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Technical Documentation: Oxygen-based Combustion Systems (Oxyfuels) with Carbon Capture and Storage (CCS)

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DEVELOPMENT AND APPLICATION OF OPTIMAL DESIGN CAPABILITY FOR COAL GASIFICATION SYSTEMS

**Technical Documentation:
Oxygen-based Combustion Systems (Oxyfuels) with
Carbon Capture and Storage (CCS)**

Final Report of

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Objective

The basic objective of this research is to develop a model to simulate the performance and cost of oxyfuel combustion systems to capture CO₂ at fossil-fuel based power plants. The research also aims at identifying the key parameters that define the performance and costs of these systems, and to characterize the uncertainties and variability associated with key parameters. The final objective is to integrate the oxyfuel model into the existing IECM-CS modeling framework so as to have an analytical tool to compare various carbon management options on a consistent basis [1].

Literature Review

Process Overview

Oxyfuel combustion for CO₂ capture was first proposed in 1981 by researchers at the Argonne National Laboratory. The basic approach is to use pure oxygen for combustion, rather than air, so as to produce a flue gas stream consisting mainly of CO₂ and water vapor. The water is then easily removed, leaving a concentrated CO₂ stream for disposal. To prevent excessively high temperatures in the boiler, a portion of the flue gas stream is recycled back to the boiler to dilute the oxygen and maintain temperatures similar to conventional air-blown designs.

A review of recent studies reveals that different organizations employ substantially different design assumptions regarding the plant configuration. Table 1 summarizes the configuration options defining the scope of an oxyfuel plant model assumed by various studies [2-15]. These studies also use different assumptions for various process parameters, as indicated in Table 2.

Table 1. Summary of oxyfuel plant configurations assumed by various studies (data not available is represented by a blank entry in the table)

| Study/Reference | Year | Plant type & size (MW _g) ¹ | Flue gas recycle ² | Particle removal ³ | FGD ⁴ | SCR ⁵ | Flue gas cooler ⁶ | Dry CO ₂ ⁷ | Other units ⁸ |
|----------------------------|------|---|-------------------------------|-------------------------------|------------------|------------------|------------------------------|----------------------------------|--------------------------|
| Dillion et al. [2] | 2004 | New, 740 | Dry | ESP | No | No | Yes | Distill | No |
| AAL [3] | 2004 | New, 533 | Wet | ESP, out | Yes* | No | No | No | ACI |
| AAL [4] | 2003 | Retrofit, multiple♦ | Wet | ESP, out | Yes* | Optional* | No | No | ACI |
| AAL [4] | 2003 | New, multiple | No | ESP | Yes | No | No | No | ACI |
| ANL [4] | 2003 | Retrofit | Wet | ESP | Yes | No | No | No | APH, O ₂ htr |
| U Waterloo [6] | 2003 | Retrofit, 400 | Wet | | | | Yes | Distill | Aux power |
| Chalmers/Vattenfall [7, 8] | 2002 | New, 933 | Wet | Cyclone | No | | No | Distill | No |
| ALSTOM/ABB/AEP [9] | 2001 | Retrofit, 463 | Wet | ESP | Yes | No | Yes | Distill | No |
| AP/BP/Babcock [10] | 2000 | New | Wet | | No | No | Yes* | Distill | No |
| Simbeck [11] | 2000 | New, 575 | Dry | Baghouse | No | No | No | No | No |
| Simbeck [11] | 2000 | Retrofit, 318 | Wet | ESP | No | No | No | No | Aux power |
| McDonald & Palkes [12] | 1999 | Retrofit, 318 | Wet | ESP | No | No | Yes | Distill | APH, O ₂ htr |
| Babcock et al. [13] | 1995 | 660 | Dry | ESP | No | No | Yes | Distill | Claus |
| Air Products [14] | 1992 | Retrofit, 572 | Wet | | | | | Distill | No |
| Japanese [15] | 1992 | New, 1000 | Wet, Dry | ESP | No | No | Yes* | No | No |

¹ Gross plant size (MW)

² Recycled flue gas may be wet (retaining the moisture) or dry (dried and then recycled)

³ Removal of particulate matter in the flue gas can be achieved using an Electro-static precipitator (ESP) or a Cyclone or a Bag house

⁴ Flue gas desulfurization system for SO₂ control

⁵ Selective catalytic reactor for NO_x control

⁶ Flue gas cooling is generally required and is achieved using a direct contact cooler

⁷ In order to achieve high purity CO₂ product, distillation (Distill) is commonly used to remove inerts

⁸ Some of the studies mention other additional units such as: Activated carbon injection (ACI) system for mercury control, Air pre-heater (APH) and oxygen heater for better heat integration, Auxiliary power generation (Aux power), and Claus plant (Claus) to recover sulfur from SO₂ stream

*All these units are located outside the recycle loop, else they are located inside the recycle loop by default

♦ Multiple plant sizes: 500, 200, 100 and 30 MW_g

Table 2. Summary of the key process parameter values assumed by various studies (data not available is represented by a blank entry in the table)

| Study/Reference | Year | FGR ratio ¹ | Oxygen purity (mole %) ² | Excess air (%) ³ | Air leakage (%) ⁴ | FGR temp (°C) ⁵ | CO ₂ purity (mole %) ⁶ | η_{CO_2} ⁷ (%) | % change ⁸ in η_{boiler} | % change ⁹ in NO _x |
|----------------------------|------|------------------------|-------------------------------------|-----------------------------|------------------------------|----------------------------|--|---------------------------------------|---|--|
| Dillion et al. [2] | 2004 | ~0.67, w/w | 95 | 19 | | 35, 330 | 96 | ~91 | | |
| AAL [3] | 2004 | 0.75 to 0.8 | 99 | | ~5 | | 95 | ~100 | | -70 |
| AAL [4] | 2003 | | | | | | | | | ? |
| AAL [4] | 2003 | 0.0 | | | | | | | | -53 to -76 |
| ANL [4] | 2003 | | | | | | | | | |
| U Waterloo [6] | 2003 | 0.71 | 99.5 | 3 | 1 | 40 | 95 | 90 | | |
| Chalmers/Vattenfall [7, 8] | 2002 | 0.64, w/w | 95 | 1.5 | | 340 | 98, w/w | ~100 | | -67 |
| ALSTOM/ABB/AEP [9] | 2001 | ~0.67 | 95, 99 | 15 | 1 | 38 | 98 | 94 | +2.7 | |
| AP/BP/Babcock [10] | 2000 | ~0.67 | 95 | | | 31 | ~97 | 93 | | |
| Simbeck [11] | 2000 | 0.71, w/w | 95 | 0 | | | | 100 | 0 | |
| Simbeck [11] | 2000 | 0.73, w/w | 99.5 | 6.5 | 1 | | 85 | | +3.5 | -31 |
| McDonald & Palkes [12] | 1999 | ~0.67 | 99 | | 1 | 38 | 98 | | +3.5 | |
| Babcock et al. [13] | 1995 | 0.65, 0.75, 0.85 | 95, 99.5 | 10, 15, 17 | 0, 1, 3, 5 | 45 | 85 to 99 | 95 to 100 | +4 to +6 | -60 |
| Air Products [14] | 1992 | | 99.5 | 2 | | | | 98 | | |
| Japanese [15] | 1992 | | 97.5 | | | | 95+ | 90 | +3 | |

History

The fact that oxygen is required for sustaining a combustion reaction has been known for centuries. The name “oxygen” comes from the Greek words “*oxy genes*” meaning “acid former.” Although oxygen was prepared by several researchers by the late 18th century, it was not recognized as an element until identified by Joseph Priestley, an English chemist, who is generally credited with the discovery of oxygen in 1774. Swedish researcher Carl Wilhelm Scheele had independently discovered oxygen and studied its properties during 1771-3, but his work was published later in 1777. Oxygen was liberated by intensely heating mercury oxide, which is a common laboratory procedure to produce oxygen even today [16-18].

¹ Flue gas recycle (FGR) ratio is the fraction of the total flue gas being recycled to the boiler

² Purity of the oxygen used in the oxyfuel combustion process

³ Excess air is the fraction of theoretical air (or oxidant), and is used to ensure complete combustion

⁴ Undesired air infiltration into the boiler, expressed as a fraction of theoretical air

⁵ The temperature at which the recycled flue gas stream is introduced back into the boiler

⁶ The percentage of CO₂ present in the product stream

⁷ The overall CO₂ capture or removal efficiency of the system

⁸ Boiler efficiency is the fraction of energy in combustion that is converted to steam energy; this column represents the relative efficiency of oxygen-firing to air-firing

⁹ NO_x emission rate (lb/MBtu) for oxygen-firing, relative to air-firing

It took more than a hundred years after the discovery of oxygen for a large-scale production of oxygen. Air, which is an abundant source of oxygen, could not be used to produce pure oxygen until the end of 19th century. Carl von Linde obtained a patent for the world's first modern refrigerator in 1877, an essential component of modern cryogenic systems. He was among the first in the world to produce large volumes of liquid air (1895), and in 1902 began constructing his first air separation unit [19, 20]. Oxygen production plants using air separation technology became commercially available in another decade or two, and many more companies entered this field. Air Products, Inc. built its first oxygen generator in the 1940s and is now one of the leading manufacturers of oxygen plants [21].

Oxygen was being produced for various industrial uses and also for use in the health care sector. High temperature flames using oxyfuel combustion (e.g., using acetylene) became popular in welding and other metal processing applications. Large amounts of oxygen are also consumed in various petrochemical industries to produce a large array of chemicals and polymers.

The idea of using oxyfuel combustion for CO₂ capture in a coal-fired furnace is much more recent. It was first proposed by Horne and Steinberg in 1981, and was also being studied by Wolsky and others at Argonne National Laboratory at that time [14]. There was a growing interest in capturing CO₂ during the 1970s, not because of greenhouse gas concerns, but due to its potential use in enhanced oil recovery. In 1982, Abraham et al claimed that oxyfuel combustion was 20% cheaper than an MEA process for CO₂ capture [22]. Later, as the oil crisis of the late 1970s subsided, and real oil prices fell, the interest in capturing CO₂ also diminished. Some experimental work continued at ANL and a few other places through the 1980s and early 1990s [23]. Many more research groups started looking into this technology in late 1990s, when greenhouse gas control emerged as a global environmental issue.

