System Design and Analysis for Conceptual Design of Oxygen-Based PC Boiler

Topical Report

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ABSTRACT

The objective of the system design and analysis task of the Conceptual Design of Oxygen-Based PC Boiler study is to optimize the PC boiler plant by maximizing system efficiency. Simulations of the oxygen-fired plant with CO_2 sequestration were conducted using Aspen Plus and were compared to a reference air-fired 460 Mw plant. Flue gas recycle is used in the O_2-fired PC to control the flame temperature. Parametric runs were made to determine the effect of flame temperature on system efficiency and required waterwall material and thickness. The degree of improvement on system efficiency of various modifications including hot gas recycle, purge gas recycle, flue gas feedwater recuperation, and recycle purge gas expansion were investigated. The selected O_2-fired design case has a system efficiency of 30.1% compared to the air-fired system efficiency of 36.7%. The design O_2-fired case requires T91 waterwall material and has a waterwall surface area of only 44% of the air-fired reference case. Compared to other CO_2 sequestration technologies, the O_2-fired PC is substantially better than both natural gas combined cycles and post CO_2 removal PCs and is slightly better than integrated gasification combined cycles.

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1.0 Introduction

This report describes the results and conclusions of Task 2, system design and analysis of the Conceptual Design of Oxygen-Based PC Boiler study. The objective of the Conceptual Design of Oxygen-Based PC Boiler study is to develop a conceptual pulverized coal (PC)-fired power plant, which facilitates the practical capture of carbon dioxide capture for subsequent sequestration. The system design and analysis task, which was performed using the Aspen Plus computer program, is aimed at optimizing the PC boiler plant operating parameters to minimize the overall power plant heat rate. The flow rates and other properties of individual streams of the power plant were calculated as the results of the Aspen Plus simulations. The required performance characteristics of such operating components as pulverized coal-fired furnace, heat recovery area, flue gas recuperator, and economizer were determined.

Two plant configurations were simulated: 1) a conventional air-fired PC power plant and 2) the proposed oxygen-based PC plant. In order to compare the performance of the oxygen-based plant with that of the conventional plant, the power output of the steam turbine in the both plants was kept constant.

2.0 Executive Summary

The objective of the Conceptual Design of Oxygen-Based PC Boiler study is to develop a conceptual pulverized coal-fired power plant, which facilitates the practical capture of carbon dioxide capture for subsequent sequestration. The system design and analysis task, which was performed using the Aspen Plus computer program, is aimed at optimizing the PC boiler plant operating parameters to minimize the overall power plant heat rate.

The reference plant applied is a subcritical pressure, natural circulation boiler firing high-volatile bituminous coal generating 460 MWe. A conventional air-fired case was simulated as the comparison basis. The air-fired plant has a boiler efficiency of 88.2% and a net plant efficiency of 36.7%.

The oxygen-based plant model contains all the components in the conventional plant (with the exception of the FGD) model plus the addition of an air separation unit and a flue gas cooler. Flue gas is recycled is to control the flame temperature inside the PC-fired boiler to minimize NOx formation, minimize ash slagging in the furnace combustion zone, and avoid the application of exotic materials.

Parametric runs were made varying the amount of recycled flue gas (which directly affects the flame temperature) while maintaining the same boiler outlet O_2 concentration as the air-fired case. The results show that by reducing the recycled flue gas flow rate by 33%, the flame temperature increases from 3574ºF to 4337ºF, increasing the system efficiency (without CO_2 compression) from 31.2% to 31.9%. The system efficiency was further increased to 32.1% by raising the temperature of the recycled flue gas from 96ºF to 148ºF.

Equipment to compress and liquefy the CO_2 effluent to 2000 psia and to reduce the moisture to 50 ppm (to avoid transport pipe corrosion) was added to the system model. This equipment reduces the system efficiency of the 3574ºF flame temperature case from 31.2% to 27.9%. Recycling the residual O_2 before CO_2 compression increases the system efficiency with CO_2 compression to 28.3%. For the high temperature case (3574ºF) recycling of the purge gases results in a system efficiency of 29.5%. A further enhancement of this high flame temperature case was simulated by adding a flue gas feedwater heater prior to CO_2 separation and a turbine expander to recover power from the recycled purge gas pressure reduction. The addition of this equipment raised the system efficiency to 29.8% with a conservative 1% unburned carbon loss and an efficiency of 30.1% with negligible unburned carbon loss.

Calculations were made using the computer program, EMISS, to determine the furnace waterwall temperature and required material and tube wall thickness for the various cases run. For the air-fired reference case, the waterwalls are carbon steel with a 0.285” wall thickness. For the maximum temperature O_2-fired case, the maximum wall temperature is 955ºF for which 0.24” thick T91 material is required. Furthermore, due to the greater temperature and greater concentrations of radiating gas species, the required waterwall surface area is only 44% of the air-fired reference case.

The efficiency of carbon sequestration in oxygen-firing boilers even can rival competing gasification plants**.** The power consumption of CO_2 removal for O_2-fired PC plants is about one-third of natural gas combined cycles, about one-half of post CO_2 removal PCs and slightly less than integrated gasification combined cycles. The reduction in power plant efficiency of CO_2 removal for O_2-fired PC plants is nearly half of either natural gas combined cycles or post CO_2 removal PCs and nearly the same as integrated gasification combined cycles. And note that the O_2-firing PC is the only technology that removes 100% of the CO_2.

The oxygen-based PC boiler incorporates cryogenic O_2 separation which can produce very pure oxygen; but it requires substantial capital and operating costs. Membrane separation of O_2 has been demonstrated at small scale employing very thin membrane fibers which preferentially allow O_2 to permeate, but not N_2. Although the purity of O_2 from membranes may not be as high as in the cryogenic separation, lower purity oxygen may be sufficient for the oxygen-based PC boiler power cycle. Membrane separation has the potential to use less power at a lower capital cost.

3.0 Experimental

This work performed for this report was performed utilizing computer program simulations. No experimental equipment was used.

4.0 Results and Discussion

4.1 Reference Site and Fuel Conditions

In December, 2000 Parsons published a study of the cost of electricity of several case studies in CO_2 sequestration from a PC boiler by post capture (Ref. 1). To provide a consistent comparison with the cases analyzed in the Parsons report, the same site conditions (59ºF, 14.7 psia, 60% RH) and the same fuel (Illinois #6) were used. Site Conditions and fuel properties are presented in Figure 1. Fuel HHV and LHV were estimated by a DuLong's method and the stoichiometric air ratio of 867 lb_{air}/lb_{coal} was calculated based on the fuel ultimate analysis.

The liquid CO_2 produced from the oxygen-based PC power plant is not chemically pure, but can readily sequestered in geologic formations (depleted oil and gas reservoirs, unmineable coal seams, saline formations, and shale formations) or in oceans. The liquid CO_2 exits the plant at 2000 psia (Parsons study used 1200 psia). The other gases in the delivered CO_2 are limited to H_2O < 50 ppm (to avoid acid corrosion), and $Ar+N_2$ < 3% (to avoid phase separation). The excess gases in CO_2 stream either have to be purged or recycled. However, since SO_2, as an acid gas, similar to CO_2, it can be sent to pipeline directly, and as mentioned in literature, it does not need to be separated out from CO_2 product. Furthermore it is also not necessary to remove the small concentration of NOx in the CO_2 effluent since it can be sequestered along with the CO_2.

4.2 Air-Fired Reference Case

To study the effects of CO_2 removal on the performance of power plant, an air-fired PC boiler was been simulated in detail as a reference case. This model was used as the base, which was then extended to include the air separation unit (ASU) and CO_2 compression for O_2-fired PC cases.

The reference plant employs a subcritical pressure, natural circulation boiler firing high-volatile bituminous coal producing 2400 psig steam at 1000ºF and reheat steam at 1000ºF to generate 460 MWe. A condenser pressure of 2.5" Hg was applied along with seven feedwater heaters, which raise the feedwater temperature to 494ºF.

Case 1 is the reference air-fired PC boiler case and the model and results shown in Figure 2.

The Aspen Plus model includes coal mills, flue gas heater, pulverized coal-fired furnace, steam generator, superheater, reheater, economizer, ash-removal unit, nitrogen oxides (NOx) selective catalytic reactor (SCR), flue gas de-sulfurization reactor (FGD), air blower, induced draft (ID) fan, feed water pump, cooling water pump, feed water heaters, and a single reheat steam turbine. A coal drying function has been modeled and added into mill module to result in the proper correct mill exit gas temperature. The furnace was simulated by a zero dimensional model. However, all key tube banks of the heat recovery area (HRA) were individually modeled. The furnace roof heat absorption is also simulated. The high pressure steam temperature is controlled by water spray for de-superheat. The simulation also included heat losses from boiler and HRA sides, as well as steam pipes. Some user-defined models are included to perform emission calculations. User built-in calculations have been added to determine boiler efficiency, system net efficiency, and net power. The heat carried by exhaust streams was automatically calculated by program.

For a given steam turbine output and a fuel, ASPEN PLUS iterates for to determine the feed rates of air, coal, etc., based on specified temperature approaches and excess air requirement.

The system configuration, detailed setup parameters and summary of results for the case 1 reference case are shown in Figure 2 and Figure 3. This system has a steam turbine cycle efficiency (generator power divided by heat transferred to the steam cycle) of 45.78%, a boiler efficiency (heat to steam cycle divided by heat input from fuel to boiler) of 88.2%, an unburned carbon loss (UBC) of 1.0%, and a net plant efficiency of 36.68% (net plant heat rate of 9302 Btu/kwh). It has a gross power as 460 MW at generator, an auxiliary power of 42 MW, and a net power of 418 MW. Total heat input from fuel is 3890 MM Btu/hr.

The temperature of the flue gas exhausted to the stack is 292ºF. The flue gas exiting the boiler contains 3.0%, vol., wet O_2 (18% excess air) and contains 773 klb/hr (1.95 lb/kwh) of CO_2.This 3.0% O_2 level is kept constant for all of the O_2-fired cases. An SCR is applied to control NOx with NH_3/NOx=1.0, while an FGD is used to control SOx by lime solution with Ca/S=1.05, L/G=10, and 85% excess air for aeration (Figure 3).

The break down of auxiliary power for case 1 is listed in Figure 4. Most of these power consumptions were simulated directly by the ASPEN module. Some required user Fortran for those processes lacking ASPEN modules, such as solids handling. The power consumption was based on stream flows and design data. Fan power consumption was simulated based on the pressure drops from both air side and gas side. The total auxiliary power consumption, including FGD, for case 1 is approximately 9.2% of the gross power.

4.3 Oxygen-Based PC Plant

4.3.1 Boiler Plant Modifications

The oxygen-based (or oxygen-fired) plant model contains essentially all the components in the conventional plant model. In addition, it also includes an air separation unit (ASU) and a flue gas cooler. In the O_2-fired plant, the FGD is not needed because the SO_2 is acid gas similar to CO_2 and can thus be sent to pipeline together with the CO_2. A substantial portion of the SOx and H_2S will be removed as the flue gas is cooled down in the CO_2 cooling and compression equipment.

The steam side components remain very similar to the air-fired case with only some changes in heat bundle duties in the heat recover area (HRA).

In O_2-fired cases, flue gas is recycled is to control the flame temperature inside the PC-fired boiler to minimize NOx formation, minimize ash slagging in the furnace combustion zone, and avoid the application of exotic materials.

Before the flue gas is separated into a recycled and effluent stream (to the pipeline), it is cooled to 90ºF. Since this is below the acid/moisture dew point a heat exchanger containing acid-resistant materials must be used. The recycled gas is then reheated, before the forced draft (FD) fan, by mixing it with a bypassed hot gas to avoid reaching the dew point. After the O_2 from ASU plant is mixed with recycled flue gas, it is heated by the flue gas exiting the boiler in a gas-gas heat exchanger, which acts as a recuperator to improve cycle efficiency and reduce fan power requirements.

It is assumed in this study that there is no tramp air ingress through the sealed boiler.

4.3.2 Air Separation Unit

For an O_2-fired PC, O_2 purity is a key parameter for system performance and economics. A high purity O_2 will produce a high purity of product CO_2 gas which will reduce CO_2 purification and compression power. However, producing high purity O_2 requires high ASU plant operational and equipment costs. Furthermore, too high O_2 purity is not necessary because the fuel combustion itself will generate some gases, such as N_2 and some excess O_2 is required for complete combustion, in additional to CO_2 as flue gas. Therefore there is a balance point to give an optimum. After literature review and some trade studies, it was determined to use a complete ASU to separate air to O_2 and N_2/Ar, (Ar is separated out from the O_2 by additional column) to maintain an O_2 purity between 99.0 and 99.5%.

The method of air separation chosen for this study is the commercially available large-scale cryogenic air separation technique. A traditional cryogenic ASU plant, shown in Figure 5, was simplified in the simulation to include the power consumption, but without details of distillation columns and cold heat exchangers. The ASPEN simplified ASU model is shown in Figure 6. The ASPEN model does not include the air purifier, which removes moisture, hydrocarbons, CO_2, and NOx in an adsorber and is located between the cold box and air compressor. Although the separated N_2/Ar gases could potentially be sold as byproducts, no economic credit for this is taken in this study. No heat recovery from the ASU air compressor inter-stage coolers is included, because recovery of this low grade heat recovery is very inefficient.

Power consumption is 23.6 kw/klb_{air} from a Ref. 1 for a 95% O_2 purity under 67 psia ASU pressure. From Ref. 2, the power increases by 4% when O_2 purity is increased from 95% to 99.5%. Thus, for the O_2 purity of 99.5% used in this study a power consumption of 24.5 kw/klb_{air} was applied. For a 460 MW steam turbine generation, the ASU plant consumes about 70 MW, or 15% of generated power.

4.4 Parametric Cases

There are five O_2-fired cases for parametric studies as follows:

1. Case 1: air fired reference case
2. Case 2: with ASU & gas recycle, the same mass flows of air and flue gas as case 1
3. Case 3: with ASU, reduced air, the same flue gas flow and O_2% as case 1
4. Case 4: reduced recycle gas flow, the same flame temperature and O_2% as case 1
5. Case 5: reduced recycle gas flow, high flame temperature and the same O_2% as case 1
6. Case 6: reduced recycle gas flow, higher flame temperature and the same O_2% as case 1

Case 2 has the same system net excess air (exit plant O_2 flow rate divided by stoichiometric O_2 flow rate) as case 1, but because of gas recycle, more O_2 is carried by recycled gas back into boiler, which raised boiler excess air (boiler exit O_2 flow rate divided by stoichiometric O_2 flow rate) to a very high number of 69%, and a very high oxygen content of flue gas of 15.3%. The cycle diagram for case 2 is shown in Figure 7.

In Case 3 the air flow rate was reduced to produce a 3%, vol. O_2 concentration at the boiler exit (similar to case 1). This corresponds to a boiler excess of 13.5% and a net excess of 3.1%. Compared to case 2 the air flow rate of case 3 was reduced by 13% from 3422 klb/hr to 2981 klb/hr. O_2 concentration in the boiler is 26.9%, compared to 20.4% for the air-fired case, and yields a higher combustion efficiency. Flue gas flow to the sequestration plant is reduced from 927 klb/hr in case 2 to 820 klb/hr in case 3, which results in less CO_2 compression duty. Case 3 maintains the same flue gas flow rate as cases 1 and 2. The cycle diagram for case 3 is shown in Figure 8.

