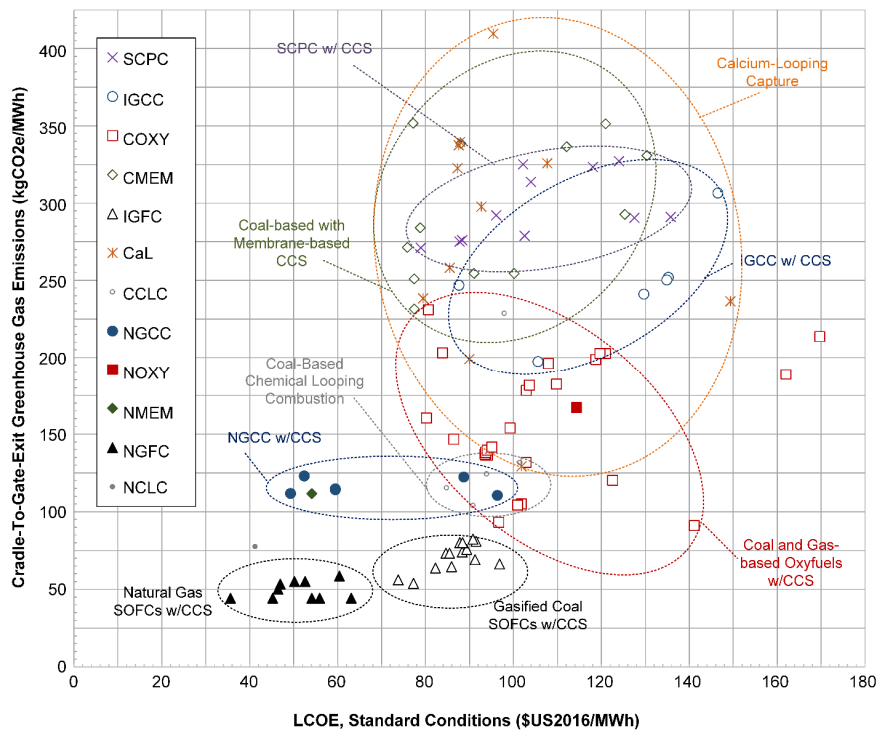
Figure 13. Scatter plot of the calculated LCOE at standard conditions of each plant vs. the calculated life cycle greenhouse gas emissions, but using average fuel prices for 2016 instead of as-reported prices. Ellipses are drawn as a visual aid only and do not specifically identify outliers.
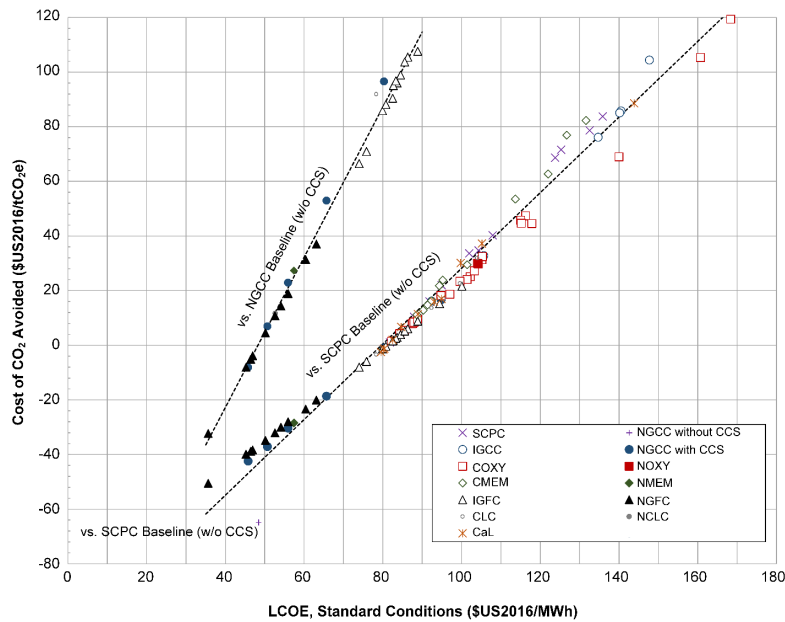