Oxyfuel technology is now being promoted as a promising option for CO₂ capture from power plants. However, it is still in the early stages of development. Although various parts of this system (such as oxygen production and flue gas treatment) are commercially available today, only laboratory-scale studies of oxyfuel combustion for coal-fired power generation have been conducted so far, with some pilot plant studies also in progress. Recently, Vattenfall has announced a plan to build a 40MWt demonstration plant using oxyfuel combustion technology.

Advantages

A number of features make oxyfuel combustion technology a potentially attractive option for capturing CO₂ from power plants [24-26].

- **Use of steam cycle technology:** Oxyfuel combustion systems use conventional boiler technology, which the power plant community is familiar with. This also makes it a potential candidate for CO₂ retrofits to existing steam plants. As it does not use any major chemical processes (like gasification, water-gas shifting, etc.), it is perceived as a more reliable system. More importantly, independent of greenhouse gas concerns, there are on-going efforts to improve the steam cycle efficiency. Oxyfuel combustion systems would benefit from these developments as well.
- **Nitrogen-free combustion:** When air is used in conventional combustion, it introduces a large amount of nitrogen which is inert (does not help the combustion reaction). When pure oxygen is used in place of air, the quantity of flue gas generated reduces substantially.

This leads to reductions in equipment sizes and heat losses, and to savings in the cost of flue gas treatment.

- **Lower emissions:** Use of oxyfuel combustion technology with CO₂ capture opens up the possibility of a zero-emission (or close to zero emissions) coal power plant. Almost all the CO₂ from the plant can be captured using this process, whereas other CO₂ capture technologies become increasingly expensive as the CO₂ capture efficiency approaches 100%. Various experimental studies using O₂/CO₂ recycle show significant reduction in NO_x formation, as part of the NO_x in the recycle stream is believed to get dissociated to form nitrogen [27-30]. Thus, the NO_x levels in these boilers may fall significantly. Some studies have also reported substantial reduction in mercury emissions as well as enhanced SO₂ removal efficiency in FGD units [31, 32]. Finally, there is a possibility of co-capture of other pollutants (especially SO₂) along with CO₂, if co-disposal becomes feasible and acceptable.
- **Potential cost savings:** At present, oxyfuel configurations assume an externally recycled flue gas stream, required to control temperature in the boiler in order to avoid ash melting problems. However, better materials and boiler designs may help eliminate (or substantially cut down) the need for recycled flue gas. This would lead to very compact boilers and flue gas cleanup devices, which cost substantially less. Further cost savings are also expected from improved efficiency, elimination of certain flue gas cleanup devices (e.g. SCR) and improvements or new developments in oxygen production technology (e.g. use of ion transport membranes).

Issues and Challenges

Several key issues or challenges need to be addressed in order to make oxyfuel capture systems feasible and competitive [24-26].

- **Boiler design:** Today there is lack of fundamental knowledge in order to design a boiler using pure oxygen for combustion. For example, how much excess oxygen would be required? What kind of oxygen distribution system needs to be used to ensure complete combustion of the fuel into CO₂ and water (and avoiding CO formation)? There is a need for more experimental and modeling work, as well as for verification and validation of reliable heat transfer models. Use of pure oxygen for combustion leads to very high flame temperature. This may lead to problems such as ash melting and high-temperature NO_x formation. Also required are new materials that can be used to fabricate the high temperature boilers, especially if flue gas is not recycled. Another potential problem with the boiler design is air leakage. The main reason for using pure oxygen for combustion is to obtain a flue gas which is almost all CO₂ when dried. However, air leakage may lead to significant amount of nitrogen in the flue gas. Designing an air-sealed boiler is a challenge.
- **Large-scale oxygen production:** For a typical power plant, the oxygen requirement would be very large, several multiples of other current industrial applications. Current air separation technology (cryogenic) has a very large energy requirement and capital cost.

Determining the optimum level of oxygen purity is another challenge, and is dependent on the CO₂ product purity requirements.

- **Co-capture and co-disposal:** Although it is commonly assumed that CO₂ could be disposed of along with SO₂ and NO_x, it is not clear yet, if this would be technically feasible (e.g., because of potential problems in compression of this mixture, as well as corrosion issues in pipeline transport). Economic viability and environmental acceptability are other key factors. Depending upon the CO₂ purity requirement as dictated by regulation or the end user, further purification of the flue gas may be required.
- **Other environmental emissions:** An oxyfuel combustion system, especially if it uses near-stoichiometric or low excess oxygen, may have higher CO emissions, and may also leave some unburned carbon. Secondly, the condensate from this process has higher amounts of dissolved acidic gases and hence needs treatment. Also, it is possible that trace toxic substances might be introduced to storage sites (e.g. geologic formations) through co-capture and co-disposal.

Performance Model

A preliminary model has been developed to simulate the performance of oxyfuel combustion system for CO₂ capture. It is designed to yield mass and energy flows across the various units such as the ASU (oxygen generator), boiler, air preheater, oxygen preheater, flue gas recycle fan and other plant components. The model has been built and integrated with the existing IECM-CS modeling framework [1].

In order to determine the most suitable configuration, it is necessary to consider the following questions:

- Is the recycled flue gas dried (dry recycle) or is it recycled along with its moisture content (wet recycle)?
- Is it necessary to cool the flue gas prior to recycle?
- What type(s) of particulate control unit should be used (cyclone, baghouse or cold ESP)? If ESP is used, how would the performance get affected because of different flue gas composition (as compared to conventional system)?
- Is the recycled flue gas stream treated for particulate control?
- Should the flue gas be treated for SO_x and NO_x control?
- If the flue gas is treated for SO_x and NO_x control, where would these units be placed with respect to the recycle point?
- If the CO₂ product is disposed with SO_x and NO_x content, how might it affect the performance of the compression system and the cost of transport and storage?

We have attempted to address each of these questions based on a review of the literature. In defining the scope of an oxyfuel system model, we have also considered tradeoffs between the number of configuration options and the resulting data requirements and complexity added to the model.

Model Configurations

The following menu system configuration options are included in the current model:

- Plant type: New or Retrofit
- Steam cycle: Sub-critical or supercritical
- Oxygen generator: Cryogenic or ITM (advanced)

- Flue gas recycle: Wet recycle or dry recycle
- Particulate removal: ESP/cyclone/baghouse
- Flue gas cooler: Yes (within recycle loop) or No
- If FGR and FG cooler: Where should it be located?
- FGD - optional
- SCR - optional
- CO₂ purification system - optional
- Heat integration features: APH, O₂ heater, use of N₂

Default Configuration

The default model configuration for the oxyfuel combustion system for CO₂ capture in IECM-CS is as follows:

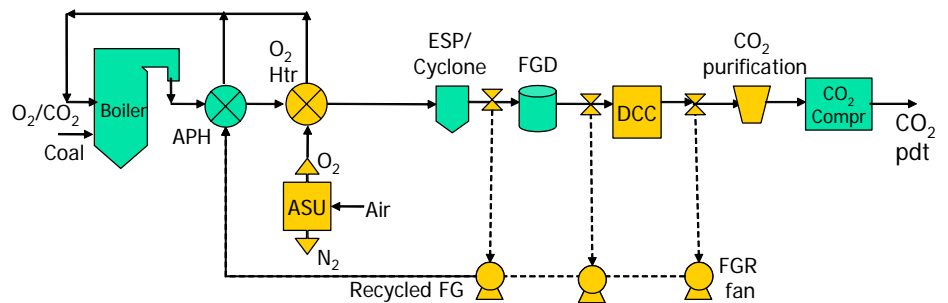


Figure 1. Oxyfuel combustion (O₂/CO₂) system configuration in IECM-CS model

Key Model Parameters

The key model parameters defining the performance of the oxyfuel combustion system for CO₂ capture are as follows:

- **Oxygen purity:** Air contains about 21% oxygen on molar basis. The oxygen product obtained from an air separation unit (ASU) is typically in excess of 90%. It may be noted that the energy penalty (and the cost of separation) increases sharply with higher product purity. However, at higher oxygen purity there are less non-condensable impurities in the CO₂ product obtained from the system. Many studies have reported that 95% is an optimal level of oxygen purity. This value has been used as a model default. Argon is the main impurity in the oxygen product, with some traces of nitrogen.
- **Oxygen pressure:** This is the pressure at which the oxygen product is delivered from the air separation unit. The total energy requirement for the ASU also depends on this pressure.
- **Excess oxygen:** Excess oxidant is generally provided to ensure complete combustion of the fuel and to avoid formation of carbon monoxide. Conventional coal combustion is carried out using about 15-20% excess air. Since pure oxygen is an expensive commodity as compared to air, it is necessary to minimize the use of excess oxygen.

The optimum level of excess oxygen needed to ensure complete combustion is not yet clear. Various studies assume values in the range 0-19%, the majority of them being on the lower side. Hence a default value of 5% is used.

- **Air leakage:** Ideally, the oxyfuel system aims at using only pure oxygen for combustion. However, it may not be practically feasible to seal the boiler and flue gas ductwork completely to avoid air ingress. Such air infiltration into the system is termed as air leakage. It is crucial to keep it at minimum level. Values in the range of 1-5% have been assumed by various studies, while many others tend to ignore this parameter and assume zero air leakage. In a conventional air-fired boiler, the amount of air leakage is typically 15-20% of the theoretical air requirement. It is expected that oxyfuel systems would be better sealed and the default value for air leakage is thus assumed to be 2% of theoretical (stoichiometric) oxygen.
- **Flue gas recycle ratio:** Oxyfuel combustion systems with flue gas recycle are also commonly referred to as “O₂/CO₂ combustion systems”. The flue gas recycle ratio (FGRR) is the fraction of total flue gas generated that is recycled back into the boiler. Higher FGRR implies a lower oxygen mole fraction in the O₂/CO₂ oxidant entering the boiler, whereas zero FGRR is the case of pure oxygen combustion with no flue gas recycle. Studies using flue gas recycle assume FGRR values in the range 0.6-0.85. The IECM-CS uses a nominal value of 0.7.
- **Recycled flue gas temperature:** The temperature of the recycled flue gas would decide the temperature of oxidant stream (after mixing with pure oxygen) entering the boiler, and hence affect the working of the air preheater and the boiler efficiency. It is recommended that the flue gas be cooled down to near ambient temperature (say 40 degC), especially in the retrofit applications, in order to make use of the existing air preheater. Not all the studies use flue gas cooler, and the FGR temperature is quite high in those configurations.
- **FGR fan pressure head:** A fan is used to provide a small pressure head for the recycled flue gas stream going back to the boiler. This FGR fan pressure head along with the recycled flue gas flow rate, determine the energy used by the fan. The nominal (default) value for this pressure head is 0.14 psi.
- **Flue gas moisture removal:** The recycled flue gas may be sent back to the boiler with or without moisture. The flue gas moisture removal level is the fraction of moisture removed from the recycled flue gas stream. It would be zero in case of a wet recycle system (the more prevalent assumption), and close to one in case of dry recycle. IECM-CS uses the value of zero (wet recycle) as default value.
- **CO₂ product purity:** The flue gas from oxyfuel combustion is a mixture of CO₂ with other compounds. Even after drying (i.e. removal of H₂O, which is the second largest component in the flue gas), the concentrated CO₂ stream may contain various non-condensable gases (e.g. N₂, O₂, Ar) and pollutants (SO₂, NO_x, HCl), depending on the combustion conditions and various parameters discussed before. Some studies assume that the CO₂ product may be compressed and disposed together with all these impurities (co-disposal), while other studies propose schemes for CO₂ product purification. The CO₂ product purity

is a parameter that would dictate the kind of post-treatment required for the CO₂ stream. It would also affect the energy requirement for CO₂ purification and compression. A nominal purity of 97.5% is assumed in the IECM-CS.