In cases 4 to 6 the amount of flue gas recycle was reduced to increase the O_2 level in the boiler. This increase in boiler O_2 creates a higher flame temperature, which reduces the size of the furnace and increases the overall cycle efficiency. The effect of flame temperature on cycle efficiency is shown in Figure 13 and Figure 14. Case 4 has the same flame temperature as air-fired reference case, while cases 5 and 6 have higher flame temperatures. Although there is little change in gas exhaust flow to CO_2 compressor among cases 3-6 (listed as gas exit system in Figure 13), the decreasing recycle gas flow results in reduced auxiliary power consumption for both the FD and ID fans. The reduction in FD and ID fan auxiliary power requirements is presented in Figure 15 (approximately 2.5 MW from case 3 to case 6) along with the reduction in ASU power requirements (approximately 1 MW from case 3 to case 6)

Figure 16 shows the relationship of flue gas flow to boiler flame temperature. Figure 17 is a similar plot to Figure 16, but it uses volumetric flow rate as the abscissa instead of mass flow rate. Both figures show the air-fired data for comparison. It can be observed that the O_2-fired PC has a lower volume flow rate than does the air-fired PC due to the higher molecular weight of the flue gas (i.e. CO_2 versus N_2). From case 3 to case 6 the ratio of the O_2-fired PC volume flow rate to the air-fired PC volume flow rate drops from 75% to 57%, which means for a constant flue gas velocity, boiler size is reduced.

Another advantage from decreasing the quantity of recycle gas is to increase the O_2 content in the boiler (from 27%v to 34% by vol. from case 3 to case 6) which should improve the fuel combustion and reduce the required height of the furnace. This credit has not been simulated in this system study, but will be modeled in the 3-D CFD boiler simulation study (Task 4).

4.5 Hot Recycle

A case with hot gas recycle (case 7, Figure 12) was run to evaluate its effect on the system performance. A hot gas recycle will bring more energy back into boiler, will reduce fuel and O_2 feed rates, and reduce ASU duty, but it requires more power to the fan because of the increased recycle gas volume flow.

Figure 13 shows that increasing the recycle gas temperature from 95ºF to 148ºF increases the boiler efficiency from 89.60% to 90.94% and net efficiency from 31.90% to 32.12%. The additional fan power increase of 0.3 MW was more than made up by a reduction in ASU power of 0.5 MW. The resultant fuel saving is approximately 0.7%. The size of the gas-gas heat exchanger increases due to the reduction in LMTD (fluid temperature difference is reduced from 213ºF to 176ºF for the hot end, and from 104ºF to 74ºF for the cold end).

There is a limit to increasing the recycle gas temperature without increasing the stack gas temperature, which will reduce efficiency and increase cooling duty. One option mentioned in literature is to raise both stack gas and recycle gas temperatures, and then recover heat from stack gas to replace part of the feedwater heaters. The merits of this approach are questionable because the efficiency improvement will be very small from replacing the low pressure feedwater heater.

Figure 18 shows the cooling curve for flue gas cooling before compressor. Part of the heat can be recovered to preheat the condensate. If 90 MMBtu/hr of heat is recovered (exit temperature = 142ºF), the steam saved from extraction will generate an additional 2.2 MW, or an efficiency increase of 0.2% point. If 150 MMBtu/hr of heat is recovered (exit temperature = 133ºF), the steam saved from extraction will generate an additional 3.4 MW, or an efficiency increase of 0.3% point. This potential energy savings will be further explored in Section 4.8.

4.6 CO_2 Compression

The flue gas effluent stream (mainly CO_2) has to be compressed to the high pipeline pressure of 1200 to 2000 psia. Using case 3 as a basis, case 8 is simulated in which the CO_2 sequestration equipment is added to the system and the effluent is conservatively compressed to 2000 psia. The dominant moisture in flue gas is condensed out first during flue gas cooling before the first stage compression. The condensed water contains acid gases and has to be treated before recycle or discharge.

A flue gas dry composition before the first stage CO_2 compressor from case 8 is:

CO_2	O_2	N_2+Ar	SOx	H_2O
90.9	2.9	1.3	1.3	3.6

In literature, the O_2 as low as 1.3% was used. Reducing O_2 content, such as from 3.0% to 2.0% by reducing excess air, would be helpful in reducing CO_2 compression power, but it is judged that and oxygen content of approximately 3.0% is required for good combustion efficiency. Both CO_2 and SO_X are acid gases. They combine with moisture to form acid, which causes a corrosion problem along CO_2 pipeline. Therefore, after the 2nd stage, a chemical method of active dehydration with TEG (Triethyleneglycol), regarding hydrate formation and corrosion, has been applied to remove the rest of moisture out to a very low level as less than 50 ppm, where the TEG can be regenerated by heating. In the model, the TEG dehydration was simulated, but the TEG itself was not simulated.

A four-stage compression with inter-stage cooling was applied in case 8. To reduce power, an equal compression pressure ratio of approximately 3.4 was applied. The results are shown in Figure 19 and Figure 23. The effects of the addition of the ASU and the CO_2 sequestration plant can be seen in Figure 23 as follows:

Efficiency = 36.7% with air (case 1)
Efficiency = 31.2% with ASU (case 3)
Efficiency = 27.9% with ASU & CO_2 sequestration (case 8)

4.7 Vent Gas Recycle

Because of limitation of the other gases in CO_2 pipeline (i.e., H_2O < 50 ppm and $Ar+N_2$ < 3%), the excess gases have to be purged. A novel idea applied in this study is to recycle this O_2-rich purge gas back to system and to reduce ASU duty and to recover power. The vent gas recycle has been simulated in case 9 (Figure 20). To effectively separate out the non-CO_2 gases, the compressed CO_2 stream from the 3rd stage is cooled down to its dew point (Figure 24) and condensed out by phase separation by a strip tower (not simulated in this study). Instead of using a compressor for the 4th stage CO_2 of compression as in case 8, a CO_2 liquid pump is used in case 9 to save power and cost.

The composition of the purge gas is as follows:

	O_2	N_2+Ar	others
Case 9	71.6	28.1	0.3

Although the flue gas flow discharged to CO_2 plant is nearly identical for both cases 8 and 9 as shown in Figure 23, the net excess O_2 reduces to near zero when vent gas recycle is applied, and consequently the air to system is reduced from 2981 to 2905 klb/hr. However since the boiler is still operated at 27.5% O_2 inlet, and a 3% O_2 outlet the boiler combustion performance will not be affected.

The addition of vent gas recycle (compare case 8 to case 9 in Figure 23) increases the net efficiency from 27.9% to 28.3%, increases the net power from 317.3 to 321.8 MW, and decreases the auxiliary power from 143.1 to 138.6 MW (a net 4.5 MW saving). Gas recycle will also reduce NOx in the effluent since the majority of the NOx will be separated out with the purge gas and removed by the gas adsorber in ASU plant.

Since the vent gas from the CO_2 compression plant is at 920 psia and the ASU is at 67 psia, the vent gas can be sent to an expander to generate power and reduce its pressure and temperature. This power generation is not included in case 9, but is added in a subsequent case (see Section 4.8). Furthermore, the low temperature expanded recycled gas could be used as extra coolant for the ASU distillation column. However, this energy savings has not been accounted at present time because the model does not include the details for ASU plant.

In cases 10 and 11 the amount of flue gas recycled was reduced to increase the O_2 level in the boiler (similar to cases 4 to 6). This increase in boiler O_2 creates a higher flame temperature, which reduces the size of the furnace and increases the overall cycle efficiency. The effect of flame temperature on cycle efficiency is shown in Figure 23, which demonstrates that raising the flame temperature from 3575ºF to 4343ºF increases the net efficiency from 28.3% to 29.5% (boiler efficiency is also increased from 88.3% to 90.8%). Cycle diagrams for cases 10 and 11 are shown in Figure 21 and Figure 22, respectively.

4.8 Design Case

Based on the results of the parametric cases, case 12 (design case) is modeled

Since all processes that capture CO_2 must include some kind of wet gas cooler, due to the low temperature to which the flue gas is cooled, case 12 utilizes a wet-end economizer to recover as much of this heat as practical. In case 12 a heat exchanger is added to recover the flue gas sensible energy and prior to separation into recycle and outlet streams. The heat exchanger cools the flue gas to 142ºF and removes 90 MM Btu/hr. This significantly increases the boiler efficiency by reducing the energy content of the flue gas effluent. The LMTD of the heat exchanger is 34ºF and it is judged that reducing the flue gas temperature further (below 142ºF) could be uneconomical since the required heat exchanger would be too large. Figure 27 shows that reducing the flue gas temperature from 142ºF to 139ºF (increasing the heat exchanger absorption from 90 MM Btu/hr to 100 MM Btu/hr) reduces the LMTD from 34ºF to 26ºF.

Although the flue gas outlet temperature of this additional heat exchanger is nominally at the moisture condensation temperature, the presence of SOx in the flue gas raises the dew point of gas such that sulfuric acid solutions will condense below 300ºF. Consequently the heat exchanger must be constructed from acid-resistant materials.

There are two possible applications for this wet-end economizer: 1) replace the first stage FWH1 or 2) heat up a split stream from the condensate pump. The second option results in a higher thermal efficiency since it reduces the higher pressure steam extractions. This is demonstrated in Figure 28 which shows that the efficiency (work/steam thermal energy) of the steam, if it is not extracted, increases with increasing pressure (stage). Therefore the split stream method yields higher efficiency and is applied in the design case (Figure 26).

Case 12 also incorporates a turbine expander to recover power from the recycled purge gas pressure reduction. The turbine generates 0.41 MW of power as the pressure is reduced from 925 psia to 67 psia. The expanded gas is quite cold (-183ºF) and could potentially be used as a coolant for distillation operation. However, since the ASU was not modeled in detail in this study no credit for this cooling effect is taken.

Figure 26 shows the cycle diagram for case 12. The boiler efficiency is 95.3% and the overall cycle efficiency is 29.8% (Figure 23). The 4.7% losses in the boiler efficiency is comprised of 17% sensible heat, 66% latent heat (H_2O), 14% unburned carbon, 3% radiation. Since it is likely that with O_2-fired combustion and a high flame temperature, the unburned carbon loss will be nearly zero (to be confirmed in Task 4), the boiler efficiency would be 94.1% and the overall cycle efficiency would be 30.1%.

4.9 Furnace Waterwall Temperature

The level of radiation in the O_2-fired boiler is significantly higher than an air-fired boiler due to greater concentrations of radiating gas species (CO_2 and H_2O) and higher flame temperature. Consequently, it is important to select the proper amount of recycled flue gas to limit the water wall temperature such that a reasonable waterwall material can be used.

The maximum waterwall temperature and furnace heat flux was calculated from the ASPEN results using the Foster Wheeler computer program, EMISS. The EMISS computer program calculates radiative heat flux of CO_2 and H_2O gases as follows:

$$Q/A = \left(\frac{\varepsilon_{wall}+1}{2}\right)\sigma(\varepsilon_g T_g^4 - \alpha_g T_w^4)$$

$$Q/A = U_O(T_W - T_f)$$

where,

ε	$= \varepsilon_{CO2} + \varepsilon_{H2O} - \Delta\varepsilon$
$\Delta\varepsilon$	= Correction factor due to spectral overlap
ε_{CO2}	= Emissivity of CO_2 (function of temperature, mean beam length, and partial pressure of CO_2)
ε_{H2O}	= Emissivity of H_2O (function of temperature, mean beam length, and partial pressure of H_2O)
ε_g	= Emissivity of gas at gas temperature
α_g	= Absorptivity of gas at wall temperature (equal to emissivity)
ε_{wall}	= Tube wall emissivity (assumed to be 0.7)
σ	= Stefan-Boltzmann constant = 1.714×10^{-9} Btu/hr-ft^2-R^4
Q/A	= Heat Flux
T_f	= Water/Steam fluid temperature (ºR)
T_g	= Gas temperature (ºR)
T_w	= Wall temperature (ºR)
U_o	= Heat transfer coefficient from outside of wall to steam/water fluid

Figure 25 presents the calculated furnace heat flux for cases 1 to 12. Both the maximum heat flux (based on the maximum furnace gas temperature) and the average heat flux (based on the average furnace gas temperature) are presented. Based on the maximum heat flux, the maximum water wall temperature is computed. From this maximum wall temperature and the selected material, the tube wall minimum thickness is computed using stress allowables from the ASME Boiler and Pressure Vessel Code.

Figure 25 shows that for the air-fired reference case (case 1), the waterwalls are carbon steel with a 0.285” wall thickness. Even though the O_2-fired cases 2 and 3 have similar furnace gas temperatures than case 1, the substantially greater concentrations of CO_2 and H_2O results in a radiative heat flux of approximately 50% higher. Thus, to maintain the same wall thickness requires an upgrade in the material to the T2 alloy. As the flame temperature is increased in cases 4 to 6 and cases 9 to 11, the heat flux and wall temperature increases requiring further material upgrades. For case 11 (and case 12), the maximum wall temperature is 955ºF for which 0.24” thick T91 material is required.

The ratio of the average furnace heat flux of the O_2-fired furnace to the average furnace heat flux of the air-fired furnace is also presented in Figure 25. For case 12 this ratio is 0.44, which means that case 12 requires only approximately 44% of the case 1 heating surface area. This can be used as a preliminary estimate of the O_2-fired furnace size. Figure 25 shows that the air-fired furnace dimensions of 36’ x 51’ x 207’ (D x W x H) are substantially reduced to 27’ x 38’ x 140’ (D x W x H) in case 12. Task 4 of this study will perform a detailed design of the O_2-fired boiler in detail by performing a three-dimensional CFD simulation.

4.10 Comparison With Post CO_2 Capture

CO_2 cannot be free captured and sequestrated without reducing both the plant power and efficiency because of a potential energy stored in the pressurized liquid CO_2. A minimum of 40 kw/klb_{CO2} additional auxiliary power is required for CO_2 compression. The difference between technologies lies in the difference in power requirements of the different CO_2 or O_2 separation techniques.

Parsons (Ref. 1) performed some studies on CO_2 removal by a post capture method for a conventional PC boiler. The plant efficiency drops from 40.5% to 28.9% for a supercritical (3500 psia/1050ºF/1050ºF/1050ºF/2.0”Hg) boiler, and from 42.7% to 31.0% for an ultra supercritical (5000 psia/1200ºF/1200ºF/1200ºF/2.0”Hg) boiler. In the study presented herein, the CO_2 removal using an O_2-fired PC is used, which relies on an ASU. The efficiency drops from 36.7% (case 1) to 30.1% (case 12 with minimal UBC loss) for a subcritical (2415 psia/1000ºF/1000ºF/2.5”Hg) boiler. Since the Parsons study compressed the effluent CO_2 to 1200 psia, whereas this study herein compressed the CO_2 to 2000 psia, the 1200 psia pressure is used as comparison basis, which increases the case 12 efficiency to 30.2%. The net efficiency drops for these cases are

11.7% points for supercritical, post removal
11.7% points for ultra supercritical, post removal
6.5% point for subcritical, O_2 fired

Based on the Parson’s study it appears that the efficiency reduction is independent of steam cycle. However, there is a big difference between the efficiency reductions for post capture and O_2-fired. Furthermore, the O_2-fired method gives near 100% CO_2 removal, while the post capture is practically limited to about 90% (limited by a vapor-liquid equilibrium from absorption-regeneration cycle).

Another comparison basis is the kw/lb_{CO2} removal, where the kw is power generation difference between cases with and without CO_2 removal. Comparing the post CO_2 capture to the O_2-fired case:

187 kwh/lbCO_2 for supercritical, 90% post removal
188 kwh/lbCO_2 for ultra supercritical, 90%post removal
93 kwh/lbCO_2 for subcritical, O_2 fired 100% removal

Again, the change in power penalty for CO_2 removal appears independent of steam cycle. From above data, it is very clear that the O_2-PC has advantages over the post CO_2 capture.

The efficiency of carbon sequestration in oxygen-firing boilers even can rival competing gasification plants**.** Figure 29 compares the power consumption of adding CO_2 removal equipment to various competing technologies. Figure 29 shows that the power consumption of CO_2 removal for O_2-fired PC plants is about one-third of natural gas combined cycles (NGCC), about one-half of post CO_2 removal PCs and slightly less than integrated gasification combined cycles (IGCC). Figure 30 compares the reduction in power plant efficiency of adding CO_2 removal equipment to various competing technologies. Figure 30 shows that the reduction in power plant efficiency of CO_2 removal for O_2-fired PC plants is nearly half of either natural gas combined cycles (NGCC) or post CO_2 removal PCs and nearly the same as integrated gasification combined cycles (IGCC). And once again note that the O_2-firing PC is the only technology that removes 100% of the CO_2.