Figure 14. Scatter plot of the calculated LCOE at standard conditions of each plant vs. the calculated life CCA using as-reported prices, when compared against SCPC and NGCC without CO_2 capture. The NGCC baseline plant without CO_2 capture is also included as a point of reference. COXY-9 has been removed as an outlier. The straight dashed lines are regression lines for each reference group for the points included. Points with CCA above 120 \$US2016/tCO_2e were not included in this plot for clarity. Note that all data plotted above are for CCS-enabled plants, except for a single NGCC without CCS plant.

However, the assumptions and calculations made concerning the capital costs of the plants themselves play a larger role. For example, capital costs for SOFC systems include some degree of speculation of the SOFC stack prices. Adams and Barton [104] varied stack and operating costs as a parameter for their NGFC studies, which are reflected in Figure 14. NGFC-2 assumes a very optimistic SOFC stack cost of $500/kW installed, resulting in an extremely low CCA vs. NGCC of about −$32/tCO$_2$e. NGFC-5 assumes that the capital costs for the entire plant at 50% higher than their base case assumptions, resulting in a much higher CCA of $37/tCO$_2$e, a very significant difference.

The calcium looping studies have very low computed CCAs in Figure 14 (on par with the IGFC systems), with some even being negative compared to SCPC without CO$_2$ capture. This is overly optimistic, because calcium looping is effectively a parasitic load added on to an SCPC plant (or IGCC), and so it should not somehow cost less than a plant without capture. This is mostly explained by the very low fuel prices assumed in most of those studies. As shown in Figure 15, when the standard fuel price is used instead, the average CCA for the calcium-looping studies included in the meta-study is about $41/tonne, which is much more reasonable.

The average values of each of the major metrics by process types is shown in Figure 15 for the design classifications in which sufficient data are available. One key takeaway point is that in the greater scheme of things, IGCC is generally the worst choice power plant for the purposes of CO$_2$ emissions reduction, when considered as a replacement for SCPC without CCS. Coal-based membranes and coal-based oxyfuels are approximately equivalent as an SCPC replacement, since one tends to be better than the other depending on fuel prices. Gas significantly outperforms coal in terms of efficiency, cost, and environmental benefits, except for IGFC which has lower lifecycle emissions and slightly higher efficiency than NGCC on average.
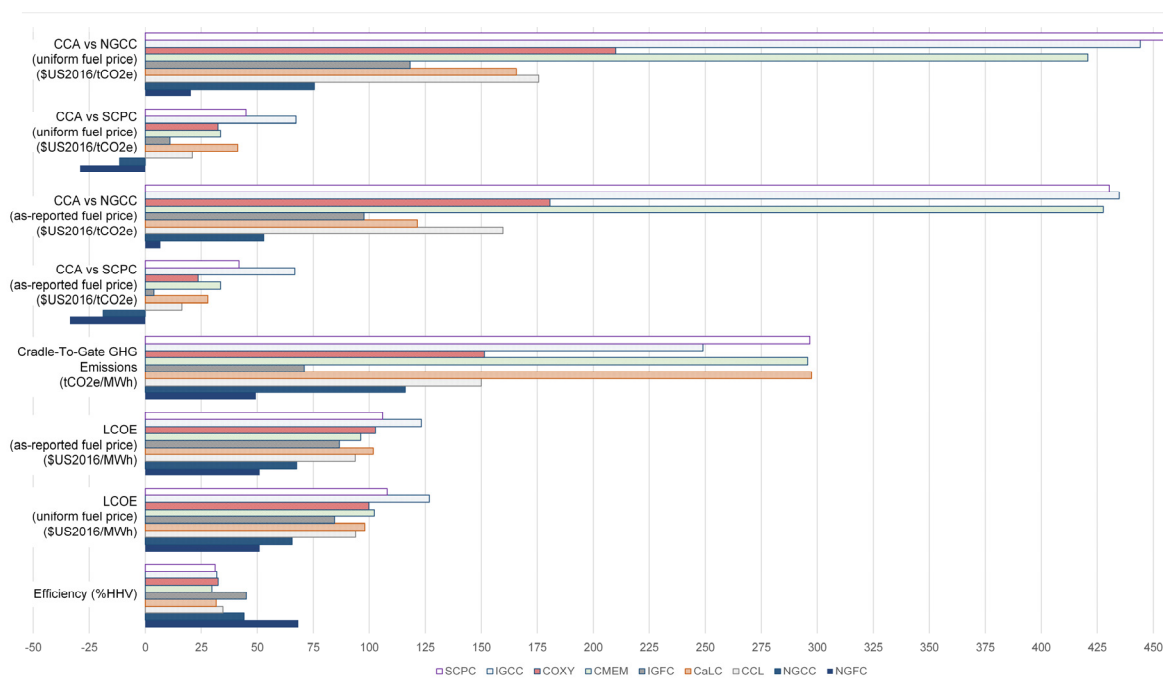