- **CO₂ capture efficiency:** Under ideal conditions, oxyfuel combustion system with flue gas recycle should be able to capture all the CO₂ present in the flue gas, i.e. the theoretical capture efficiency of this system is 100%, as assumed by some studies [3, 7, 8, 11]. However, CO₂ emissions do occur while operating this plant, especially during drying and purification of the concentrated CO₂ stream. Accounting for these undesired and unavoidable losses, the CO₂ capture efficiency of this system as reported by various studies is in the range of 90-98% [2, 6, 9, 10, 13-15].
- **CO₂ product pressure:** This is the final pressure at which the CO₂ product is delivered at the plant boundary. A typical value is about 2000 psig (13.7 MPa). This parameter, along with the CO₂ compression efficiency, determines the total energy requirement for CO₂ compression, which is a major energy penalty item second only to that of the air separation unit.
- **CO₂ compressor efficiency:** Based on our previous studies a nominal (default) value of 80% is assumed for the CO₂ compressor.

Table 3 summarizes the nominal parameter values for the oxyfuel model, along with the ranges employed in the IECM-CS.

Table 3. Key process parameter values in IECM-CS oxyfuel model

| Parameter | Units | Default value | Range |
|--|----------|-----------------|----------|
| Oxygen purity | %mole | 95 | 90-100 |
| Oxygen pressure | MPa | 0.1 | 0.1 |
| Excess oxygen | % theor. | 5 | 0-19 |
| Air leakage | % theor. | 2 | 0-5 |
| Flue gas recycle ratio | fraction | 0.7 | 0.6-0.85 |
| Flue gas recycle temperature | degC | 38 | 35-40 |
| FGR fan pressure head | psi | 0.14 | 0.14 |
| FG moisture removal | % | 0 (wet recycle) | 0-100 |
| CO ₂ product purity | % mole | 97.5 | 90-100 |
| CO ₂ product pressure | MPa | 13.8 | 7.6-15.2 |
| CO ₂ compression efficiency | % | 80 | 75-85 |

ASU Model

The oxyfuel system model nominally assumes a conventional cryogenic air separation unit. The ASU performance and cost models previously developed for the IGCC plant systems [33] also is used for the oxyfuel model.

Calculation Strategy

The IECM-CS is an integrated modeling framework that simulates the performance and cost of fossil-fuel power generation systems with environmental controls. All major plant components and multi-pollutant interactions are taken into consideration. The following sub-sections describe the algorithm used to estimate the performance of the oxyfuel combustion system. The algorithm is illustrated in the Case Study section later in this document.

Input Parameters

To begin, the following parameters are specified by the user (or the model defaults):

1. ASU product composition (as an elemental volume percent)
2. Coal composition (as an elemental weight percent, plus ash and water weight percent)
3. Excess oxygen to boiler (as a percent of the stoichiometric oxygen)
4. Air leakage to flue gas (as a percent of the stoichiometric oxygen)
5. Flue gas recycle ratio (as a percent of the total flue gas produced)
6. Gross size of plant (as megawatts of internal power produced)
7. Gross plant heat rate (as a combination of the steam cycle heat rate and the boiler efficiency)

Coal Flow Rate

Calculate the coal flow rate based on MW_g , heat rate and coal properties (heating value). The relationship in Equation (1) can be used to determine the coal flow rate required to generate the desired (or actual) gross power, given the coal properties and gross heat rate.

$$M_{coal} = \frac{MW_g \times HR_{steam}}{2 \times \eta_{boiler} \times HHV_{coal}} \quad (1)$$

where,

M_{coal} = mass flow rate of coal (ton/hr)

MW_g = gross cumulative power produced by the entire power plant; this does not consider power used by equipment in the power plant (MW)

HR_{steam} = heat rate of the steam cycle, which excludes the effects of the boiler efficiency (Btu/kWh)

η_{boiler} = boiler efficiency (fraction)

HHV_{Coal} = higher heating value of the coal on a wet basis (Btu/lb)

Oxygen Requirement

The oxygen flow rate required by the air separation unit is done through the following steps.

1. Calculate the stoichiometric O₂ requirement based on the coal flow rate, coal composition, and emission factors for incomplete combustion reactants
2. Calculate the total O₂ requirement based on the excess oxygen specified
3. Calculate the total oxygen product (i.e., oxidant) flow rate based on the oxygen purity and total O₂ requirement

Air Leakage

The air leakage stream is calculated based on air composition and air leakage input parameter.

Combustion Product

The combustion product is referred to as a flue gas stream. Given the coal and oxygen flow rates into the boiler, the composition and flow rate of the flue gas stream can be calculated.

Recycled Flue Gas

For the next iteration, part of the flue gas is recycled back to the boiler. The recycled flue gas is then added to the coal and oxygen streams described above. The flow rate of the recycled stream is calculated using FGRR and the total flue gas flow rate; this amount is then included in the estimation of the total flue gas combustion product.

The calculation is repeated until a steady state is achieved. Once the mass flow rates are balanced (it may take few iterations), the temperatures of various streams are estimated through heat balance over each unit (boiler, air preheater, O₂ heater). The temperature of the oxidant stream (mixture of recycled flue gas and pure oxygen) is estimated through simple energy balance over the air preheater (APH) unit.

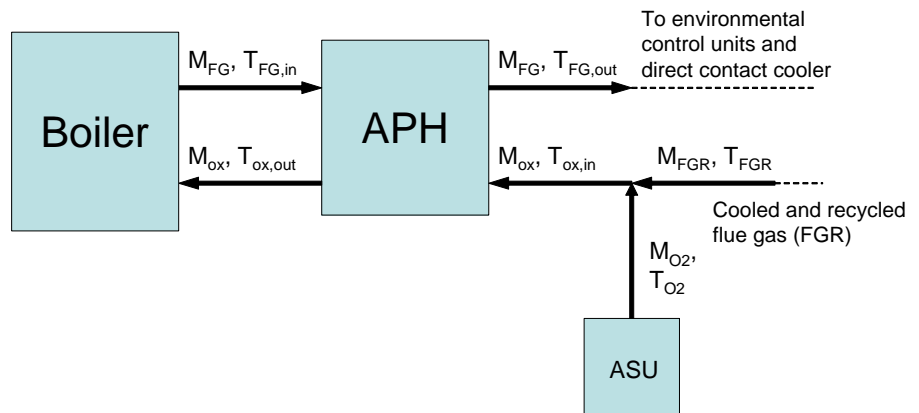


Figure 2. Energy balance over the air preheater unit

The energy balance equations yields:

$$T_{ox,in} = \frac{M_{O_2} \times c_{p,O_2} \times T_{O_2} + M_{FGR} \times c_{p,FGR} \times T_{FGR}}{M_{ox} \times c_{p,ox}} \quad (2)$$

$$T_{ox,out} = T_{ox,in} + \frac{M_{FG} \times c_{p,FG} \times (T_{FG,in} - T_{FG,out})}{M_{ox} \times c_{p,ox}} \quad (3)$$

where,

$c_{p,FG}$ = average specific heat of the flue gas (FG)

$c_{p,FGR}$ = average specific heat of the recycled flue gas (FGR)

c_{p,O_2} = average specific heat of pure oxygen (O₂)

$c_{p,ox}$ = average specific heat of the combined oxidant (ox)

CO₂ Product Stream

The CO₂ product composition and flow rate calculation is based on the CO₂ capture efficiency and CO₂ purity requirement.

Cooling Water

The cooling water requirement is based on the flue gas flow rate and the desired temperature difference. The reference case study reports a cooling water requirement of 93,200 gpm for a plant treating a flue gas flow rate of 809,763 ft³/min, the flue gas being cooled from 144°F to 100°F. So, the cooling water requirement is obtained as $3.3(10)^{-3}$ gpm per ft³/min per °F.

$$M_{cooling} = 3.3(10)^{-3} \times V_{fg} \times \Delta T \quad (4)$$

where,

$M_{cooling}$ = cooling water requirement (gpm)

V_{fg} = flue gas flow rate (actual ft³/min) at 100 °F

ΔT = desired temperature difference (°F).