Figure 1 – Site Conditions and Coal Properties

Elevation	ft	0	
Ambient pressure	psia	14.70	
Ambient Temperature	F	59.0	
Ambient Temperature, wet	F	51.5	
Relative Humidity	%	60.0	
P-H2O	psia	0.247	
Y-H2O	%, vol	1.010	
Condenser Pressure	" Hg	2.50	
Air Composition		Dry	Wet
N2	%, vol	78.085	77.297
O2	%, vol	20.947	20.735
Ar	%, vol	0.935	0.926
CO2	%, vol	0.033	0.033
H2O	%, vol	0.000	1.010
Total	%, vol	100.000	100.000

Illinois No. 6 Coal		
C	%	63.75%
H	%	4.50%
O	%	6.88%
N	%	1.25%
Cl	%	0.29%
S	%	2.51%
Ash	%	9.70%
H2O	%	11.12%
Total	%	100.00%
LHV	Btu/lb	11,283
HHV	Btu/lb	11,631

Figure 2 – Cycle Analysis of Case 1 (Air-Fired Reference Case)

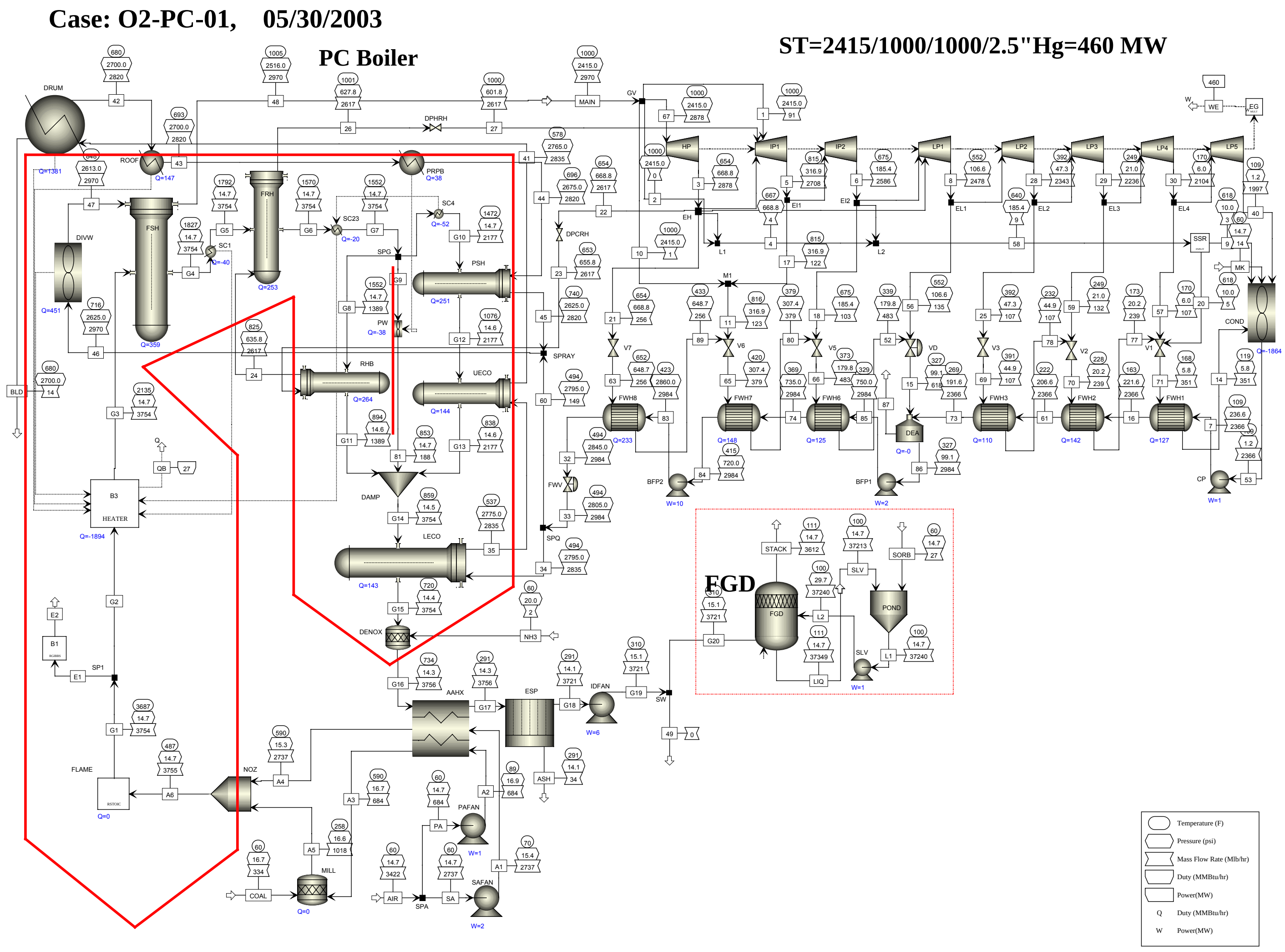


Figure 3 – Case 1 (Air-Fired Reference) Boundary Conditions and Results

SETUPS & RESULTS							
amb		**dP-air**	"H2O	**ST**		**result**	
elevation, ft	0	AAHX	4.0	main P, psia	2415	net power, MWe	418
amb T, F	60	duct	6.0	main T, F	1000	net eff, %	36.68
amb P, psia	14.7	nozzle	10.0	RH P, psia	602	gross @ST, MW	460
RH, %	60	sum	20.0	RH T, F	1000	aux power, MW	42.1
				FWHs	7	as %	9.2
air	%v	**dp-gas**	"H2O	end wet, %	9.9		
O2	20.74	FSH	0.5	end P, "Hg	2.5	HHV in, mmbtu	3890
N2	77.3	FRH	0.5			Q to ST, mmbtu	3431
Ar	0.93	RH	1.0	**FWH**	F	Q, cond, mmbtu	1864
CO2	0.03	PSH	0.7	TD	5	boiler eff, %	88.2
H2O	1.01	UECO	0.3	DC	10	ST cycle eff, %	45.78
sum	100.0	ECO	2.0	FW T	494	Generator eff, %	98.3
		AAHX	1.6				
coal	%w	Damper	4.3	**DeSuperheat**		air, klb	3422
C	63.75	BHG	5.5	SH, %	5	coal, klb	334
H	4.5	FGD	12.0	water T, F	494	sorb, klb	26
O	6.88	sum	28.4			flue gas, klb	3721
N	1.25			**Boiler**		O2, %	3.0
S	2.51	**dP**	"H2O	UBC, %	1.0	H2O, %	8.5
A	9.99	PAFan	60	margin, %	0.5	CO, ppmv	71
M	11.12	IDFan	28	radiation, %	0.22	NOx, ppmv	733
V	34.99	SAFan	20	Exa, %	18	SOx, ppmv	2082
F	44.19			flame T, F	3839	after FGD	42
sum	100.0	**eff**	%	stack T, F	292	Ash, klb	34
		FDFan	75	blowdown, %	0.5	C, %	6.0
fuel HHV	btu/lb	IDFan	70	miller exit T, F	258		
given	11666	CWPump	80			main st flow, klb	2970
aspen	11631	BFPump	80	**FGD & SCR**		RH st flow, klb	2878
		Motor/mechanic	95	L/G	10	end st flow, klb	1997
sorb	%w			Ca/S	1.05		
CaCO3	100	**air**	%	Excess air, %	85	DeSOx, %	98
		PA	20	NH3/NOx	1.0	DeNOx, %	90

Figure 4 – Auxiliary Power Requirements for Case 1

Aux power		MWe
condensed water pump		0.6
LP feed water pump		2.5
HP feed water pump		9.8
circulating water pump		3.9
FGD pump		0.7
PA Fan		1.5
SA Fan		2.0
ID Fan		5.6
FGD Fan		4.1
cooling tower Fan		2.1
coal handling		2.1
sorb handling		0.8
ash handling + ESP		1.8
others (=1%)		4.6
total		42.1

Figure 5 – Air Separation Unit

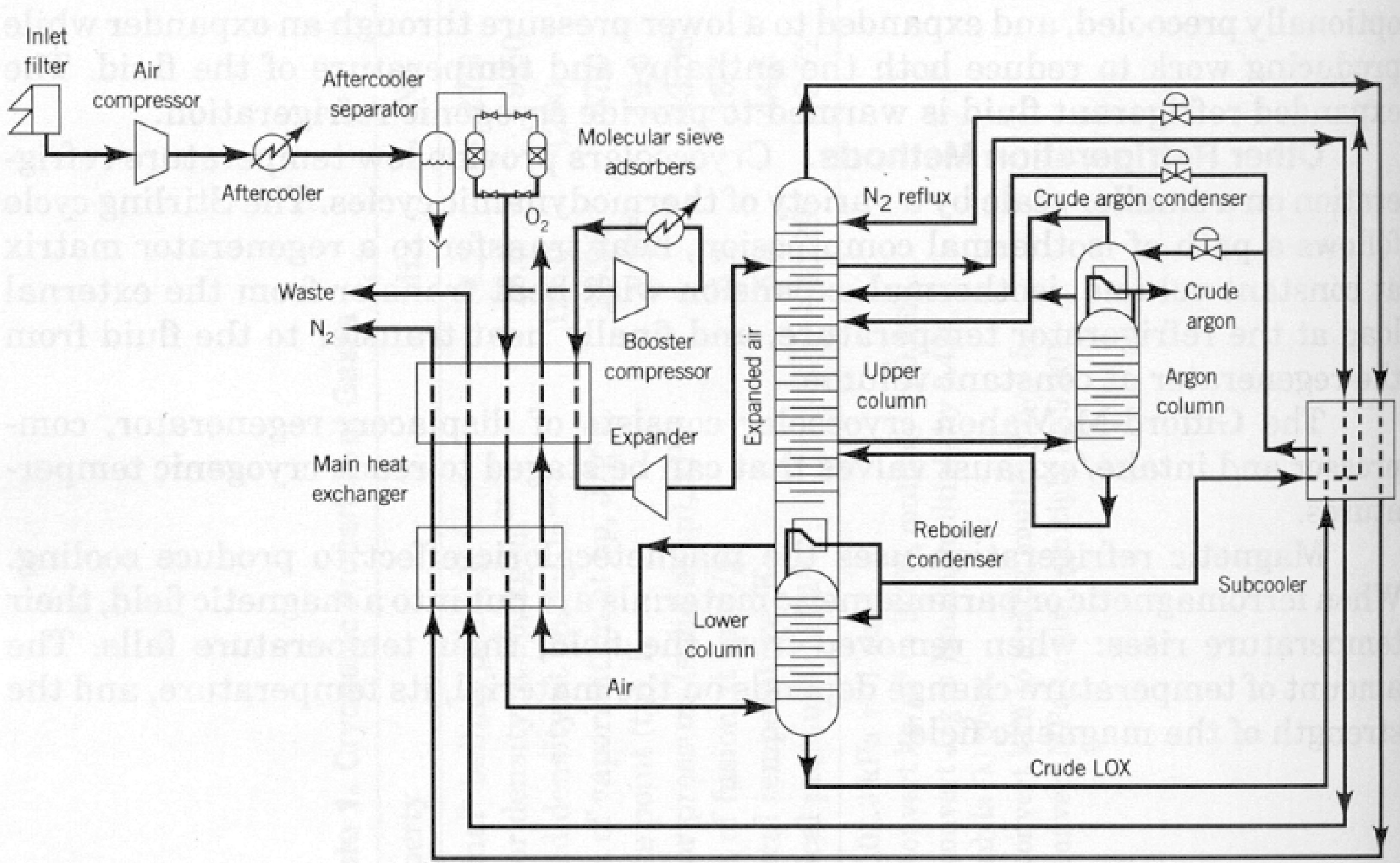


Fig. 1. Cryogenic air separation process. LOX = liquid oxygen.