Figure 15. The average values of the key metrics by process category. COXY-9, COXY-12, and COXY-13 were not included (outliers as described in Section 3.2). Note that all plants in the above plot are CCS-enabled.

The SOFC systems, on the whole, had significantly higher efficiencies than the other systems, which is expected. Although capital costs of these systems are generally high, the non-fuel costs on a per MWh basis are in the mid-range because of the high efficiency. Similarly, fuel costs are the lowest,

and because nearly 100% of direct CO_2 emissions are eliminated, the life cycle CO_2 emissions are also the lowest as well. As a result, it has the lowest CCA numbers. Although the SOFCs themselves exist and are commercially available at 250 kW scales [38], the cited studies assume lower SOFC stack costs than are currently available (either by way of expected reductions in costs from manufacturing improvements, or by assuming longer lifetimes than are currently available) and no such plant has been constructed at the scales used in this study. In addition, existing 1st generation SOFC systems are not designed for CO_2 capture, because they do not use high-purity O_2 to catalytically oxidize the unspent portions of the fuel [38]. SOFC systems with CO_2 capture (2nd generation) have not been demonstrated at any scale in their entirety.

3.4. Sensitivity Analysis

As mentioned in Section 2.2, we used an assumed power law factor of $p = 0.9$ to estimate how the as-reported non-fuel costs would change when resized to a common 550 MW net power output scale. A sensitivity analysis was conducted in order to determine the impact of p on these six key metrics shown in Figure 15: average LCOE, average CCA vs. SCPC without CO_2 capture, and average CCA vs. NGCC without CO_2 capture, each using both as-reported and uniform fuel prices. In nearly every case, the relative order of each technology class remained unchanged. The only case where the value of p affected the relative order of the technology classes is shown in Figure 16, which is for CCA vs. SCPC using as-reported fuel prices. In this case, the coal-based oxyfuels and coal-based membrane classes had very similar CCAs at any p, with COXY decreasing a small amount with increasing p and CMEM increasing slightly with p. However, as noted earlier, there was already significant enough uncertainty and noise in the data such that there is no change in the final assessment. As such, the original value of $p = 0.9$ assumed in this analysis is perfectly suitable.
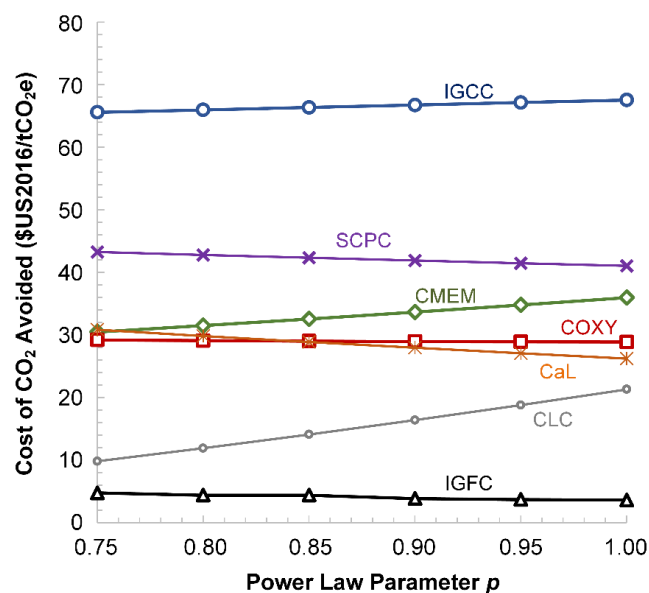
Figure 16. The average values of the CCA vs. SCPC without CO_2 capture for as-reported fuel prices as a function of the assumed power law parameter p.