Power Requirement

The energy requirements must be calculated for the flue gas recycle fan, the air separation unit, the CO₂ purification unit, and the CO₂ compression unit. The following expressions derived in other studies are used to estimate these power requirements [33, 34]:

ASU Unit Power

$$MACP = 0.0049\phi + 0.4238, \text{ for } \phi \leq 97.5\% \quad (5)$$

$$MACP = 0.0736 / (100 - \phi)^{1.3163} + 0.8773, \text{ for } \phi > 97.5\% \quad (6)$$

where,

MACP = kWh/100 ft³ O₂ product

ϕ = O₂ product purity (mole%)

ASU Total Power

$$MW_{ASU} = 3.798(10)^{-3} \times MACP \times M_{O_2} \quad (7)$$

where,

M_{O_2} = Total oxygen requirement from ASU (lbmole/hr)

FGR Fan

$$MW_{FGR} = 3.255(10)^{-6} \times V_{FG} \times \Delta P_{FGRF} / \eta_{fgrf} \quad (8)$$

where,

V_{FG} = flue gas flow rate (ft³/min)

ΔP_{FGRF} = FGR fan pressure head (psi)

η_{fgrf} = fan efficiency (%), usually 75%

Flue Gas Cooling

$$MW_{FGcooling} = 4.7(10)^{-5} \times M_{cooling} \quad (9)$$

where,

$M_{cooling}$ = cooling water flow rate (gpm)

CO₂ Purification and Compression

$$MW_{compr_purif} = (e_{compr} / 1000 + e_{purif}) \times M_{CO_2} \quad (10)$$

where,

M_{CO_2} = total mass of CO₂ captured (ton/hr)

e_{purif} = 0.109 MWh/ton, for high purity product (purity > 97.5%)

= 0.0018 MWh/ton, for low purity product [13]

e_{compr} = $[-51.632 + 19.207 \times \ln(P_{CO_2} + 14.7)] / (1.1 \times \eta_{compr}/100)$, kWh/ton

P_{CO_2} = CO₂ product pressure (psig)

η_{compr} = CO₂ compression efficiency (%), usually 80%

Net Power Generation

Finally, calculate the net power generation based on user-specified gross output and calculate energy requirements for all environmental control units, including the oxyfuel combustion system.

Performance Model Results

The oxyfuel system model is able to estimate the key intermediate and final results.

Final Result Parameters

These are the results a user is most likely to be interested in. They include:

- CO₂ product flow rate
- Environmental emissions
- Total energy penalty
- Net power output
- Plant heat rate

Intermediate Result Parameters

These additional parameters, which are estimated based on other user-specified input parameters, are crucial in calculating the key result parameters. They include:

- Boiler efficiency
- Oxygen product flow rate from ASU
- Flue gas recycle flow rate

Cost Model

The cost model for the oxyfuel system for CO₂ capture is directly linked to the performance model, and follows the framework used elsewhere in the IECM [35] to ensure consistency in economic calculations. There are four types of costs calculated by this model based on available data: capital cost, operating and maintenance (O&M) cost, incremental cost of electricity (COE), and cost of CO₂ avoidance.

A conventional pulverized coal plant consists of a base plant (consisting of boiler, steam turbine, air preheater), and environmental control units such as ESP, FGD and SCR system. All these process areas have their own capital and O&M costs associated with them, and IECM calculates each of them. The oxyfuel combustion system for CO₂ control requires special equipment/process units in addition to the units mentioned above. The cost model described here reports the costs associated with only the additional units required for the oxyfuel system. The costs of the remainder of the plant are calculated by the IECM model, depending on the new plant versus retrofit application case as explained later in this chapter.

Capital Cost

The total capital requirement (TCR) of a system is calculated as the sum of the installed equipment costs (called the process facilities capital, PFC, which depends on one or more performance variables that determine the size or capacity of each component), plus various indirect costs that are typically estimated as fractions of the process facilities cost following the EPRI cost estimating guidelines [36].

The PFC of the oxyfuel combustion system for CO₂ capture consists of several cost areas, most of which are scaled using a 0.6 cost scaling index and adjusted using the plant cost index as follows:

$$C_i = C_{i,\text{reference}} \times (X_i / X_{i,\text{reference}})^{0.6} \times (\text{PCI} / \text{PCI}_{\text{ref}}), \quad (11)$$

where:

C_i = installed capital cost of cost area (i) for a case study of interest

$C_{i,\text{reference}}$ = reference cost for cost area (i), for a particular reference case

X = scaling parameter relevant to the cost area, such as the flue gas flow rate, gross plant size, or CO₂ product flow rate

X_i = value of the relevant scaling parameter (i) for the case study of interest

$X_{i,\text{reference}}$ = value of the corresponding scaling parameter (i) for the reference case study

PCI = Plant cost index for the year in which the capital cost is being calculated

PCI_{ref} = Plant cost index for the year in which the reference cost was reported

The plant cost indices are listed in Table 4.

Table 4. Plant cost index (PCI) by year (Chemical Engineering magazine)

| Year | Cost Index | Year | Cost Index | Year | Cost Index |
|------|------------|------|------------|------|------------|
| 1977 | 204.1 | 1987 | 323.8 | 1997 | 386.5 |
| 1978 | 218.8 | 1988 | 342.5 | 1998 | 389.5 |
| 1979 | 238.7 | 1989 | 355.4 | 1999 | 390.6 |
| 1980 | 261.1 | 1990 | 357.6 | 2000 | 394.1 |
| 1981 | 297.0 | 1991 | 361.3 | 2001 | 394.3 |
| 1982 | 314.0 | 1992 | 358.2 | 2002 | 395.6 |
| 1983 | 316.9 | 1993 | 359.2 | 2003 | 402.0 |
| 1984 | 322.7 | 1994 | 368.1 | 2004 | 444.2 |
| 1985 | 325.3 | 1995 | 381.1 | 2005 | 468.2 |
| 1986 | 318.4 | 1996 | 381.7 | 2006 | 499.6 |

The oxyfuel system cost areas may be broadly categorized into three categories, namely those related to oxygen production (air separation unit), those related to flue gas recirculation and heat integration (flue gas cooler, flue gas recycle fan, flue gas recycle ducting, and oxygen heater), and those related to CO₂ processing (CO₂ compressors and CO₂ purification system). In addition to these, some cost will be associated with boiler modifications required in case of retrofit applications. The cost model for each of these cost areas are described below:

Air Separation Unit

The model is taken from Frey and Rubin [39]. This paper documents mathematical models of coal gasification combined cycle power plants. The cost model is the result of a statistical study of several oxygen plants that are incorporated into power plants. The cost equation is stated below. It gives the process facilities cost of the air separation unit in thousands of 1989 dollars [37].

$$C_{ASU,ref} = \frac{14.35 \times N_t \times T_a^{0.067}}{(1 - \phi)^{0.073}} \left(\frac{M_{ox}}{N_o} \right)^{0.852} \quad (12)$$

where,

T_a = Ambient air temperature (°F)

N_t = Total number of production trains

N_o = Number of operating production trains

M_{ox} = Molar flow rate of output oxygen (not oxygen product) (lb-mole/hr)

ϕ = Purity of oxygen product

$$20^\circ F \leq T_a \leq 95^\circ F$$

$$625 \leq \left(\frac{M_{ox}}{N_o} \right) \leq 11,350 \text{ lbmole/hr}$$

$$0.95 \leq \phi \leq 0.995$$

So, the capital cost equation for the air separation unit is as follows:

$$C_{ASU} = C_{ASU,ref} \times (PCI / PCI_{ref}) \quad (13)$$

$$C_{ASU} = C_{ASU,ref} \times (PCI / PCI_{1989}) \quad (14)$$

where,

$C_{ASU,ref}$ is calculated using equation (4-2).

Flue Gas Recycle Fan

The cost of the fan required for recycling part of the flue gas is scaled on the basis of the flow rate of the flue gas being recycled (V_{FGR} , ft³/min). The reference cost for the fan is 2 M\$, corresponding to a flue gas flow rate of 6.474(10)⁵ ft³/min (actual) [38].

$$C_{FGR_fan} = C_{FGR_fan,ref} \times (V_{FGR} / V_{FGR,ref})^{0.6} \times (PCI / PCI_{ref}) \quad (15)$$

$$C_{FGR_fan} (\$M) = 2.0 \times [V_{FGR} / 6.474(10)^5]^{0.6} \times (PCI / PCI_{1998}) \quad (16)$$

Flue Gas Recycle Ducting

Additional ducting is necessary to recycle part of the flue gas in the oxyfuel combustion system. The cost of this ducting is assumed to be a function of the flow rate of recycled flue gas. The reference cost is 10 M\$, corresponding to a flue gas flow rate of 6.474(10)⁵ ft³/min (actual) [9].

$$C_{FGR_ducting} = C_{FGR_ducting,ref} \times (V_{FGR} / V_{FGR,ref})^{0.6} \times (PCI / PCI_{ref}) \quad (17)$$

$$C_{FGR_ducting} (\$M) = 10.0 \times [V_{FGR} / 6.474(10)^5]^{0.6} \times (PCI / PCI_{2001}) \quad (18)$$

Flue Gas Cooler

The cost of the flue gas cooler is scaled on the basis of the flow rate of the flue gas assuming the desired flue gas temperature similar to that used in the reference study. The reference cost for the direct contact cooler is 17.6 M\$, corresponding to a plant size of 500 MW gross, and treating a flue gas flow rate of 809,763 ft³/min (actual) entering the cooler at 144°F [9].

$$C_{FG_DCC} = C_{FG_DCC,ref} \times (V_{FG} / V_{FG,ref})^{0.6} \times (PCI / PCI_{ref}) \quad (19)$$

$$C_{FG_DCC} (\$M) = 17.6 \times (V_{FG} / 809,763)^{0.6} \times (PCI / PCI_{2001}) \quad (20)$$

Oxygen Heater

In addition to the air preheater that exists in a conventional PC plant, the oxyfuel combustion system includes an additional heat exchanger called the “oxygen heater” for better heat integration. The cost of this heat exchanger is scaled on the basis of the gross plant size. The reference cost is 12 M\$, corresponding to a plant size of 500 MW gross [9].

$$C_{APH_OH} = C_{APH_OH,ref} \times (MW_{gross} / MW_{gross,ref})^{0.6} \times (PCI / PCI_{ref}) \quad (21)$$

$$C_{APH_OH} (\$M) = 12 \times (MW_{gross} / 500)^{0.6} \times (PCI / PCI_{2001}) \quad (22)$$

CO₂ Purification System

The cost of the CO₂ purification system depends on the desired purity level of the CO₂ product, and the total CO₂ product flow rate. The cost of a system yielding a high purity product (>99.9%) is estimated to be about \$181,818 per ton CO₂ product/hr, corresponding to a reference product flow rate of 550 ton/hr. It is assumed that this cost would be applicable for purity range above 97.5%. In case of applications where such high product purity is not required, a cheaper system giving a low purity product may be used. Such systems are estimated to cost about \$18,182 per ton CO₂ product/hr, corresponding to a reference product flow rate of 660 ton/hr [13].