Figure 6 – ASPEN ASU Model

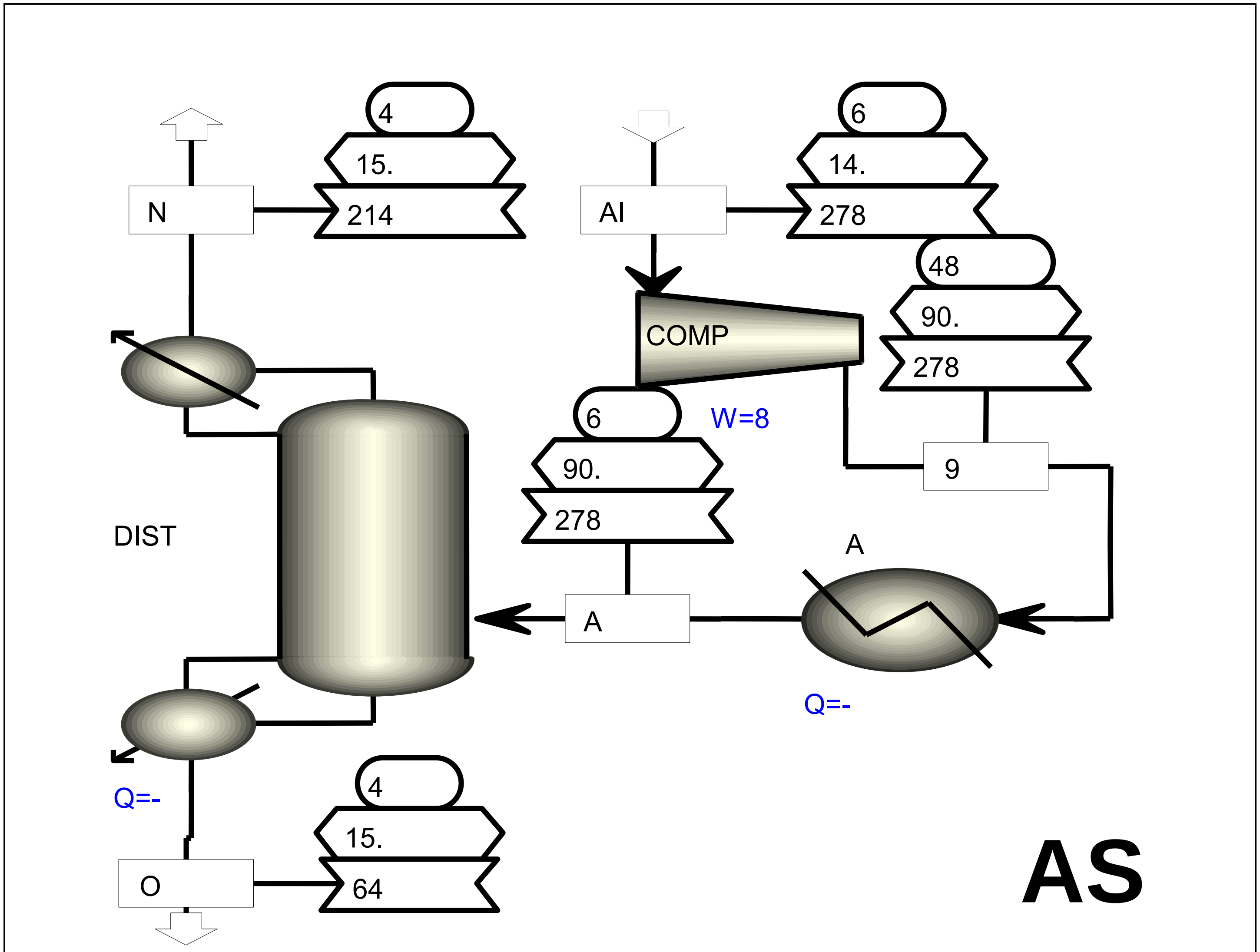


Figure 7 – Case 2 Cycle Diagram

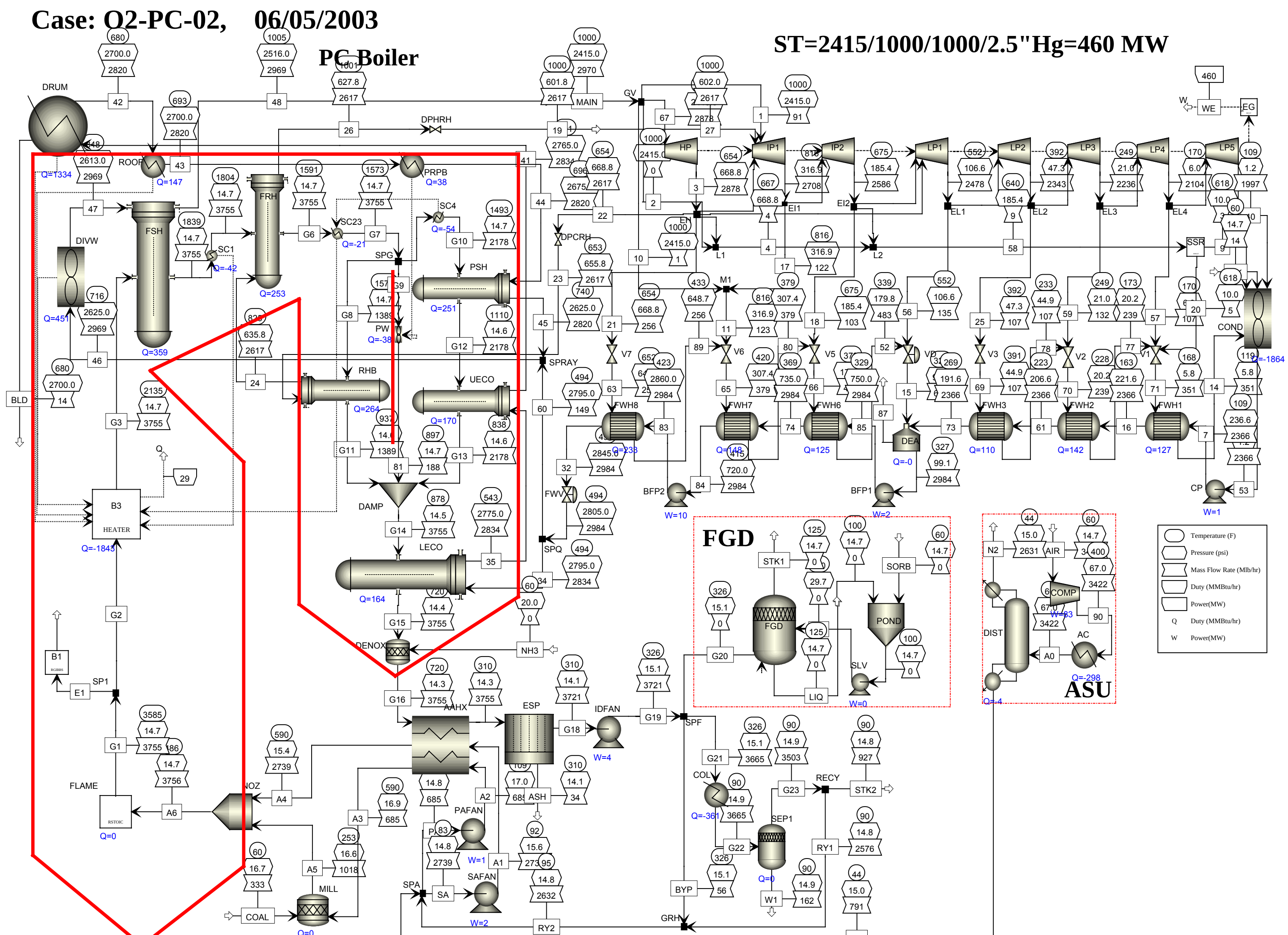


Figure 8 – Case 3 Cycle Diagram

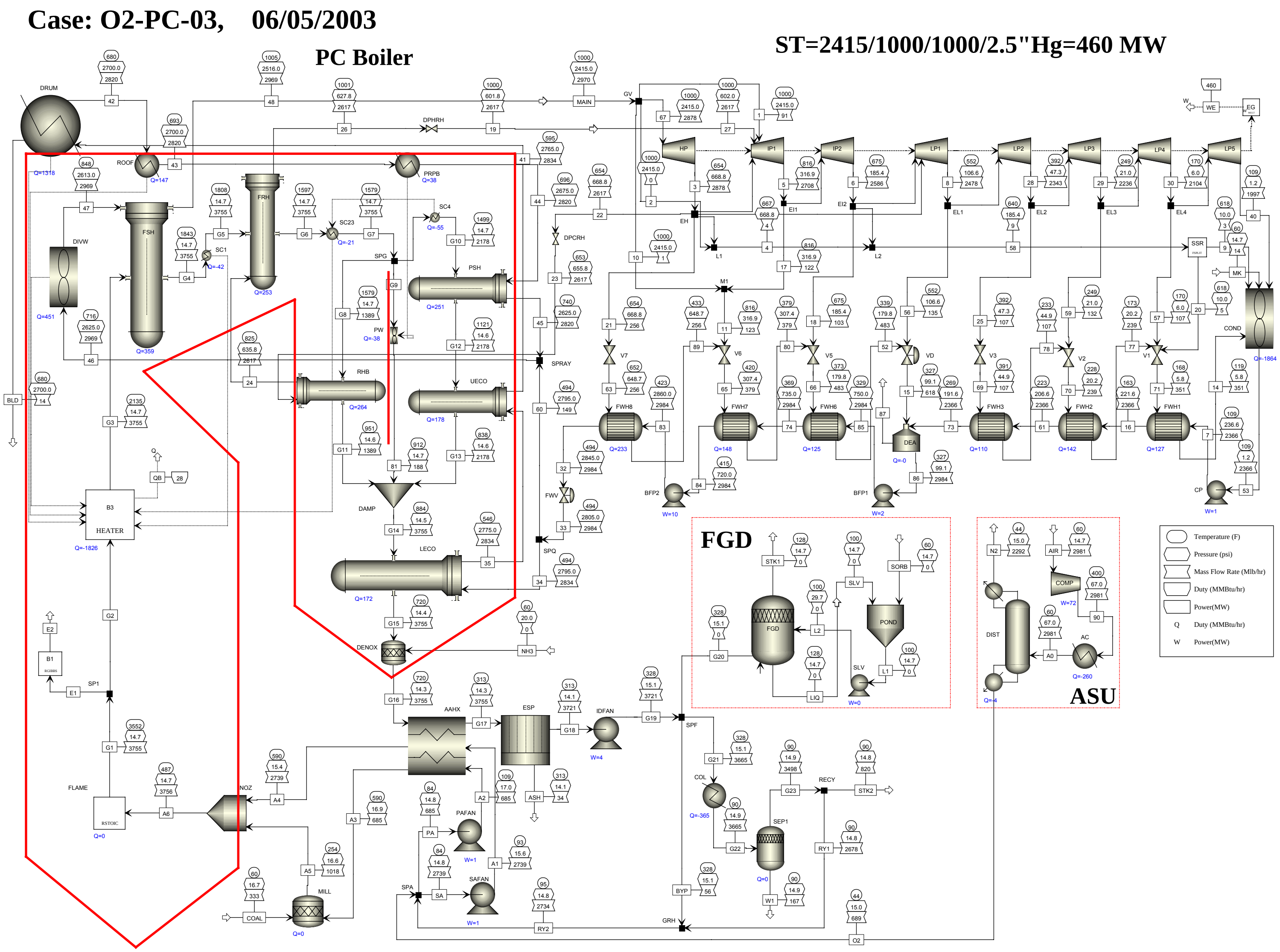


Figure 9 – Case 4 Cycle Diagram

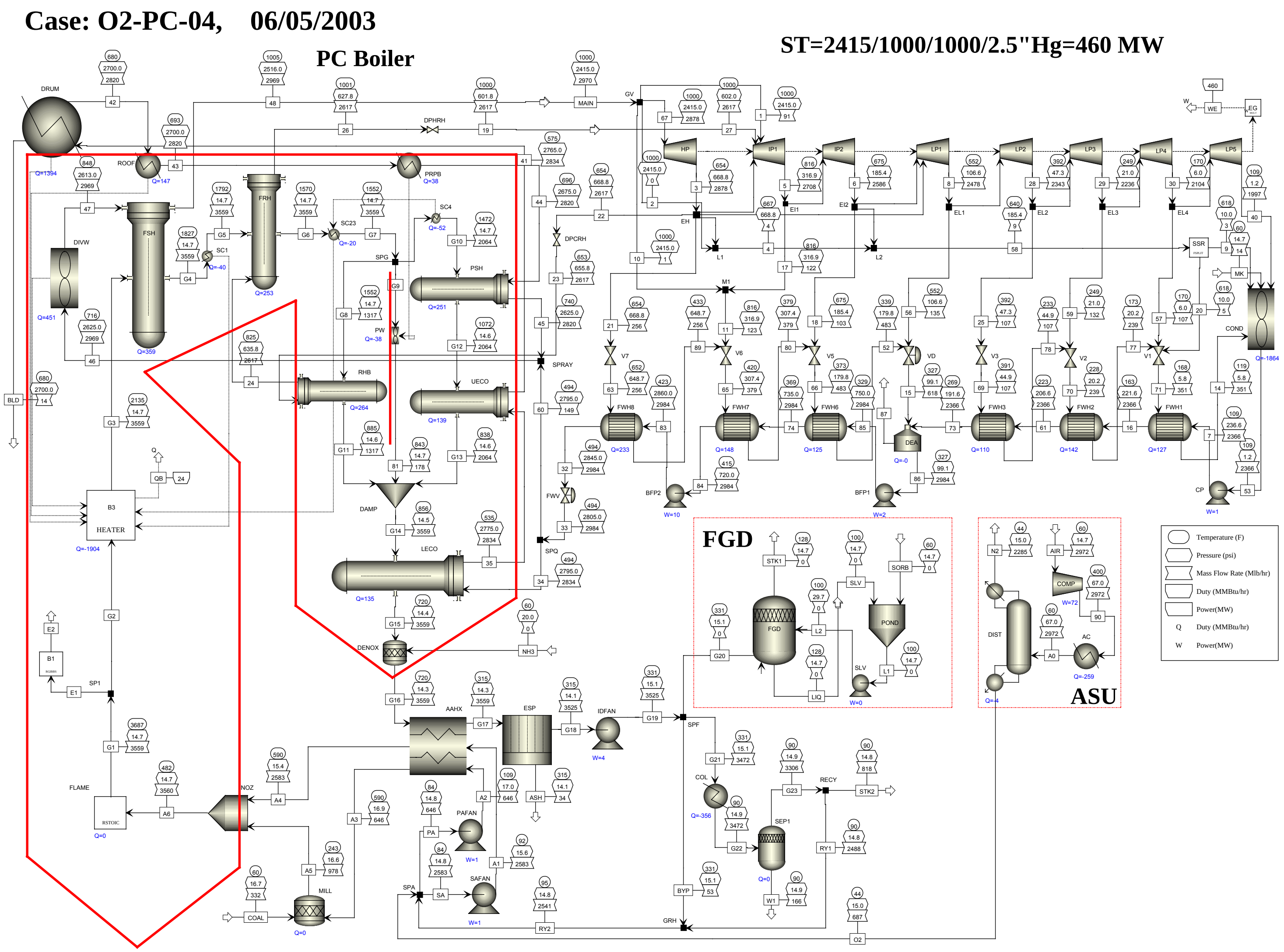


Figure 10 – Case 5 Cycle Diagram

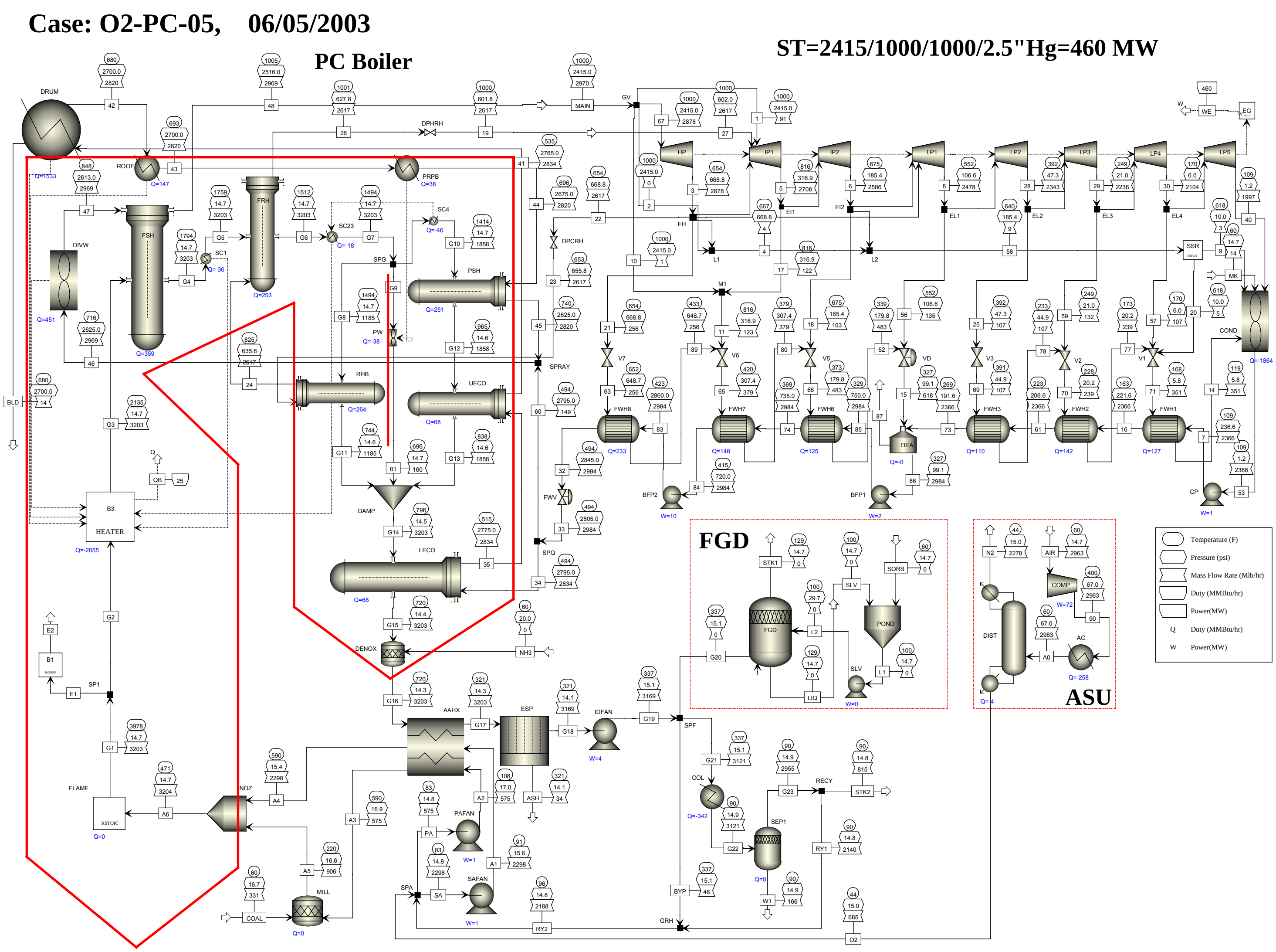


Figure 11 – Case 6 Cycle Diagram

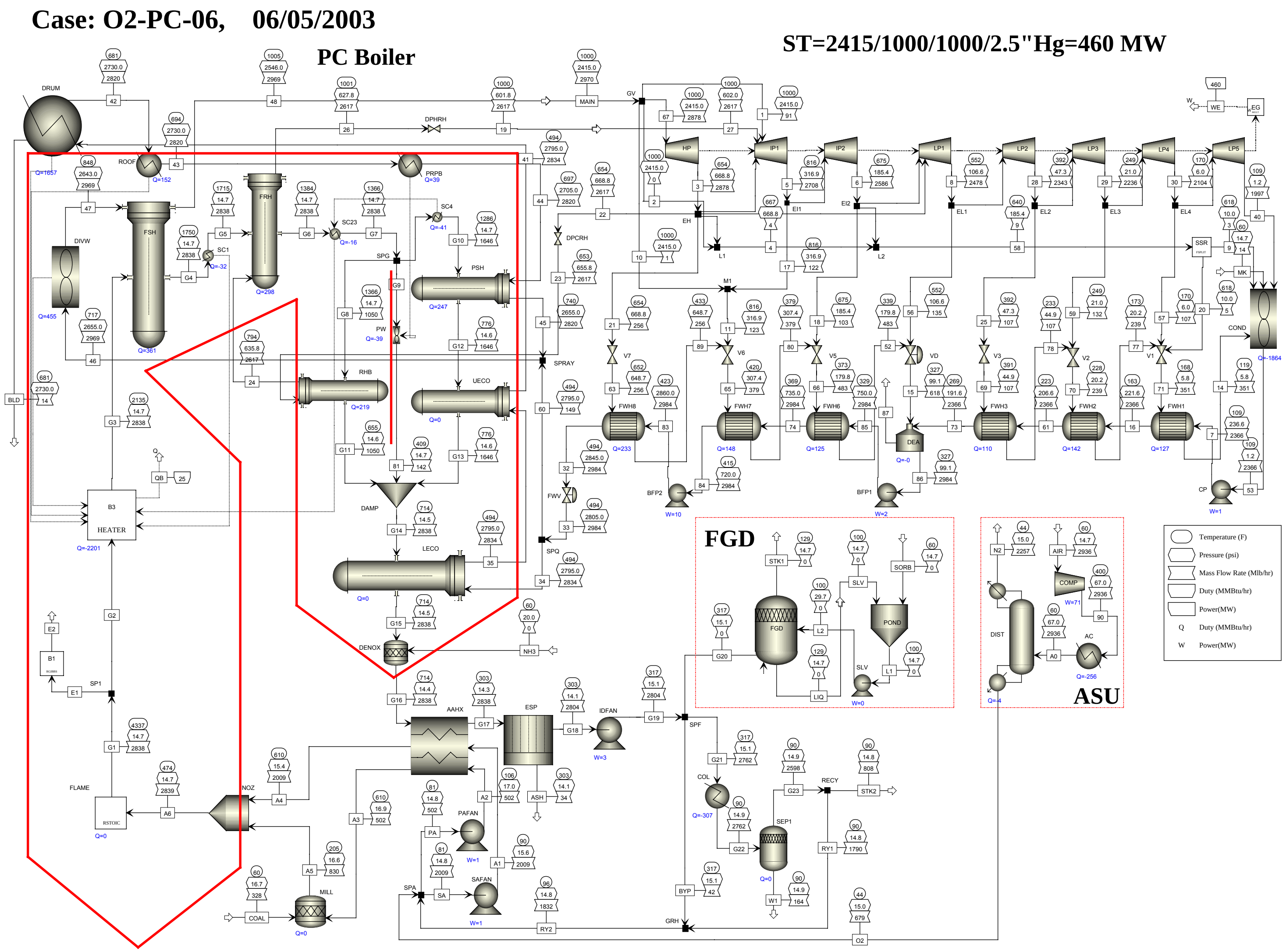


Figure 12 – Case 7 Cycle Diagram

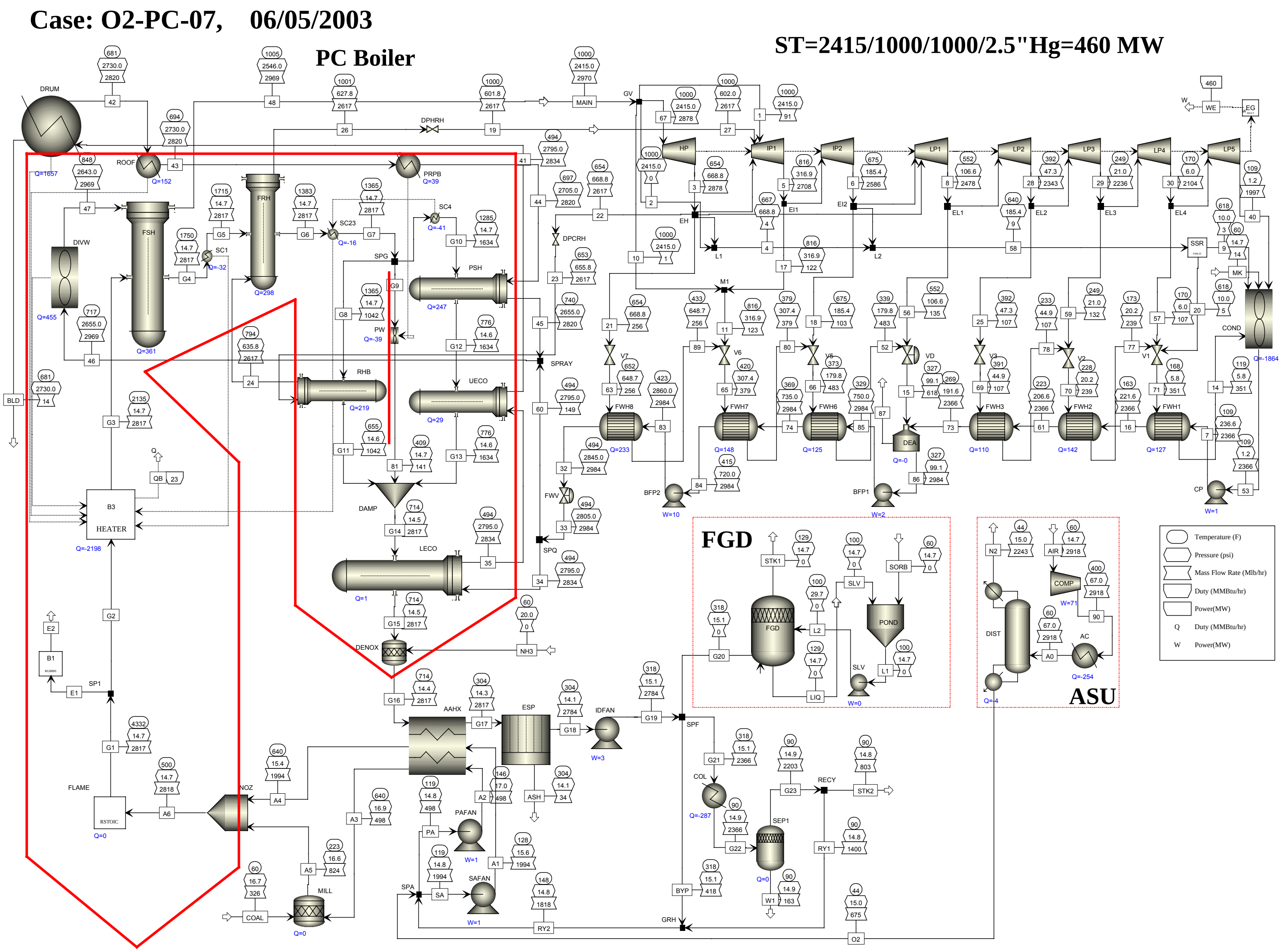


Figure 13 – Parametric Cases Summary (Without CO_2 Compression)

case	1	2	3	4	5	6	7
ASU	no	yes	yes	yes	yes	yes	yes
coal to boiler, klb/h	334	333	333	332	331	328	326
air to plant, klb/h	3422	3422	2981	2972	2963	2936	2918
O2 purity, %	20.7	99.6	99.6	99.6	99.6	99.6	99.6