4. Conclusions and Discussion

The meta-study results illustrate that there are *macro* differences and *micro* differences between technology options. The macro differences are associated with the broad design class of the power plant. For example, in Figure 12, it is clear that IGFC, NGFC, and oxyfuels stand distinctly apart from each other in the most relevant characteristics such as LCOE and lifecycle GHG emissions. SCPC,

IGCC, and coal-based membranes, however, do not differ significantly from each other at the macro level, but together are distinct from IGFC, NGFC, and oxyfuels.

These macro differences are explained by the fundamental design paradigm in how energy is converted from the chemical potential in the fuel to electric potential energy. SCPC uses classic combustion in air, which is undesirable from a CO_2 capture standpoint because the CO_2/N_2 post-combustion separation problem is an especially difficult one. The purpose of IGCC is to convert the CO_2/N_2 separation problem into a combination of O_2/N_2 and H_2/CO_2 separation problems through air separation, gasification, water gas shift, and pre-combustion CO_2 capture. However, the literature survey results showed that, on average, this is actually worse. Membrane-based solutions attempt to improve upon this by lowering the costs of the CO_2/N_2 or H_2/CO_2 separations, but at the macro level, there is no significant improvement gained in terms of CCA since, on average, global lifecycle emissions are actually worse than IGCC or SCPC. In other words, membranes can at best improve upon the CO_2 capture step itself, but cannot make macro-level improvements because it is still ultimately rooted in a fundamentally inefficient process paradigm that relies on either classic combustion or hydrogen production.

Instead, to make significant headway, completely different design paradigms are needed. The oxyfuel process converts the CO_2/N_2 problem into a combination of O_2/N_2 and H_2O/CO_2 separation problems. This requires neither solvent-based CO_2 capture nor membranes. Although the meta-study of the oxyfuels literature indicated a very large variance in key metrics, at the macro level, it is still distinctly superior to IGCC, SCPC, or coal-based membranes. The data on gas-based oxyfuels were limited, but what was available does not indicate a significant difference between coal and gas based approaches. SOFC systems also convert the CO_2/N_2 problem into a combination of O_2/N_2 and H_2O/CO_2 separation problems, but the SOFC stack itself does the job of both the O_2/N_2 separation and the electrochemical conversion, and at efficiencies fundamentally higher than oxyfuels because it avoids the limitations of the Carnot cycle. As such, the meta-study results indicated that it is the best approach at the macro level of any process class studied.

The results of the meta-study cannot be used to make determinations about which design choices within a technology class are the best at the micro-level (for example, we cannot make conclusions about which type of solvent is best to use for SCPC). This is because the differences in assumptions and methodologies from study to study are significant enough such that it is not possible to differentiate the effects of micro-level design choices from the noise. However, it is clear from Figure 12 that incremental improvements made at the micro-level (such as improved membranes or solvents) will not significantly improve key metrics such as CCA. Instead, significant improvements in CCA can only be achieved by abandoning current efforts such as SCPC, IGCC, or coal-based membranes and switching to newer design paradigms such as oxyfuels and SOFCs.