$$C_{\text{CO}_2\text{purif}} = C_{\text{CO}_2\text{purif,ref}} \times M_{\text{CO}_2\text{pdt}} \times (M_{\text{CO}_2\text{pdt}} / M_{\text{CO}_2\text{pdt,ref}})^{0.6} * (\text{PCI} / \text{PCI}_{\text{ref}}) \quad (23)$$

where,

$$M_{\text{CO}_2\text{pdt}} = \text{CO}_2 \text{ product flow rate, ton/hr}$$

So, for the high purity CO₂ product:

$$C_{\text{CO}_2\text{purif}} (\text{\$M}) = 0.2 \times (M_{\text{CO}_2\text{pdt}} / 1.1) \times (M_{\text{CO}_2\text{pdt}} / 550)^{0.6} \times (\text{PCI} / \text{PCI}_{1995}) \quad (24)$$

And for the low purity CO₂ product:

$$C_{\text{CO}_2\text{purif}} (\text{\$M}) = 0.02 \times (M_{\text{CO}_2\text{pdt}} / 1.1) \times (M_{\text{CO}_2\text{pdt}} / 660)^{0.6} \times (\text{PCI} / \text{PCI}_{1995}) \quad (25)$$

CO₂ Compression System

The multi-stage compression unit with inter-stage cooling and drying yields the final CO₂ product at the specified pressure (about 2000 psig) that contains only acceptable levels of moisture and other impurities (e.g. N₂). Multiple trains will be required if the total CO₂ product flow rate is larger than 300 ton/hr. The size (and cost) of this unit will be a function of the CO₂ product compression power, and may be estimated as follows [38]:

$$C_{\text{CO}_2\text{compr}} = C_{\text{CO}_2\text{compr,ref}} \times (\text{hp}_{\text{CO}_2\text{comp}} / \text{hp}_{\text{CO}_2\text{comp,ref}})^{0.7} \times (\text{PCI} / \text{PCI}_{\text{ref}}) \quad (26)$$

$$C_{\text{CO}_2\text{compr}} = 16.85 \times (\text{hp}_{\text{CO}_2\text{comp}} / 51,676)^{0.7} \times (\text{PCI} / \text{PCI}_{1998}) \quad (27)$$

where,

$$\text{hp}_{\text{CO}_2\text{comp}} = \text{CO}_2 \text{ product compression power requirement (hp)}.$$

In addition to the above mentioned cost areas, there will be cost of boiler modifications required in case of a retrofit application, discussed later in this section.

Boiler Modifications

In case of a *pre-existing* PC plant being retrofitted for CO₂ capture, the boiler must be modified to suit the new oxyfuel combustion system. The cost for these modifications has been estimated to be about 4% of the cost of the boiler [9].

$$C_{\text{boiler_mod}} = 0.04 \times C_{\text{boiler}} \times (\text{PCI} / \text{PCI}_{2001}), \text{ for retrofit application} \quad (28)$$

$$C_{\text{boiler_mod}} = 0, \text{ for new plant case (default)} \quad (29)$$

The sum of these individual process area equipment costs gives the total process facilities capital (PFC). The various indirect costs are then estimated as fractions of the PFC following the EPRI cost estimating guidelines [36]. Table 5 lists the elements of total capital cost. Because of data limitations some of the indirect cost factors for the amine system are estimated based on other similar technologies.

The total plant cost (TPC) is the sum of the process facilities capital (PFC), general facilities capital, engineering and home office fees, and contingencies (project and process). The project contingency is a capital cost factor covering the cost of additional equipment or other costs that would result from a more detailed design at an actual site. The process contingency is a capital cost factor (added cost) applied to a technology to reflect its level of maturity. TPC is developed on the basis of instantaneous (“overnight”) construction occurring at a single point in time, and is generally expressed in mid-year dollars of a (user-specified) reference year.

The total capital requirement (TCR) includes all the capital necessary to complete the entire project, including interest during construction (AFUDC, allowance for funds during construction) and owner costs, which include royalties, startup costs, and inventory capital.

Table 5. Oxyfuel combustion system capital cost model parameters and nominal values

| | Capital Cost Elements | Value |
|---|--|-------------------------------|
| A | Process Area Equipment Costs | (See Eqns (12) to (29) above) |
| B | Total Process Facilities Capital (PFC) | ΣC_i |
| C | Engineering and Home Office | 7% PFC |
| D | General Facilities | 10% PFC |
| E | Project Contingency | 15% PFC |
| F | Process Contingency | 5% PFC |
| G | Total Plant Cost (TPC) = sum of above | B+C+D+E+F |
| H | AFUDC (interest during construction) | Calculated |
| I | Royalty Fees | 0.5% PFC |
| J | Pre-production | 1 month’s fixed O&M cost |
| K | Pre-production | 1 month’s variable O&M cost |
| L | Inventory (startup) Cost | 0.5% TPC |
| M | Total Capital Requirement (TCR) | G+H+I+J+K+L |

O&M Costs

The major operating and maintenance (O&M) cost consists of the fixed costs and variable cost elements listed in Table 6.

Fixed O&M Costs

The *fixed O&M* (FOM) costs in the model include the costs of maintenance (materials and labor) and labor (operating labor, administrative and support labor). They are estimated on annual basis (\$/yr) as follows:

$$FOM = FOM_{\text{labor}} + FOM_{\text{maint}} + FOM_{\text{admin}} \quad (30)$$

$$FOM_{\text{labor}} = \text{labor} \times N_{\text{labor}} \times 40(\text{hrs/week}) \times 52(\text{weeks/yr}) \quad (31)$$

$$FOM_{\text{maint}} = \Sigma_i (f_{\text{maint}})_i \times TPC_i \quad \text{where } i = \text{process area} \quad (32)$$

$$FOM_{\text{admin}} = f_{\text{admin}} \times (FOM_{\text{labor}} + f_{\text{maintlab}} \times FOM_{\text{maint}}) \quad (33)$$

where,

labor = the hourly wages to the labor (\$/hr) = \$24.82/hr

N_{labor} = number of operating labor required = 2

$(f_{\text{maint}})_i$ = total annual maintenance cost expressed as the fraction of the total plant cost (TPC) = 0.04 for all areas

f_{maintlab} = fraction of maintenance cost allocated to labor = 0.4

f_{admin} = the administrative labor cost expressed as the fraction of the total labor cost = 0.3

Table 6. Oxyfuel combustion system O&M cost model parameters and nominal values

| O&M Cost Elements | Typical Value |
|---|--|
| Fixed O&M Costs | |
| Total Maintenance Cost | 4% TPC |
| Maintenance Cost Allocated to Labor (f_{maintlab}) | 40% of total maint. Cost |
| Admin. & Support Labor Cost (f_{admin}) | 30% of total labor cost |
| Operating Labor (N_{labor}) | 2 jobs/shift |
| Variable O&M Costs | |
| Chemicals Cost | \$0.26/ton CO ₂ |
| Waste Water Treatment Cost | n/a |
| CO ₂ Transport Cost | \$0.03/ton CO ₂ per mile [33] |
| CO ₂ Storage/Disposal Cost | \$4.55/ton CO ₂ [33] |

Variable O&M Costs

The *variable O&M* (VOM) costs include costs of chemicals consumed (if any, in CO₂ purification and drying), utilities (water, power), and services used (waste water treatment, CO₂ transport and storage). These quantities are determined in the performance model. The unit cost of each item (e.g., dollars per ton of reagent, or dollars per ton of CO₂ stored) is a parameter specified as a cost input to the model. The total annual cost of each item is then calculated by multiplying the unit cost by the total annual quantity used or consumed. Total annual quantities depend strongly on the plant capacity factor.

The individual components of the variable O&M cost are a function of the annual hours of operation (HPY). The following equation describes this value:

$$\text{HPY} = (\text{PCF} / 100) \times 365 \times 24 \text{ (hrs/yr)} \quad (34)$$

Chemicals

A small quantity of chemicals is used in this process, including the ASU chemicals, desiccant and lubricants. The aggregate cost of these chemicals is calculated from the reference study by normalizing the total cost of chemical consumption reported (\$613,400/yr) over the flow rate of CO₂ captured (400 ton/hr) [9].

$$\text{VOM}_{\text{chemicals}} = \text{UC}_{\text{Chemicals}} M_{\text{CO}_2} \times \text{HPY} \quad (35)$$

where

$\text{UC}_{\text{Chemicals}}$ = unit cost of the chemicals used, averaged at \$0.26/ton CO₂ captured and M_{CO_2} is the flow rate of CO₂ captured (ton/hr).

Wastewater Treatment

It is not clear if the moisture condensed from the flue gas needs to be treated in a wastewater treatment plant. If yes, the cost would be based on the quantity estimated in the performance model as:

$$VOM_{\text{wastewater}} = M_{\text{wastewater}} \times UC_{\text{wastewater}} \times \text{HPY} \quad (36)$$

where,

$$UC_{\text{wastewater}} = \text{unit cost of wastewater treatment.}$$

CO₂ Transport

Transportation of CO₂ product is assumed to take place via pipelines. The cost of CO₂ transport is estimated on the basis of two user-specified parameters namely transportation distance (TD, in km) and unit cost of transport (UC_{transport}, \$/km per tonne CO₂), plus the CO₂ product flow rate (calculated result from performance model).

$$VOM_{\text{transport}} = M_{\text{CO}_2} \times UC_{\text{transport}} \times \text{TD} \times \text{HPY} \quad (37)$$

CO₂ Storage

Depending upon the method of CO₂ disposal or storage, either there may be some revenue generated (as in enhanced oil recovery, or enhanced coal bed methane), or an additional cost (all other disposal methods). The total cost or revenue of CO₂ disposal/storage is estimated from the unit cost and CO₂ product flow rate (UC_{disp}).

$$VOM_{\text{disposal}} = M_{\text{CO}_2} \times UC_{\text{disp}} \times \text{HPY} \quad (38)$$

Power

By default, all energy costs are handled internally in the model by de-rating the overall power plant based on the calculated power requirement. The CO₂ capture unit is charged for the total electricity production foregone because of CO₂ capture and compression (E_{CO₂, tot}).

For power plants with multi-pollutant controls the desire to quantify costs for a single pollutant requires an arbitrary choice of how to charge or allocate certain costs. This is especially relevant for energy-intensive processes like CO₂ capture systems.