O2 level to boiler, %	20.4	39.7	26.9	28.1	30.9	34.1	33.7
O2 level exit boiler, %	3.0	15.3	3.0	3.0	3.0	3.0	3.0
Net EXA, %	18.0	18.0	3.1	3.1	3.0	3.2	3.2
Boiler EXA, %	18.0	69.1	13.6	12.5	11.2	10.4	10.4
flame T, F	3687	3608	3574	3687	3978	4337	4332
gas to boiler exit, klb/h	3721	3721	3721	3525	3169	2804	2784
gas to boiler exit, mmcft/h	237	181	175	167	151	134	135
gas recycle to boiler, klb/h	0	2632	2734	2541	2188	1832	1818
gas T before IDFan, F	291	307	309	315	321	303	304
Recycled gas T, F	-	95	95	95	95	96	148
gas exit system, klb/h	3721	927	820	818	815	808	803
CO2, %v	14.0	76.4	89.6	89.6	89.6	89.6	89.6
ASU plant power, MW	0	83.0	72.3	72.1	71.8	71.2	70.7
Aux power, MW	42.1	116.7	105.8	105.3	104.4	102.7	102.5
Boiler eff, %	88.2	88.3	88.3	88.6	88.9	89.60	90.94
Net eff (w/o CO2), %	36.68	30.19	31.15	31.29	31.46	31.90	32.12