Finally, it is clearly evident that in the North American energy market, there is very little reason to use coal at all, since gas prices are so low. In other markets where gas is much more expensive (particular Asian ones), coal systems may make a lot more sense (particularly IGFC) but this study did not examine those markets. Most NGCC with CCS studies had a lower CCA than IGFC, although it has much higher lifecycle GHG emissions. However, out of the options considered, the cheapest way to avoid CO_2 emissions as a first step is to simply replace each existing SCPC plant as it reaches the end of its lifetime with an NGCC plant without CO_2 capture.

Consider a future United States energy mix in which coal has been entirely phased out and replaced with NGCC without carbon capture. At that point, as those NGCC plants age, in order to reduce emissions further, what technologies should replace them? Based on Figure 14, if energy prices remain relatively the same (in $2016), the better option could likely be to use NGCC with CCS, or even better, NGFC if it is available at scale at that time. Even IGFC is still not a great solution for North America (with CCA vs. NGCC above 100 $US2016/tCO$_2$e), but it could be a very important solution for other markets.

Looking at the big picture, the meta-study results suggest that it may not be prudent to invest large amounts of money into improving solvent-based and membrane-based CO_2 capture technologies at the micro-level. Membranes, in particular, make the least sense, because they are still relatively immature, and the apparent returns on membrane investment will be very small in the greater scheme of things. Rather, there is a major incentive to invest in oxyfuel technology to some degree and SOFC technology to a large degree, where the meta-study results show that the potential gains at the macro level are very significant. SOFCs in particular are fuel-flexible, and so are likely to be the best technological choice, with the corresponding fuel chosen based on the market prices of the day. Fortunately, in the near term, there is a clear path forward to making immediate reductions in CO_2 emissions at low cost through existing NGCC technology.

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Nomenclature

Abbreviations

CaL	Calcium-looping carbon capture
CCA	Cost of CO_2 emissions avoided
CCLC	Coal-based chemical looping combustion
CCS	Carbon capture and sequestration
CO_2e	Carbon dioxide equivalents
CMEM	Coal-based membrane separations
COXY	Coal-based oxyfuel combustion
GHG	Greenhouse gases
GWP	Global warming potential
HHV	Higher heating value
IGCC	Integrated gasification combined cycle
IGFC	Integrated gasification (solid oxide) fuel cell
LCOE	Levelized cost of electricity
MEM	Membrane-based separation
NCLC	Natural gas-based chemical looping combustion
NGCC	Natural gas combined cycle
NGFC	Natural gas (solid oxide) fuel cell
NMEM	Natural gas-based membrane separations
NOXY	Natural gas-based oxyfuel combustion
OXY	Oxyfuel combustion
PPPI	Purchasing power parity index
SCPC	Supercritical pulverized coal
SOFC	Solid oxide fuel cell
t	Metric tonne (1000 kg)
WGS	Water gas shift

Variables

$\in$	electrical efficiency on a HHV basis (the net electrical power out at the plant exit divided by the total HHV of the fuel brought into the plant).
BPR	relative price ratio of fuel (the average price of coal or gas in the US divided by the price of the same amount of fuel in the country of the study in the \$US for the year of that study)
C	Cost
E	The exchange rate (in \$US per local currency units)
ECI	energy cost index
GWP	Global warming potential (in tCO_2e per MWh of net power output)
LCOE	levelized cost of electricity
NPO	net power output (in MW)
PPPI	Purchasing power parity index
U	unit price of fuel (on a per MJ of fuel on HHV basis).
p	the power law exponent for the "six-tenth's rule" used to scale non-fuel costs.

Subscripts

country	applies to the location in which a plant is built
fuel	applies to the fuel expenses associated with power plant operation (either coal or gas).
nonfuel	applies to the non-fuel portion of a plant (e.g.,: capital costs, non-fuel operating costs, business expenses, etc.).
original	applies to the plant if built at the size reported in the cited study.
originalyear	the original year of the study
scaled	applies to the plant if built at the standard size listed in Table 1.
standard	applies to a plant relocated to the standard location and standard project year as shown in Table 1.

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