The unit cost of electricity (COE_{noct1}) is estimated by the base plant module, or may be overridden by a user-specified value if this energy is assumed to be supplied from an external source. Since energy cost is one of the biggest O&M cost items for the CO₂ unit, the way in which it is accounted for is important when calculating the mitigation cost.

$$VOM_{\text{energy}} = E_{\text{CO}_2, \text{tot}} \times \text{HPY} \times \text{COE}_{\text{noct1}} \quad (39)$$

The total variable O&M (VOM, \$/yr) cost is obtained by adding all these costs:

$$VOM = VOM_{\text{chemicals}} + VOM_{\text{wastewater}} + VOM_{\text{transport}} + VOM_{\text{disposal}} + VOM_{\text{energy}} \quad (40)$$

Finally, the total annual O&M cost (TOM, \$/yr) may be obtained as:

$$\text{TOM} = \text{FOM} + \text{VOM} \quad (41)$$

Incremental Cost of Electricity

Once the total capital requirement and the total O&M costs are known, the total annualized cost of the power plant may be estimated as follows:

$$\text{Total annual revenue requirement, TRR (\$/yr)} = (\text{TCR} \times \text{CRF}) + \text{TOM} \quad (42)$$

where,

TCR = Total capital requirement of the power plant (\$)

CRF = Capital recovery factor (fraction)

The capital recovery factor, or fixed charge factor (FCF), is the factor that annualizes the total capital requirement of the plant. It depends on the applicable interest rate (or discount rate) and useful lifetime of the plant. The details of the capital recovery factor can be found elsewhere [36].

It can be seen that a higher value of this factor (*e.g.* from assumptions of shorter plant life and/or higher interest rate) leads to a higher overall annualized cost. Hence the assumption about this factor (a user-defined parameter) is crucial in the overall economics of the plant.

The IECM framework calculates the cost of electricity (COE) for the overall power plant by dividing the total annualized plant cost (\$/yr) by the net electricity generated (kWh/yr). Results are expressed in units of \$/MWh (equivalent to mills/kWh).

$$\text{Cost of electricity, COE (\$/MWh)} = \text{TRR} / (\text{MW}_{\text{net}} \times \text{HPY}) \quad (43)$$

where,

TRR = Total annual revenue requirement (\$/yr)

MW_{net} = Net power generation capacity (MW)

HPY = Annual hours of operation (hrs/yr)

Note that the COE includes the cost of all environmental control systems, not just the CO₂ control system. Thus, by running two scenarios of the power plant model, one without CO₂ capture (reference plant) and one with CO₂ capture (CO₂ capture plant), the incremental capital costs, O&M costs, and total annualized costs attributed to CO₂ capture are obtained. The addition of a CO₂ capture and sequestration system increases the COE for the plant; this incremental cost of electricity is attributed to CO₂ control.

Cost of CO₂ Avoidance

Analysts often express the cost of an environmental control system in terms of the cost per unit mass of pollutant removed. However, for energy-intensive CO₂ controls there is a big difference between the cost per tonne CO₂ “removed” and the cost per tonne “avoided” based on *net* plant capacity. Since the purpose of adding a CO₂ unit is to reduce the CO₂ emissions per net kWh delivered, the “cost of CO₂ avoidance” is the economic indicator that is widely used in this field. It can be calculated as:

$$\begin{aligned} \text{Cost of CO}_2 \text{ Avoided (\$/t)} = & \\ & \frac{(\$ / \text{kWh})_{\text{after}} - (\$ / \text{kWh})_{\text{before}}}{(\text{tonne CO}_2 / \text{kWh})_{\text{before}} - (\text{tonne CO}_2 / \text{kWh})_{\text{after}}} \end{aligned} \quad (44)$$

In contrast, the cost per unit of CO₂ removed or captured is simply the additional expenses incurred in the capture of CO₂, divided by the total quantity of CO₂ captured. This can be calculated as the difference between the total annualized cost of the plant (TRR, M\$/yr) with and without CO₂ control, divided by the total quantity of CO₂ captured (tonne CO₂/yr), with the net power generated by the two plants remaining the same. Hence, the CO₂ avoidance cost, as calculated in equation 4-33, is quite different from the cost per unit of CO₂ captured. In case of CO₂ control using an energy-intensive technology like amine-scrubbing, the *cost of CO₂ avoidance* may be substantially higher than *cost of CO₂ capture*.

Case Study

The application of the performance and cost model may be illustrated using a case study of a power plant. Let us consider the case of an existing conventional coal-fired power plant, and impact of modifying it to oxyfuel combustion plant to obtain a concentrated CO₂ product stream.

Input Parameters

The basic assumptions and input parameters are listed in Table 7. These can be entered into the IECM [1].

Table 7. Design parameters for case study of a pulverized coal plant with CO₂ control using O₂/CO₂ recycle (oxyfuel combustion) system

| Parameter | Value | Parameter | Value |
|--------------------------------|-------------------|--|--|
| Gross plant size (MW) | 500 | Emission standards | 2000 NSPS ¹ |
| Base plant steam cycle type | SC ² | NO _x Controls | LNB ³ |
| Gross plant heat rate (kJ/kWh) | 9325 ⁴ | Particulate Control | ESP ⁵ |
| Plant capacity factor (%) | 75 | SO ₂ Control | FGD ⁶ |
| Coal characteristics | | CO ₂ Control | O ₂ /CO ₂ ⁷ |
| Rank | Bit. | CO ₂ product purity (%) | 97.5 |
| HHV (kJ/kg) | 30,776 | CO ₂ capture efficiency (%) | 90 |
| % S | 2.13 | CO ₂ product pressure (kPa) | 13,790 |
| % C | 73.81 | Distance to storage (km) | 165 |
| Delivered cost (\$/tonne) | 37.10 | Cost year basis (constant dollars) | 2000 |
| Delivered cost (\$/GJ) | 1.203 | Fixed charge factor | 0.15 ⁸ |

¹ NO_x = 65 ng/J, PM = 13 ng/J, SO₂ = 81% removal (assumed to be the same as that of the reference plant case)

² Nominal case is a sub-critical unit

³ LNB = Low- NO_x Burner

⁴ Gross heat rate of the plant improves to 8,841 kJ/kWh after switching to oxyfuel combustion mode, due to higher boiler efficiency

⁵ ESP = Electrostatic Precipitator

⁶ FGD = Flue Gas Desulfurization

⁷ O₂/CO₂ = Oxyfuel combustion system with flue gas recycle

⁸ Corresponds to a 30-year plant lifetime with a 14.8% real interest rate (or, a 20-year life with 13.9% interest)

The reference plant (without CO₂ control unit) is a New Source Performance Standard (NSPS) compliant coal-fired power plant and the complete plant with multi-pollutant environmental controls is simulated using IECM. Wyoming Powder River Basin coal has been assumed to be used. The model outputs are presented later in Table 5.2 in comparison with the estimates for the CO₂ capture plant.

In case of the CO₂ capture plant, the following changes have been assumed as compared to the reference plant:

- Pure oxygen (95% purity) mixed with recycled flue gas is used as oxidant, in place of air.
- Excess air (or oxygen) level is reduced to 5% (reference plant uses the default value which is about 20%).
- Air leakage has been reduced to 2% (reference plant uses the default value which is about 19%).
- The boiler efficiency improves to 94.03% in case of oxyfuel combustion system, as compared to 89.16% for the reference plant using air.
- CO₂ handling system including CO₂ product purification, compression, transport and storage has been added.

The values for other parameters are listed in Table 3, Table 4, Table 5 and Table 6 in previous sections.

Performance

Coal Flow Rate

The required coal flow rate for this illustration is calculated using Equation (1):

$$\begin{aligned}M_{\text{coal}} &= M_{\text{coal}} = \frac{MW_g \times HR_{\text{steam}}}{2 \times \eta_{\text{boiler}} \times HHV_{\text{coal}}} \\ &= \frac{500 \text{ MW} \times 7,880 \text{ Btu} / \text{kWh}}{2 \times 0.9404 \times 13,260 \text{ Btu} / \text{lb}} \\ &= 158 \text{ ton/hr (or 143.3 tonne/hr)}\end{aligned}$$

Oxygen Requirement

Stoichiometric Oxygen

The stoichiometric O₂ requirement is calculated on the basis of the coal flow rate and coal composition. The results are shown in Table 8.

Table 8. Coal properties and associated oxygen requirements for stoichiometric combustion.

| Coal component | Mol. Wt. | wt% | ton/hr | O ₂ ton/ton | O ₂ req. ton/hr |
|----------------|----------|-------|--------|------------------------|----------------------------|
| C | 12 | 73.81 | 116.6 | 2.7 | 331.0 |
| H | 2 | 4.88 | 7.7 | 8.0 | 61.7 |
| O | 32 | 5.41 | 8.5 | -1 | -8.5 |
| S | 32 | 2.13 | 3.4 | 1 | 3.4 |
| N | 28 | 1.42 | 2.2 | 0.095 ¹ | 0.2 |
| Total | | | | | 367.7 |

So, the theoretical O₂ requirement is 367.7 ton/hr, or about 22,970 lbmole/hr.

Required Oxygen

With 5% excess oxygen, the total amount of O₂ required can be calculated.

$$M_{O_2, req} = 1.05 \times 367.7 = 386 \text{ ton/hr}$$

Required Oxidant

The oxygen product is 95% pure. Hence, the total amount of oxygen product or oxidant supplied by the ASU can be calculated.

$$\begin{aligned}M_{ox} &= 386 / 0.95 = 406.4 \text{ ton/hr} \\ &= 369 \text{ tonne/hr}\end{aligned}$$

¹ Estimated on the basis of NO_x emission rate of 0.1885 lb/MBtu and assuming 95% of NO_x is NO

Air Leakage

Air leakage stream is defined on the basis of theoretical air (oxygen) requirement. It is assumed that the air leakage is 2% which means the air leakage stream contains oxygen equivalent to 2% of theoretical oxygen requirement for combustion. So, the amount of oxygen in air leakage stream = $0.02 \times 367.7 = 7.4$ ton/hr. Air contains about 22.8% w/w of oxygen. So, the air leakage stream is estimated to be = $7.4 / 0.228 = 32.3$ ton/hr.

Combustion Product

The combustion products and composition of the flue gas stream is estimated on the basis of combustion reaction stoichiometry, and other known input streams.