Figure 14 – Net Efficiency (without CO_2 compression) Versus Flame Temperature

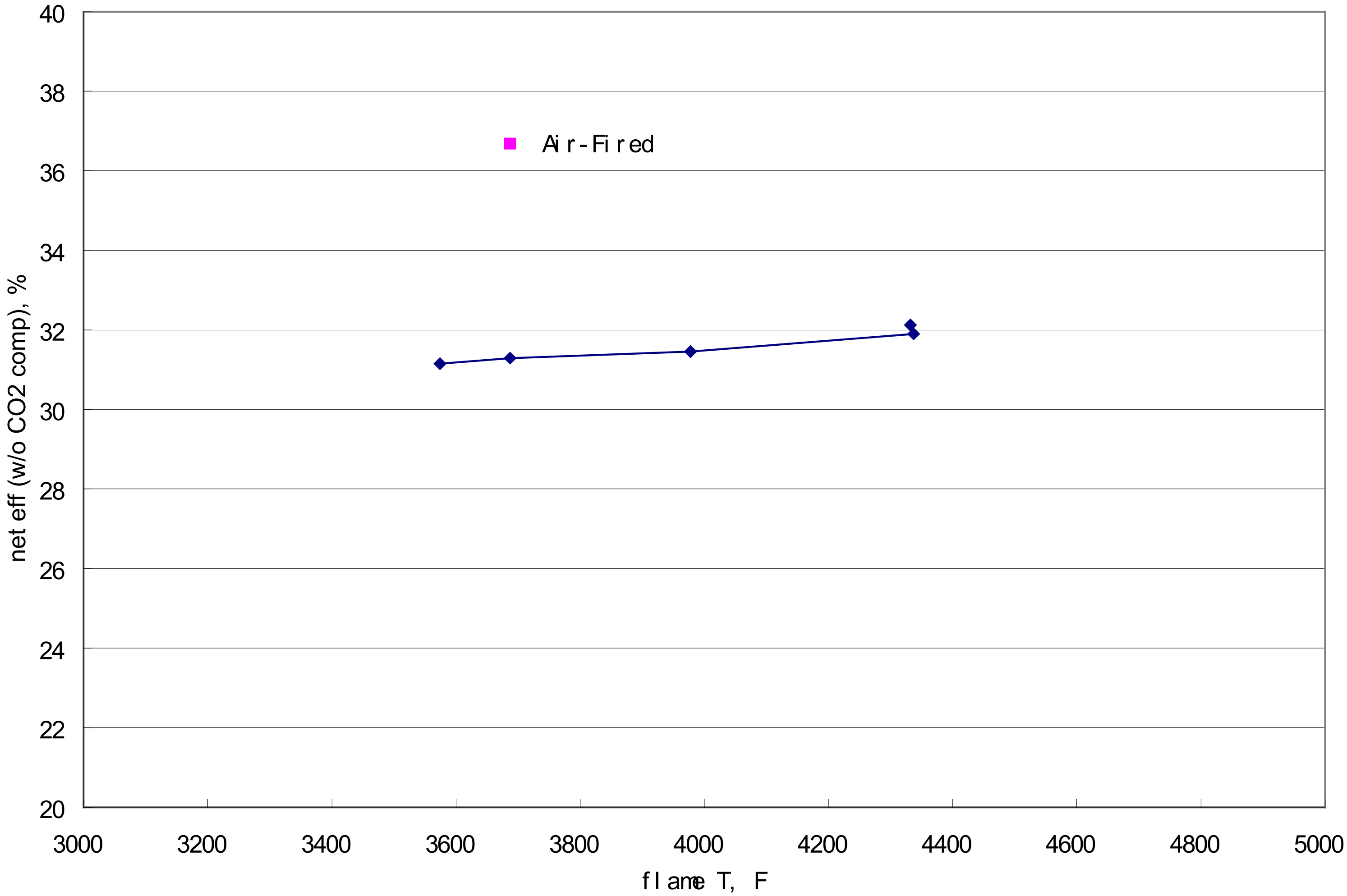


Figure 15 - ASU and Auxiliary Power Requirements Versus Flame Temperature

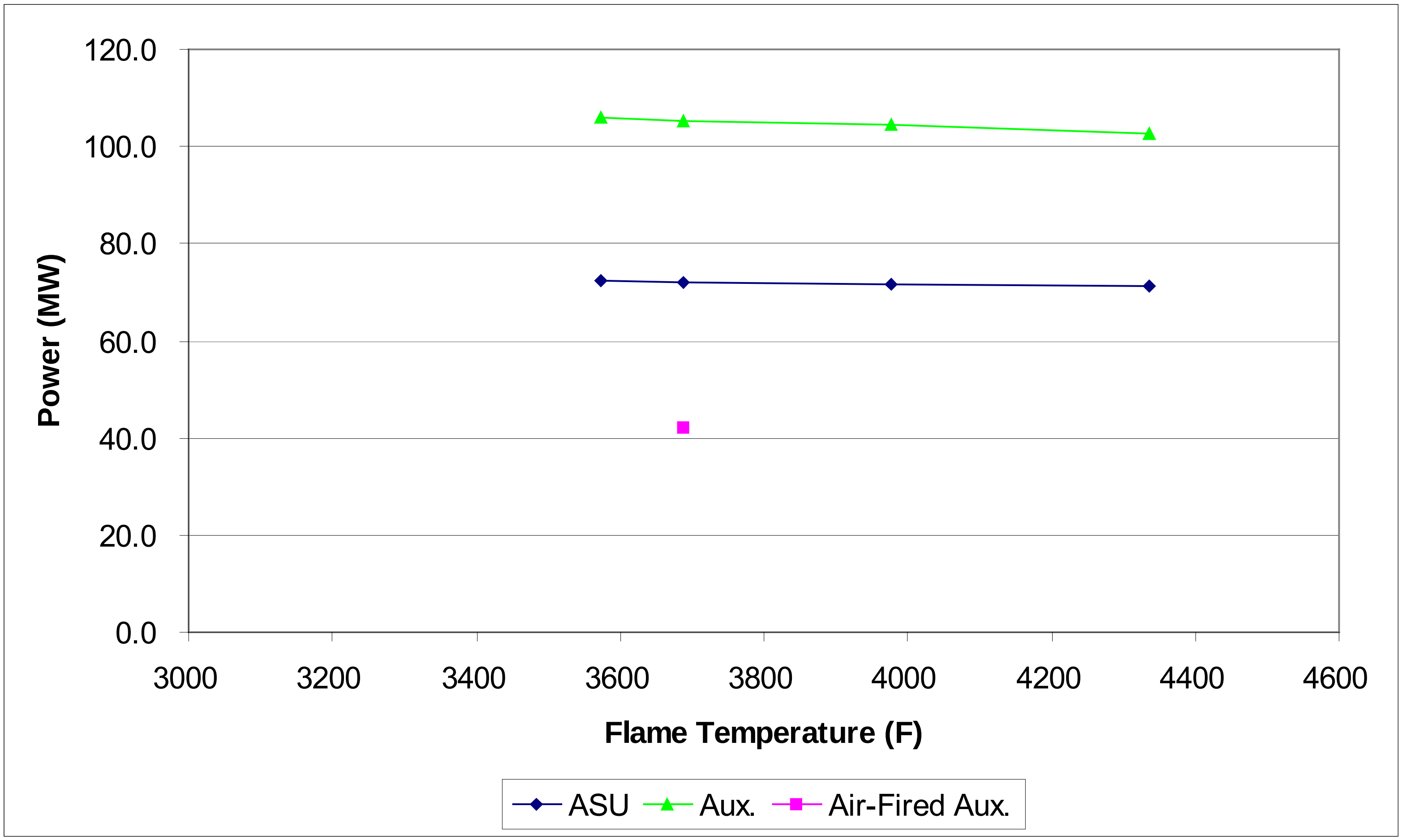


Figure 16 – Recycled Gas Mass Flow Rate Versus Flame Temperature

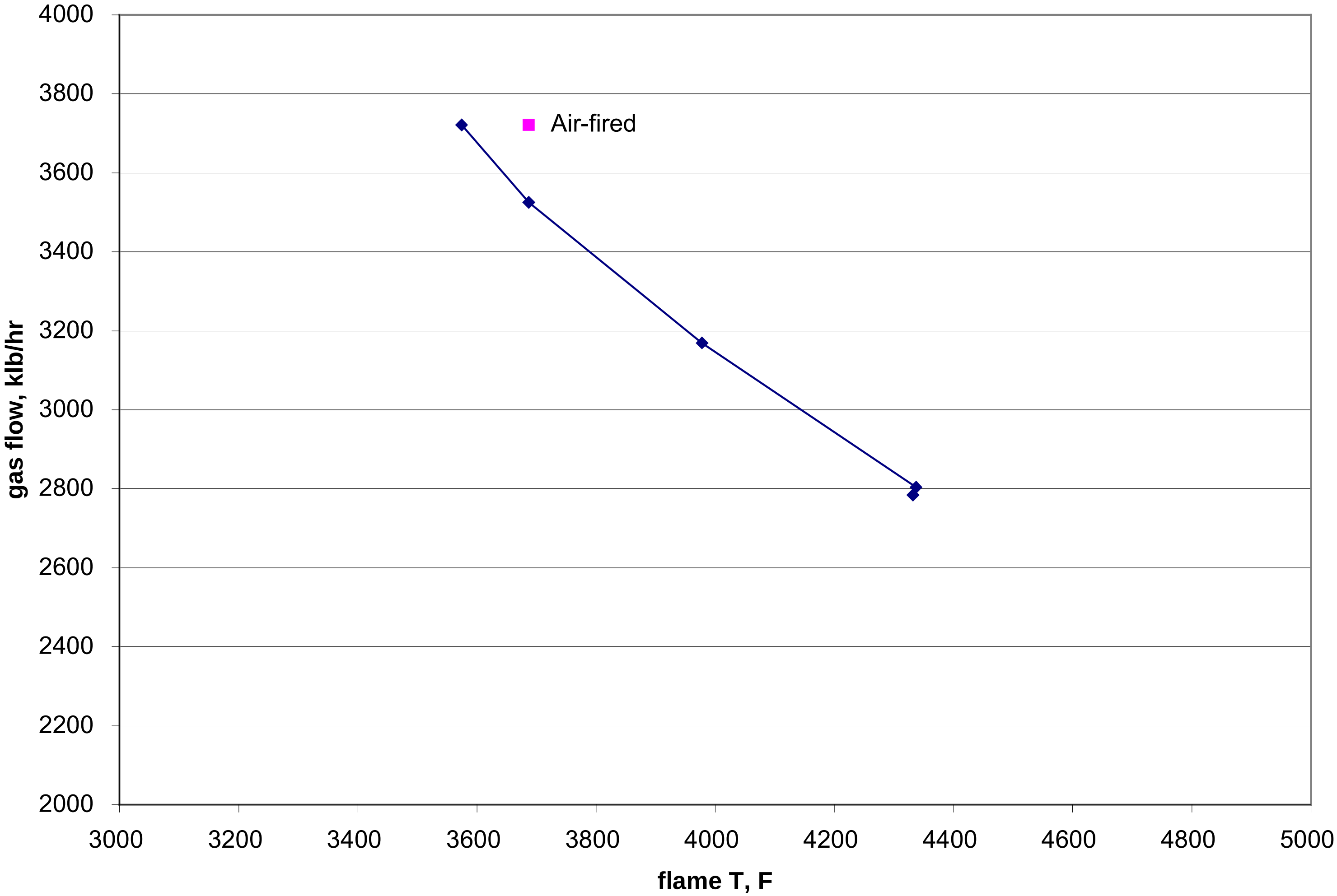


Figure 17 – Recycled Gas Volumetric Flow Rate Versus Flame Temperature

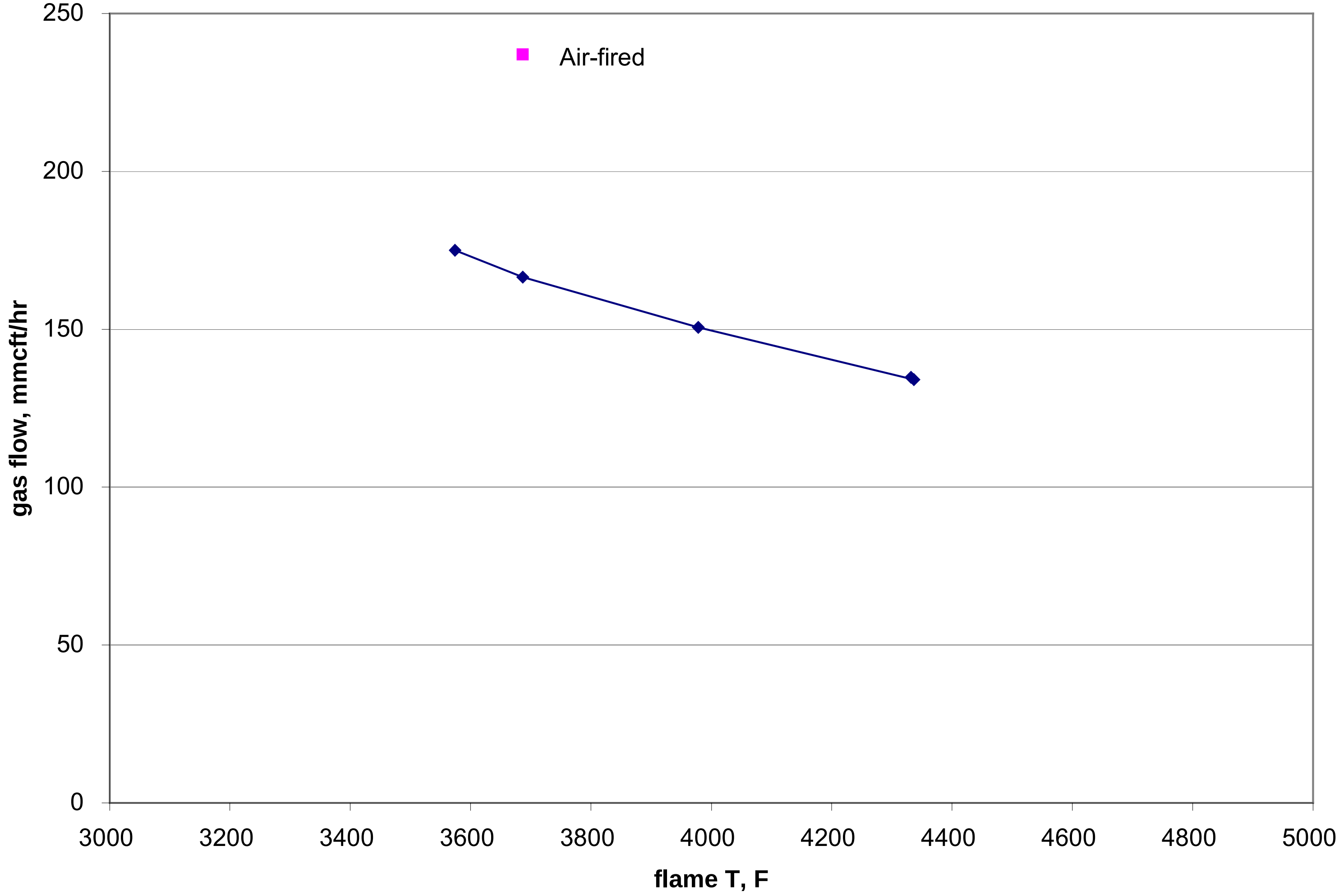


Figure 18 – Flue Gas Cooling Curve

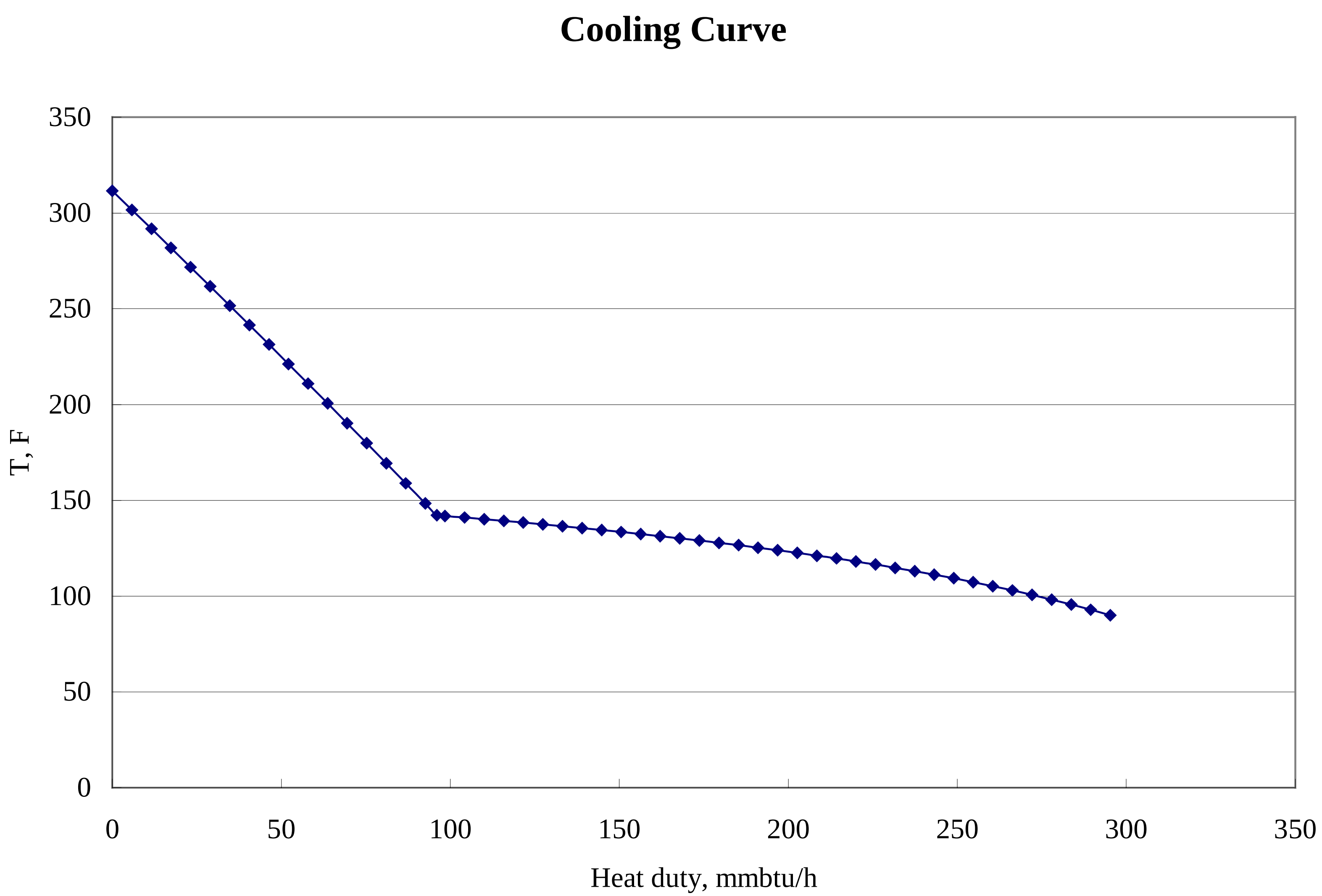


Figure 19 – Case 8 Cycle Diagram

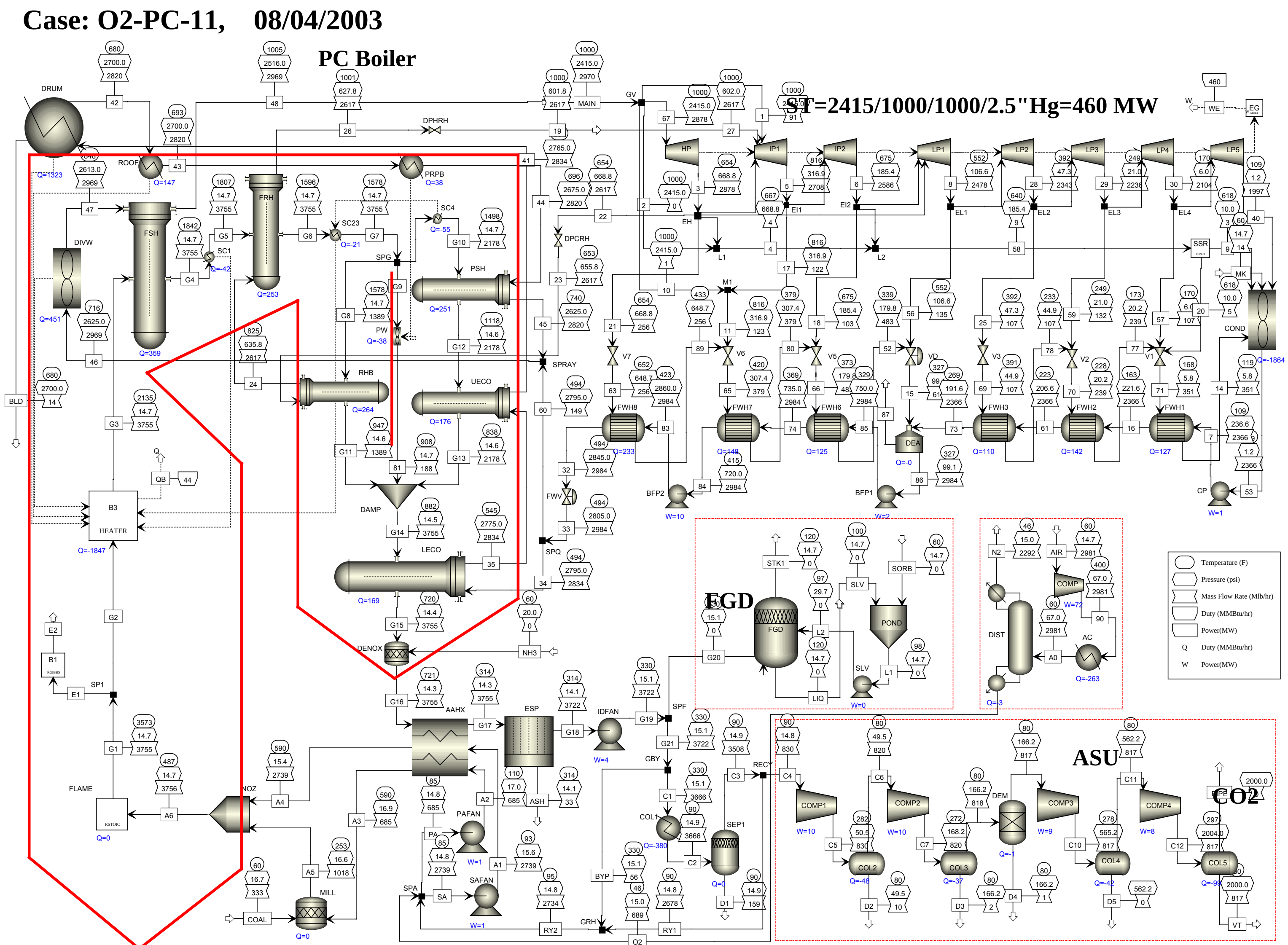


Figure 20 – Case 9 Cycle Diagram

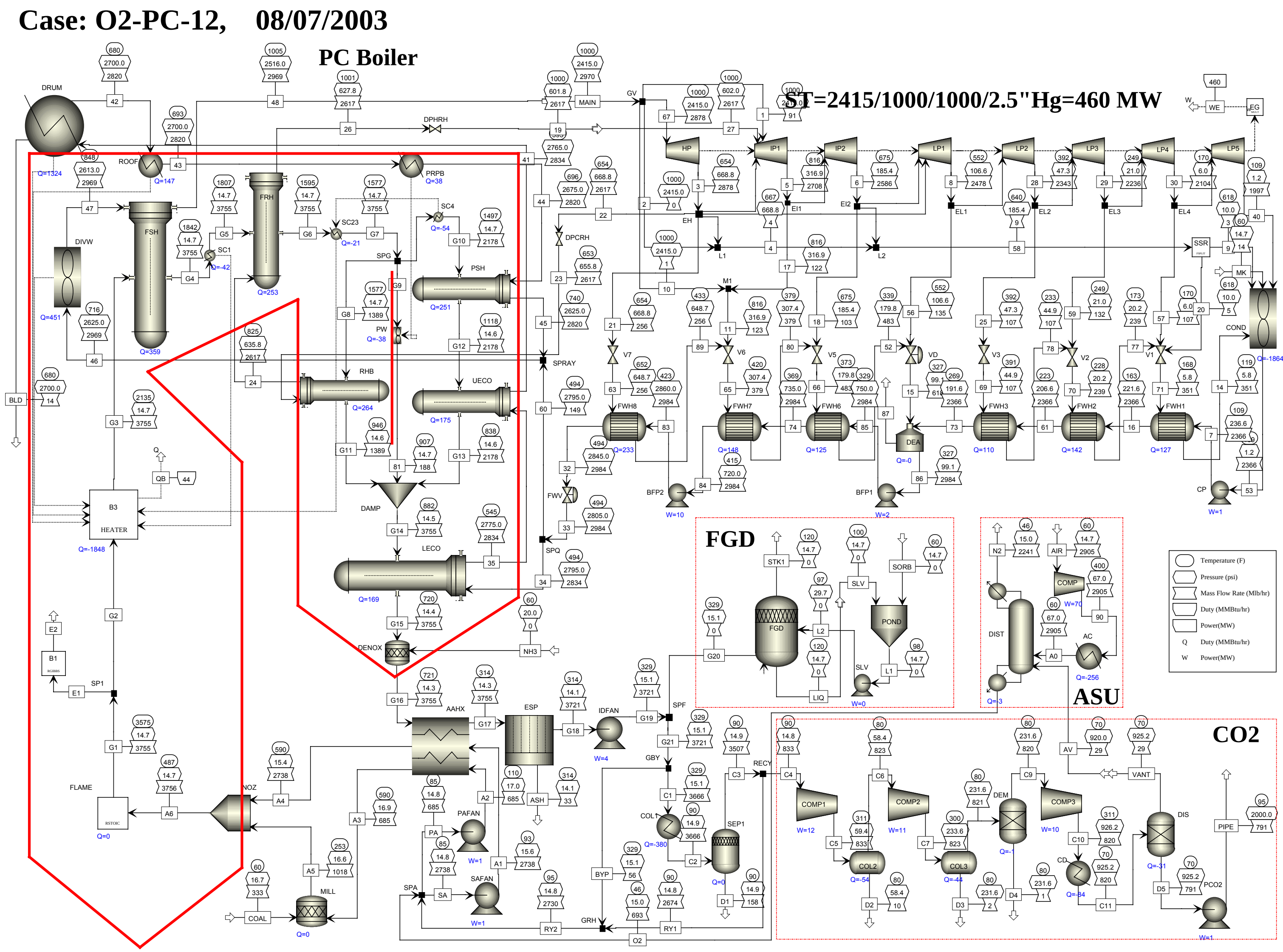


Figure 21 – Case 10 Cycle Diagram

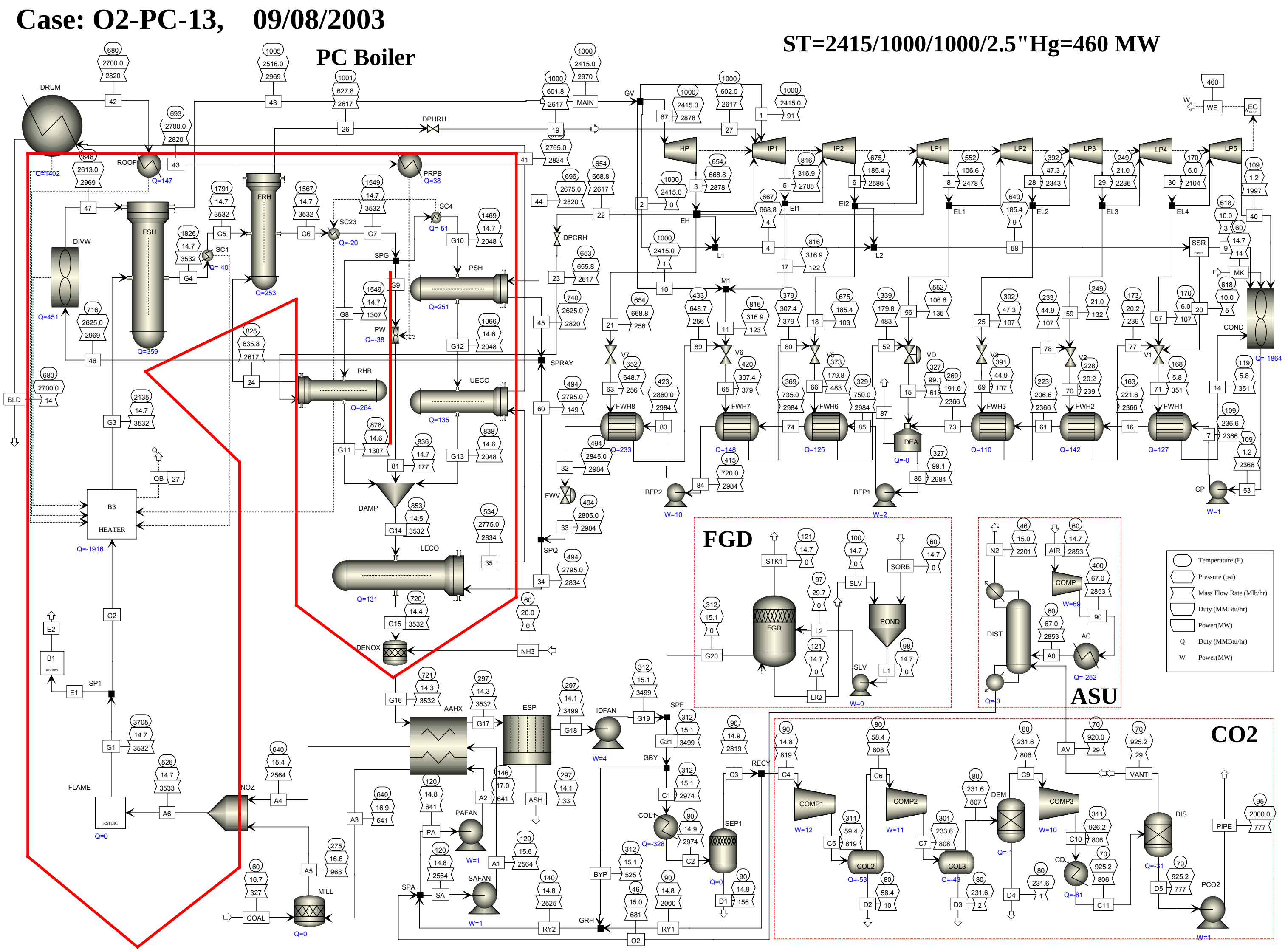


Figure 22 – Case 11 Cycle Diagram

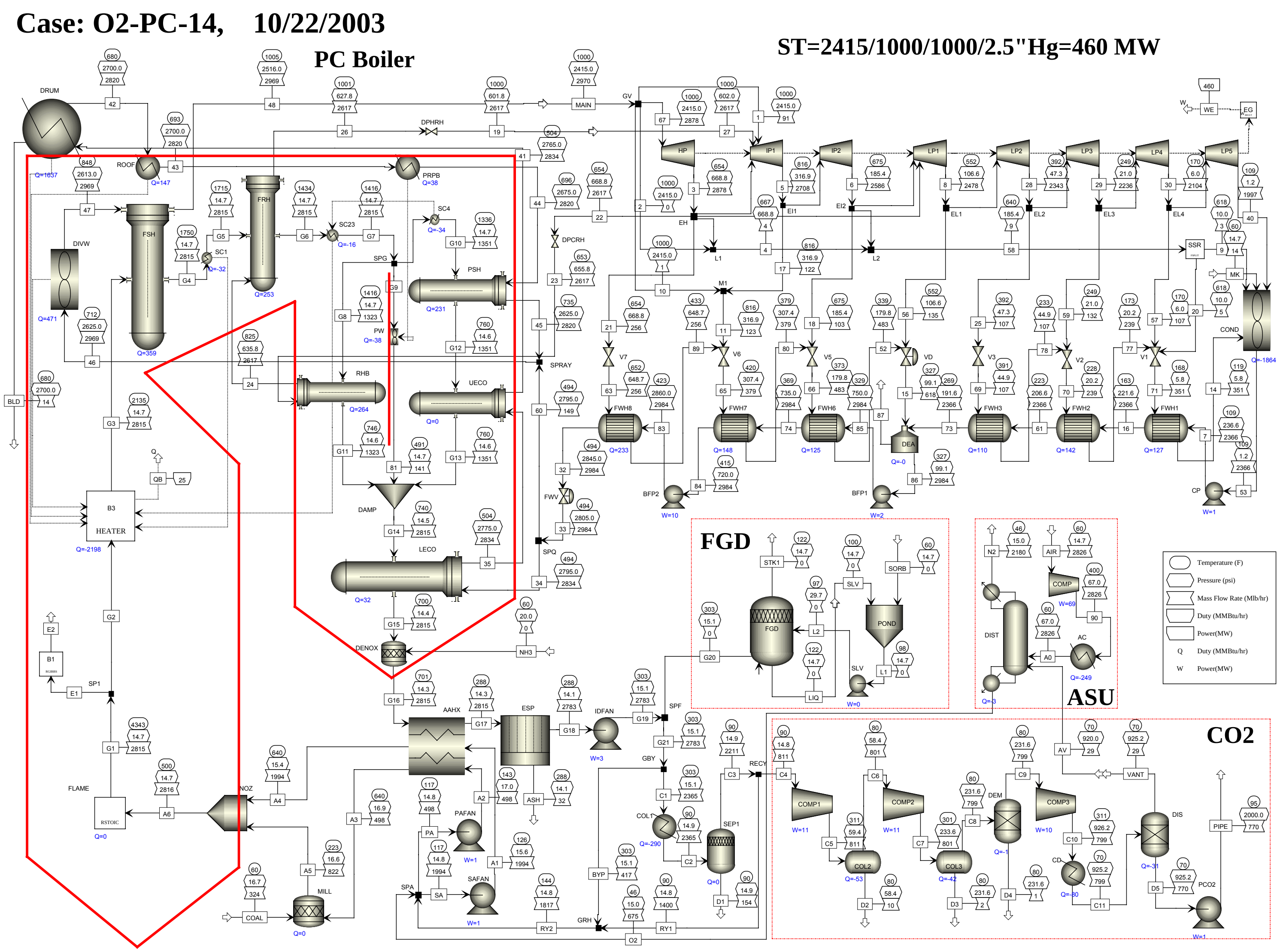


Figure 23 – Parametric Cases Summary (With CO_2 Compression)

case	1	3	8	9	10	11	12
ASU	no	yes	yes	yes	yes	yes	yes
CO2 Sequestration	no	no	yes	yes	yes	yes	yes
Vent gas recover	no	no	no	yes	yes	yes	yes