Table 9. Combustion products of the flue gas stream. All values are in units of lb-mole/hr.

| Component | Combustion Products | Oxidant | Sub-total | Air Leakage | Total |
|------------------|---------------------|-----------------|-----------------|----------------|-----------------|
| N ₂ | 134.4 | 194.4 | 328.8 | 1,712.1 | 2,040.9 |
| O ₂ | -22,967.9 | 24,116.3 | 1,148.4 | 459.4 | 1,607.8 |
| H ₂ O | 7,710.4 | 0.0 | 7,710.4 | 63.2 | 7,773.6 |
| CO ₂ | 19,436.6 | 0.0 | 19,436.6 | 0.0 | 19,426.6 |
| CO | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| HCl | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| SO ₂ | 208.7 | 0.0 | 208.7 | 0.0 | 208.7 |
| SO ₃ | 1.7 | 0.0 | 1.7 | 0.0 | 1.7 |
| NO | 25.0 | 0.0 | 25.0 | 0.0 | 25.0 |
| NO ₂ | 0.9 | 0.0 | 0.9 | 0.0 | 0.9 |
| NH ₃ | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Ar | 0.0 | 1,074.8 | 10,74.8 | 20.5 | 1,095.3 |
| Total | 4,549.7 | 25,385.5 | 29,935.2 | 2,255.1 | 32,190.3 |

Recycled Flue Gas

The flue gas is then passed through the ESP and FGD units to remove particulate matter and sulfur oxides respectively. Next, it is cooled down and most of the water is condensed out. A part of the flue gas is then recycled back into the boiler along with oxygen from ASU.

So, for the next iteration, the oxidant is a mixture of oxygen and recycled flue gas. The oxidant and flue gas streams are estimated assuming that part of the oxygen requirement comes from the leakage air and the oxygen content in the recycled flue gas. For 75% flue gas recycle ratio, we get:

Table 10. Oxidant and recycled flue gas composition. All values are in units of lb-mole/hr.

| Component | FGR | O ₂ (theory) | O ₂ (corr.) | Oxidant | Total | FG Out |
|------------------|----------|-------------------------|------------------------|----------|----------|----------|
| N ₂ | 1,530.7 | 194.4 | 181.0 | 1,711.6 | 3,558.1 | 889.5 |
| O ₂ | 1,205.8 | 24,116.3 | 22,451.2 | 23,657.0 | 1,148.4 | 287.1 |
| H ₂ O | 1,275.3 | 0.0 | 0.0 | 1,275.3 | 9,048.9 | 2,262.2 |
| CO ₂ | 14,577.5 | 0.0 | 0.0 | 14,577.5 | 34,014.1 | 8,503.5 |
| CO | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| HCl | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| SO ₂ | 156.5 | 0.0 | 0.0 | 156.5 | 365.1 | 91.3 |
| SO ₃ | 1.3 | 0.0 | 0.0 | 1.3 | 2.9 | 0.7 |
| NO | 18.8 | 0.0 | 0.0 | 18.8 | 43.8 | 10.9 |
| NO ₂ | 0.6 | 0.0 | 0.0 | 0.6 | 1.5 | 0.4 |
| NH ₃ | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| AR | 821.5 | 1,074.8 | 1,000.6 | 1,822.1 | 1,842.5 | 460.6 |
| Total | 19,587.9 | 25,385.5 | 23,632.7 | 43,220.6 | 50,025.4 | 12,506.4 |

After several iterations, we get the following stable solution:

Table 11. Final oxidant and recycled flue gas composition All values are in units of lb-mole/hr.

| Component | FGR | O ₂ (req.) | Oxidant | Total | FG Out | FGR |
|------------------|----------|-----------------------|----------|-----------|----------|----------|
| N ₂ | 6,093.4 | 184.7 | 6,278.1 | 8,124.6 | 2,031.1 | 6,093.4 |
| O ₂ | 1,205.8 | 22,910.5 | 24,116.3 | 1,607.8 | 401.9 | 1,205.8 |
| H ₂ O | 1,525.7 | 0.0 | 1,525.7 | 9,299.2 | 2,324.8 | 1,525.7 |
| CO ₂ | 58,309.3 | 0.0 | 58,309.3 | 7,7745.9 | 19,436.5 | 58,309.3 |
| CO | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| HCl | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| SO ₂ | 29.7 | 0.0 | 29.7 | 39.6 | 9.9 | 29.7 |
| SO ₃ | 0.2 | 0.0 | 0.2 | 0.3 | 0.1 | 0.2 |
| NO | 75.0 | 0.0 | 75.0 | 100.0 | 25.0 | 75.0 |
| NO ₂ | 2.6 | 0.0 | 2.6 | 3.4 | 0.9 | 2.6 |
| NH ₃ | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| Ar | 3,124.6 | 1,021.1 | 4,145.7 | 4,166.2 | 1,041.5 | 3,124.6 |
| Total | 70,366.3 | 24,116.3 | 94,482.6 | 101,087.0 | 25,271.8 | 70,366.3 |

CO₂ Product Stream

The CO₂ product flow rate is estimated as follows:

$$\text{Total CO}_2 \text{ captured} = 0.90 \times 19,436.5 = 17,492.9 \text{ lbmole/hr} = 384.8 \text{ ton/hr}$$

At 97.5% purity level, the total product flow rate would be about 17,942 lbmole/hr or 394 ton/hr.

Power Requirement

The energy requirement for various items are calculated in the following subsections.

ASU Unit Power

$$\begin{aligned} \text{MACP} &= 0.0049\phi + 0.4238, \text{ for } \phi \leq 97.5\% \\ &= 0.0736 / (100 - \phi)^{1.3163} + 0.8773, \text{ for } \phi > 97.5\% \end{aligned}$$

where,

$$\text{MACP} = \text{kWh}/100 \text{ ft}^3 \text{ O}_2 \text{ product}$$

Here,

$$\phi = \text{O}_2 \text{ product purity (mole\%)} = 95\%$$

So,

$$\text{MACP} = 0.0049 \times 95 + 0.4238 = 0.8893 \text{ kWh}/100 \text{ ft}^3 \text{ O}_2 \text{ product}$$

ASU Total Power

$$\text{MW}_{\text{ASU}} = 3.798(10)^{-3} \times \text{MACP} \times M_{\text{O}_2}$$

Where,

$$M_{\text{O}_2} = \text{Total oxygen requirement from ASU} = 22,911 \text{ lbmole/hr}$$

So,

$$\text{MW}_{\text{ASU}} = 3.798(10)^{-3} \times 0.8893 \times 22,911 = 77.38 \text{ MW}$$

FGR Fan

$$\text{MW}_{\text{FGR}} = 3.255(10)^{-6} \times V_{\text{FG}} \times \Delta P_{\text{FGRF}} / \eta_{\text{fgrf}}$$

where,

$$V_{\text{FG}} = \text{flue gas flow rate (ft}^3/\text{min)}$$

$$\Delta P_{\text{FGRF}} = \text{FGR fan pressure head (psi)}$$

$$\eta_{\text{fgrf}} = \text{fan efficiency (\%), usually 75\%}$$

Here,

$$\text{recycled flue gas flow rate} = 70,366 \text{ lbmole/hr}$$

At 100 deg F, the volumetric flow rate of this stream would be about 438,620¹ ft³/min. So,

$$\text{MW}_{\text{FGR}} = 3.255(10)^{-6} \times 438,620 \times 0.14 / 0.75 = 0.27 \text{ MW}$$

¹ $V = 22.4 \text{ (m}^3/\text{kgmole)} \times 70,366 \text{ (lbmole/hr)} \times (\text{kg}/2.2 \text{ lb}) \times (311/298) \times (\text{hr}/60 \text{ min}) \times (\text{ft}^3/0.02832 \text{ m}^3) = 438,620 \text{ (ft}^3/\text{min)}$

Flue Gas Cooling

$$MW_{FGcooling} = 4.7(10)^{-5} \times M_{cooling}$$

Now,

$$M_{cooling} \text{ (gpm)} = 3.3(10)^{-3} \times V_{fg} \times \Delta T$$

So,

$$M_{cooling} = 3.3(10)^{-3} \times 438,620 \times 40 = 57,900 \text{ gpm}$$

Hence,

$$MW_{FGcooling} = 4.7(10)^{-5} \times 57,900 = 2.7 \text{ MW}$$

CO₂ Purification and Compression

$$MW_{compr_purif} = (e_{comp} + e_{purif}) \times M_{CO_2}$$

where,

$$M_{CO_2} = \text{total mass of CO}_2 \text{ captured (ton/hr)} = 384.8 \text{ ton/hr}$$

$$e_{purif} = 0.0018 \text{ MWh/ton, for low purity product}$$

$$P_{CO_2} = \text{CO}_2 \text{ product pressure (psig)} = 2000$$

$$\eta_{comp} = \text{CO}_2 \text{ compression efficiency (\%)} = 80\%$$

So,

$$\begin{aligned} e_{comp} &= [-51.632 + 19.207 \times \ln(2,000 + 14.7)] / (1.1 \times 80 / 100), \text{ kWh/ton} \\ &= 107.39 \text{ kWh/ton} \end{aligned}$$

So,

$$MW_{compr_purif} = (0.1074 + 0.0018) \times 384.8 = 42.02 \text{ MW}$$

Net Power Generation

The net power generation is calculated by summing the power requirements in the subsections described above and subtracting that power from the gross power generated in the power plant. This is the power that is available for export and use outside the power plant.

The energy consumption from the subsections above is as follows:

$$MW_{use} = 77.38 + 0.027 + 2.7 + 42.02 = 122.13 \text{ MW}$$

The net power output of the plant can be estimated based on the gross output (500 MW) and the energy requirements for all environmental control units.

$$MW_{net} = 500 - 122.13 = 377.83 \text{ MW}$$

Direct Capital Cost

The capital costs are estimated using the equations in the Capital Cost section discussed earlier. Please refer to those previous sections for the governing equations, references, and explanations. Each process area in the power plant and the associated capital costs are given in the following subsections.

Note also that all costs are reported in \$M for year 2000 US\$ for illustration purposes. To convert costs to other years, please refer to Table 4 and substitute the appropriate cost index for the year of interest for the “PCI” term in each cost equation.