Wet-end economizer	no	no	no	no	no	no	yes
coal to boiler, klb/h	334	333	333	333	327	324	324
air to plant, klb/h	3422	2981	2981	2905	2853	2826	2826
O2 purity, %	20.7	99.6	99.6	99.5	99.5	99.5	99.5
O2 level to boiler, %	20.4	26.9	26.7	27.1	28.0	33.9	33.9
O2 level exit boiler, %	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Excess Air, %	18.0	3.1	3.1	0.3	0.3	0.3	0.3
flame T, F	3687	3574	3573	3575	3705	4343	4343
gas to boiler exit, klb/h	3721	3721	3722	3721	3499	2783	2783
gas to boiler exit, mmcft/h	237	175	175	174	166	134	134
gas recycle to boiler, klb/h	0	2734	2734	2730	2525	1817	1817
gas exit system, klb/h	3721	820	820	833	819	811	811
ASU plant power, MW	0	72.3	72.3	70.5	69.2	68.6	68.6
CO2 compression power, MW	0	0	37.2	34.6	34.0	33.7	33.7
Gross power, MW	460	460	460.0	460.0	460.0	460.0	464
Aux power, MW	42.1	105.8	143.1	138.6	136.4	133.9	134.6
Net Power, MW	417.9	354.2	317.3	321.8	324.0	326.5	329.6
Boiler efficiency, %	88.2	88.3	88.3	88.3	89.9	90.8	93.2
Net efficiency, %	36.68	31.15	27.87	28.27	28.98	29.48	29.75