Air Separation Unit (ASU)

Maximum train capacity = 11350 lbmole/hr

Hence, three operating trains would be required.

$$C_{ASU,ref} = \frac{14.35 \times N_t \times T_a^{0.067}}{(1-\phi)^{0.073}} \left(\frac{M_{ox}}{N_o}\right)^{0.852}$$

where,

$$20^\circ F \leq T_a \leq 95^\circ F$$

$$625 \leq \left(\frac{M_{ox}}{N_o}\right) \leq 11,350 \text{ lbmole/hr}$$

$$0.95 \leq \phi \leq 0.995$$

So,

$$C_{ASU,ref} = \frac{14.35 \times 3 \times 59^{0.067}}{(1-0.95)^{0.073}} \left(\frac{22,911}{3}\right)^{0.852} = \$143,168,000 \text{ (1989 \$)}$$

So, the capital cost equation for the air separation unit is as follows:

$$\begin{aligned} C_{ASU} (\$M) &= C_{ASU,ref} \times (PCI / PCI_{1989}) \\ &= 143.2 \times (394.1 / 355.4) \text{ M\$} \\ &= \$ 158.8 \text{ M} \end{aligned}$$

Flue Gas Recycle Fan

$$C_{FGR_fan} (\$M) = 2.0 \times [V_{FGR} / 6.474(10)^5]^{0.6} \times (PCI / PCI_{1998})$$

So,

$$\begin{aligned} C_{FGR_fan} (\$M) &= 2.0 \times [438,620 / 6.474(10)^5]^{0.6} \times (PCI / PCI_{1998}) \\ &= 1.58 \times (394.1 / 389.5) = \$ 1.6 \text{ M} \end{aligned}$$

Flue Gas Recycle Ducting

$$C_{FGR_ducting} (\$M) = 10.0 \times [V_{FGR} / 6.474(10)^5]^{0.6} \times (394.1 / PCI_{2001})$$

So,

$$C_{\text{FGR_ducting}} (\$M) = 10.0 \times [438,620 / 6.474(10)^5]^{0.6} * (\text{PCI} / \text{PCI}_{2001})$$

$$= 7.9 \times (394.1 / 394.3) = \$ 7.9 \text{ M}$$

Flue Gas Cooler

$$C_{\text{FG_DCC}} (\$M) = 17.6 \times (V_{\text{FG}} / 809,763)^{0.6} \times (\text{PCI} / \text{PCI}_{2001})$$

So,

$$C_{\text{FG_DCC}} (\$M) = 17.6 \times (438,620 / 809,763)^{0.6} \times (\text{PCI} / \text{PCI}_{2001})$$

$$= 17.6 \times (0.692) * (394.1 / 394.3) = \$ 12.2 \text{ M}$$

Oxygen Heater

$$C_{\text{APH_OH}} (\$M) = 12 \times (\text{MW}_{\text{gross}} / 500)^{0.6} \times (\text{PCI} / \text{PCI}_{2001})$$

So,

$$C_{\text{APH_OH}} (\$M) = 12 \times (500 / 500)^{0.6} \times (\text{PCI} / \text{PCI}_{2001})$$

$$= 12 \times (394.1 / 394.3) = \$ 12.0 \text{ M}$$

CO₂ Purification System

For the low purity CO₂ product:

$$C_{\text{CO}_2\text{_purif}} (\$M) = 0.02 \times (M_{\text{CO}_2\text{_pdt}} / 1.1) \times (M_{\text{CO}_2\text{_pdt}} / 660)^{0.6} \times (\text{PCI} / \text{PCI}_{1995})$$

So,

$$C_{\text{CO}_2\text{_purif}} (\$M) = 0.02 \times (394 / 1.1) \times (394 / 660)^{0.6} \times (\text{PCI} / \text{PCI}_{1995})$$

$$= 5.3 \times (394.1 / 381.1) = \$ 5.5 \text{ M}$$

CO₂ Compression System

$$C_{\text{CO}_2\text{_compr}} (\$M) = 16.85 \times (\text{hp}_{\text{CO}_2\text{_compr}} / 51,676)^{0.7} * (\text{PCI} / \text{PCI}_{1998})$$

where,

$$\text{hp}_{\text{CO}_2\text{_compr}} = \text{CO}_2 \text{ product compression power requirement (hp)}.$$

So,

$$C_{\text{CO}_2\text{_compr}} (\$M) = 16.85 \times (55,394^1 / 51,676)^{0.7} * (\text{PCI} / \text{PCI}_{1998})$$

$$= 17.7 \times (394.1 / 389.5) = \$ 17.9 \text{ M}$$

Boiler Modifications

In case of a pre-existing PC plant being retrofitted for CO₂ capture, the boiler must be modified to suit the new oxyfuel combustion system. The cost for these modifications has been estimated to be about 4% of the cost of the boiler [9].

$$C_{\text{boiler_mod}} (\$M) = 0.04 \times C_{\text{boiler}} \times (\text{PCI} / \text{PCI}_{2001}), \text{ for retrofit application}$$

$$C_{\text{boiler_mod}} (\$M) = 0, \text{ for new plant case (default)}$$

¹ $\text{hp}_{\text{CO}_2\text{_compr}} = 107.4 \text{ (kWh / ton)} \times 384.8 \text{ (ton/hr)} \times (\text{hp} / 0.746 \text{ kW}) = 55,394 \text{ hp}$

So,

$$C_{\text{boiler_mod}} (\$M) = 0$$

So, the total process facilities cost (PFC) is sum of the individual costs estimated above, which is \$ 221.3 M.

Other Capital Costs

Next, the indirect capital costs are estimated using Table 5, and hence the total capital requirement (TCR) of the O₂/CO₂ recycle system is estimated.

$$\text{PFC} = \$ 221.3 \text{ M}$$

$$\text{TPC} = \$ 303.2 \text{ M}$$

$$\text{TCR} = \$ 337.9 \text{ M}$$

O&M Costs

The O&M costs for this system are estimated, as in the O&M Costs section previously discussed.

Fixed O&M Costs

The fixed O&M (FOM) costs in the model include the costs of maintenance (materials and labor) and labor (operating labor, administrative and support labor). They are estimated on annual basis (\$/yr) for a \$2000 year basis as follows:

$$\text{FOM} = \text{FOM}_{\text{labor}} + \text{FOM}_{\text{maint}} + \text{FOM}_{\text{admin}}$$

$$\begin{aligned} \text{FOM}_{\text{labor}} &= \text{labor} \times N_{\text{labor}} \times 40(\text{hrs/week}) \times 52(\text{weeks/yr}) \\ &= \$24.82/\text{hr} \times 2 \times 40 \text{ hr/wk} \times 52 \text{ wk/yr} \\ &= \$103,251/\text{yr} \end{aligned}$$

$$\begin{aligned} \text{FOM}_{\text{maint}} &= \sum_i (f_{\text{maint}})_i \times \text{TPC}_i \text{ where } i = \text{process area} \\ &= 0.04 \times \text{TPC} = 0.04 \times \$303.2 \text{ M} = \$ 12,128,000/\text{yr} \end{aligned}$$

$$\begin{aligned} \text{FOM}_{\text{admin}} &= f_{\text{admin}} \times (\text{FOM}_{\text{labor}} + f_{\text{maintlab}} \times \text{FOM}_{\text{maint}}) \\ &= 0.3 \times (103,251 + 0.4 \times 12,128,000) = \$ 1,486,335/\text{yr} \end{aligned}$$

So,

$$\begin{aligned} \text{FOM} &= 103,251 + 12,128,000 + 1,486,335 \\ &= \$ 13.72 \text{ M/yr} \end{aligned}$$

Variable O&M Costs

The variable O&M (VOM) costs are estimated on the basis of Table 6 and the Variable O&M Costs section previously discussed, as follows:

Chemicals

$$\begin{aligned} \text{VOM}_{\text{chemicals}} &= \text{UC}_{\text{chemicals}} \times M_{\text{CO}_2} \times \text{HPY} \\ &= \$0.26/\text{ton CO}_2 \text{ captured} \times 384.8 \text{ ton/hr} \times 6575 \text{ hr/yr} \end{aligned}$$

$$= \$657,815.6/\text{yr}$$

CO₂ transport

$$\begin{aligned} \text{VOM}_{\text{transport}} &= M_{\text{CO}_2} \times \text{UC}_{\text{transport}} \times \text{TD} \times \text{HPY} \\ &= 394 \text{ ton/hr} \times \$0.03/\text{ton.mile} \times 100 \text{ mile} \times 6,575 \text{ hr/yr} \\ &= \$7,771,650/\text{yr} \end{aligned}$$

CO₂ Storage

$$\begin{aligned} \text{VOM}_{\text{disposal}} &= M_{\text{CO}_2} \times \text{UC}_{\text{disp}} \times \text{HPY} \\ &= 394 \text{ ton/hr} \times \$4.55/\text{ton} \times 6,575 \text{ hr/yr} \\ &= \$11,787,003/\text{yr} \end{aligned}$$

Power

$$\begin{aligned} \text{VOM}_{\text{energy}} &= E_{\text{CO}_2,\text{tot}} \times \text{HPY} \times \text{COE}_{\text{noct1}} \\ &= 119.67 \text{ MW} \times 6,575 \text{ hr/yr} \times \$ 37.5 /\text{MWh} \\ &= \$29,506,134/\text{yr} \end{aligned}$$

Total Variable O&M Cost

The total variable O&M (VOM, \$/yr) cost is obtained by adding these particular costs just calculated:

$$\begin{aligned} \text{VOM} &= \text{VOM}_{\text{chemicals}} + \text{VOM}_{\text{transport}} + \text{VOM}_{\text{disposal}} + \text{VOM}_{\text{energy}} \\ &= 0.658 + 7.772 + 11.787 + 29.506 = \$ 49.723 \text{ M/yr} \end{aligned}$$

Total O&M Costs

So, the total O&M cost for the CO₂ capture unit is:

$$\text{TOM} = \text{FOM} + \text{VOM} = \$ 13.72 \text{ M/yr} + \$ 49.723 \text{ M/yr} = \$ 63.44/\text{yr}$$

Total Revenue Required

Finally, the overall annualized cost of the CO₂ capture system using O₂/CO₂ recycle technology can be estimated. The total revenue required is calculated as follows:

$$\text{TRR} (\$/\text{yr}) = (\text{TCR} \times \text{CRF}) + \text{TOM}$$

where,

$$\text{CRF} = \text{Capital recovery factor (fraction)} = 0.148$$

So

$$\text{TRR} = 337.9 \times 0.148 + 63.44 = \$ 113.5 \text{ M/yr}$$

So, the total annualized cost of capturing CO₂ using oxyfuel combustion based O₂/CO₂ recycle system has been estimated to be about \$ 113.5 M/yr.

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