Figure 24 – Vapor Pressure of CO_2

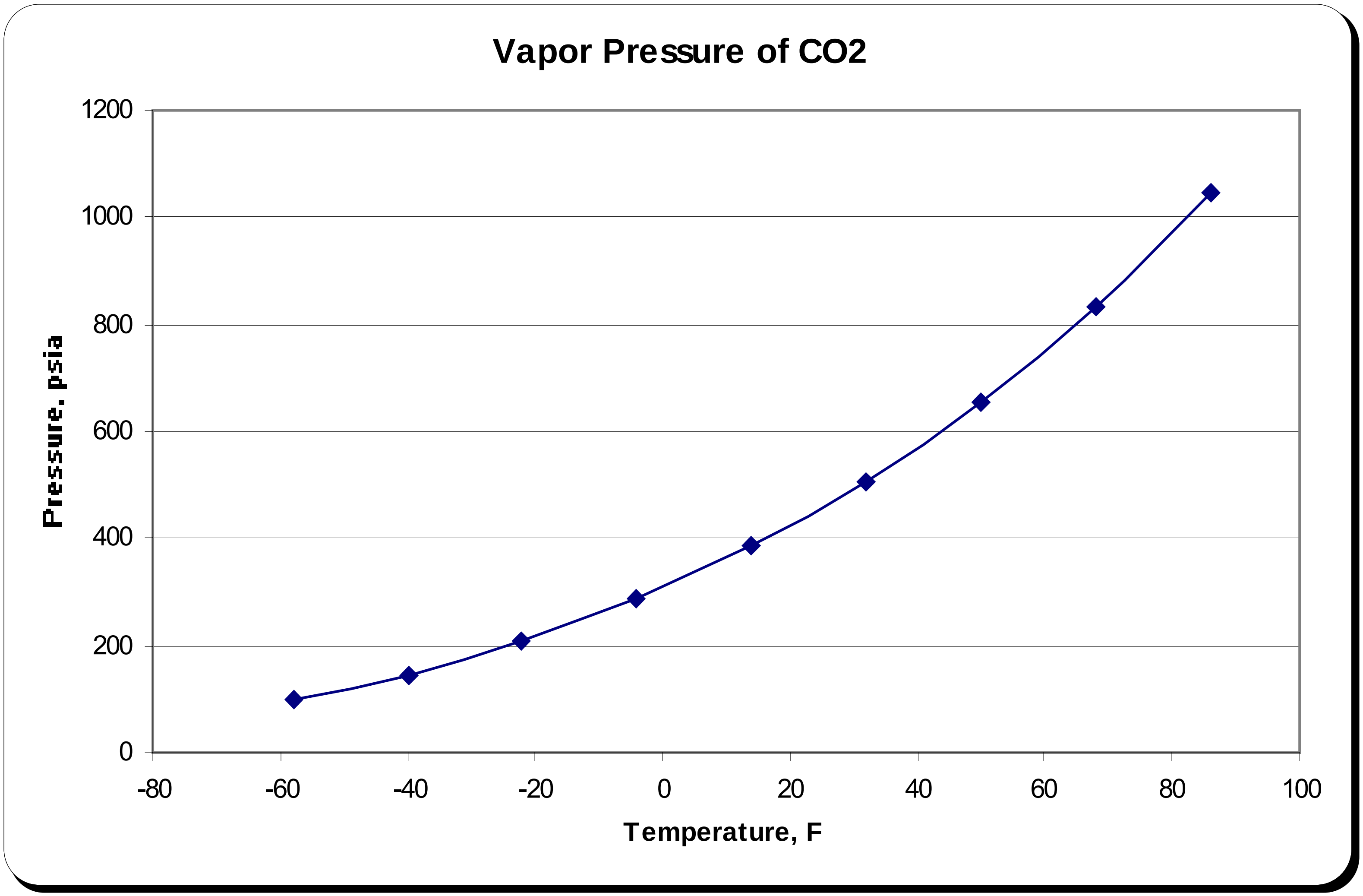


Figure 25 – Radiation Heat Flux and Water Wall Temperature

		1	2	3	4	5	6	7	8	9	10	11, 12
Gas Flow Rate	MM ft3/hr	237.0	181.0	175.0	166.5	150.6	134.0	134.8	175.0	174.4	166.0	134.0
Furnace Heat Duty	MM Btu/hr	1894	1848	1826	1904	2055	2201	2198	1847	1848	1916	2198
Furnace H2O	%	5.0	11.1	11.6	11.9	12.8	13.8	15.2	11.1	11.0	12.6	14.7
Furnace CO2	%	7.0	59.2	68.9	68.6	67.8	66.9	65.5	69.8	69.4	67.8	65.9
Depth	ft	35.9	31.3	30.8	30.1	28.6	27.0	27.0	30.8	30.8	30.0	27.0
Width	ft	51.0	44.6	43.8	42.7	40.6	38.3	38.5	43.8	43.7	42.7	38.3
Height	ft	206.9	155.7	155.0	155.8	153.4	142.2	139.9	158.1	158.8	154.5	140.2
Mean Beam Length	ft	30.9	26.2	25.8	25.2	24.0	22.5	22.5	25.9	25.8	25.1	22.5
Tube wall Properties												
Outside Diameter	in	2.75	2.75	2.75	2.75	2.75	2.75	2.75	2.75	2.75	2.75	2.75
Wall Thickness	in	0.285	0.285	0.285	0.285	0.285	0.24	0.24	0.285	0.285	0.285	0.24
Inside Diameter	in	2.18	2.18	2.18	2.18	2.18	2.27	2.27	2.18	2.18	2.18	2.27
inside H.T.C.	Btu/hr-ft2-F	2500	2500	2500	2500	2500	2500	2500	2500	2500	2500	2500
Material		SA-210C	SA-213-T2	SA-213-T2	SA-213-T2	SA-213-T22	SA-213-T91	SA-213-T91	SA-213-T2	SA-213-T2	SA-213-T2	SA-213-T91
Wall Therm. Cond.	Btu/hr-ft-F	22.0	21.8	22.2	22.2	20.0	16.1	16.1	22.2	22.2	22.2	16.1
Overall U	Btu/hr-ft2-F	583.3	579.6	587.0	587.0	544.9	540.6	540.6	587.0	587.0	587.0	540.6
Stress Allowable	psi	12,663	14,514	14,506	14,388	12,794	17,722	17,661	14,512	14,515	14,353	17,640
Min. Wall	in	0.29	0.26	0.26	0.26	0.29	0.22	0.22	0.26	0.26	0.26	0.22
Max. Furnace Temp.	F	3587	3508	3474	3587	3878	4237	4232	3473	3475	3605	4243
Max. Heat Flux	Btu/hr-ft2	63,090	97,505	99,320	105,300	127,712	147,366	148,470	98,880	98,725	106,975	148,850
Max. Wall Temp.	F	788	848	849	859	914	953	955	848	848	862	955
Ave. Furnace Temp.	F	3079	3020	2995	3079	3299	3579	3575	2994	2995	3092	3583
Ave. Wall Flux	Btu/hr-ft2	47,830	69,904	70,688	75,380	87,224	106,620	107,710	70,244	70,126	76,579	107,829
flux/flux(case1)		1.00	0.68	0.68	0.63	0.55	0.45	0.44	0.68	0.68	0.62	0.44

Figure 26 – Case 12 Cycle Diagram

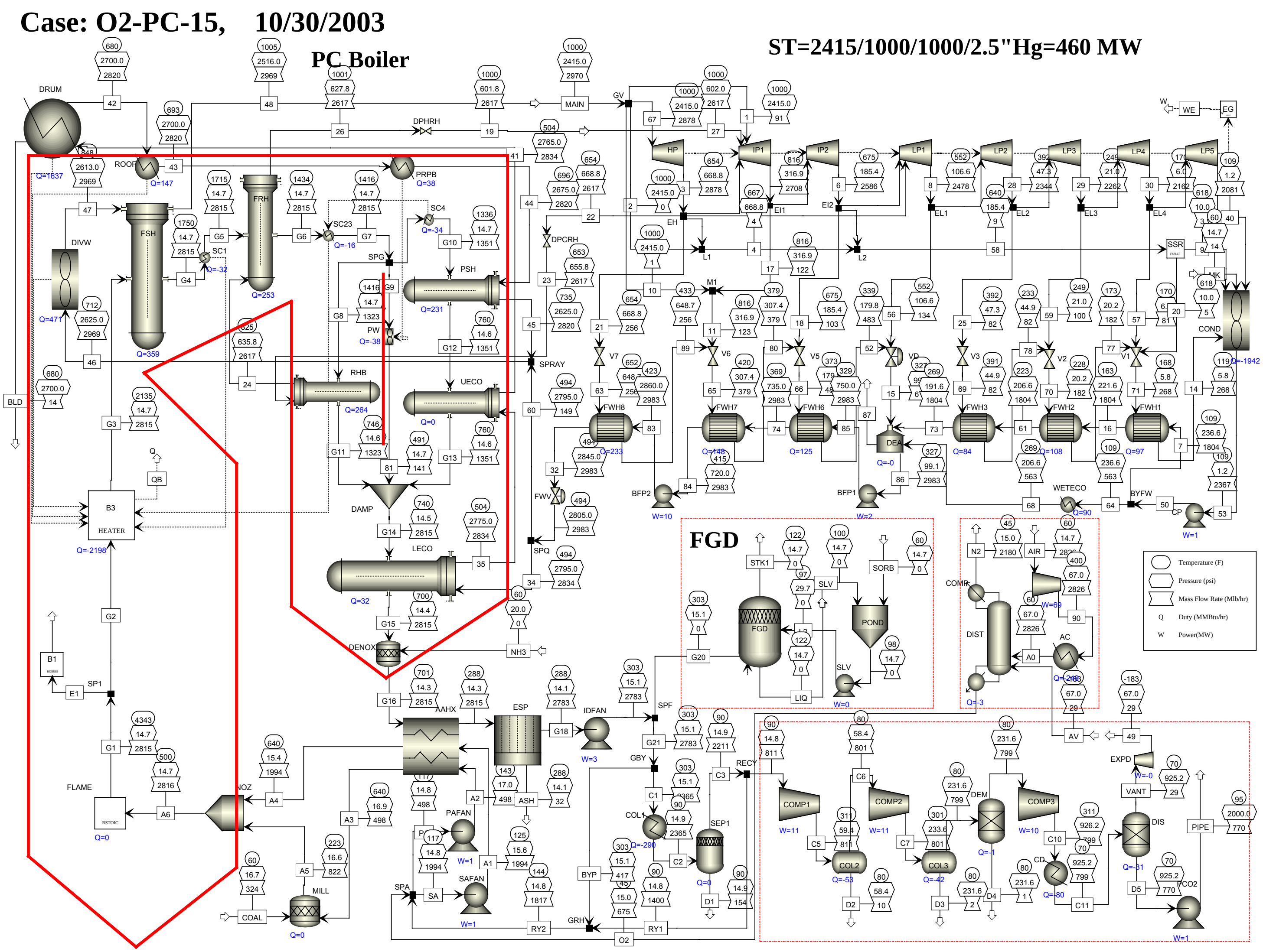


Figure 27 – Wet End Economizer Temperature Vs. Heat Duty

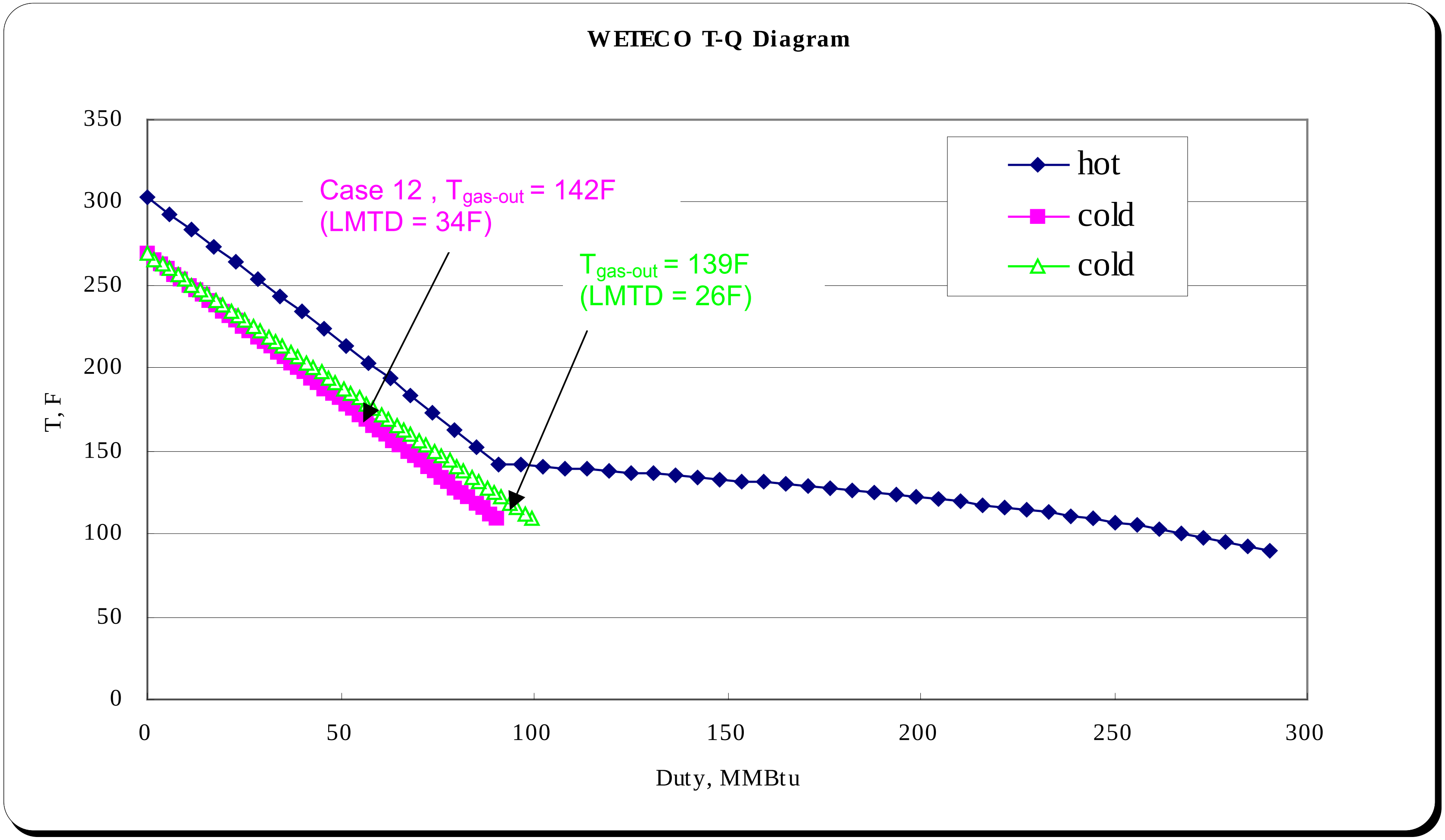


Figure 28 – Efficiency of Saved Extraction Steam Versus Stage

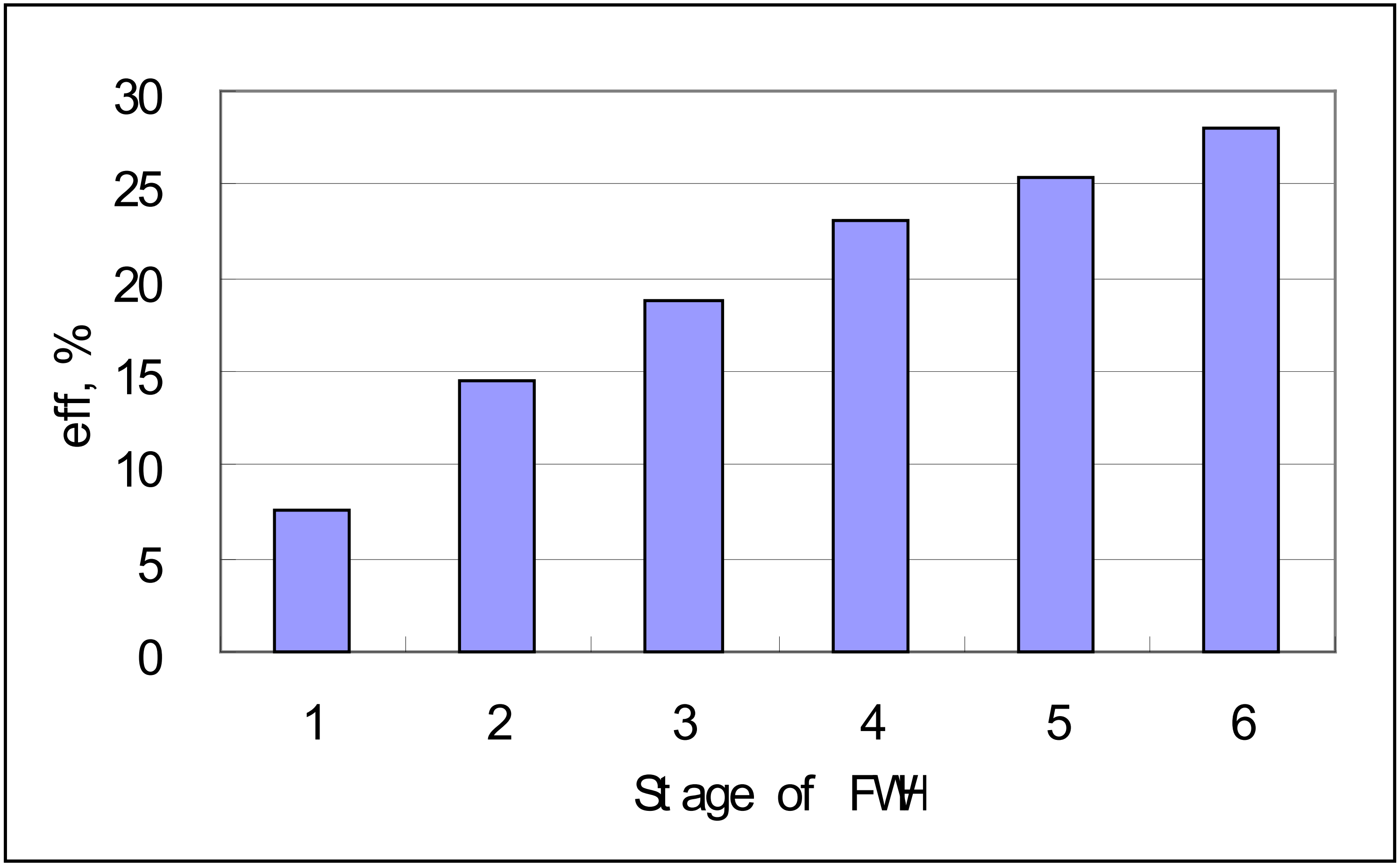


Figure 29 – Comparison of Power Consumption of CO_2 Removal of Different Technologies

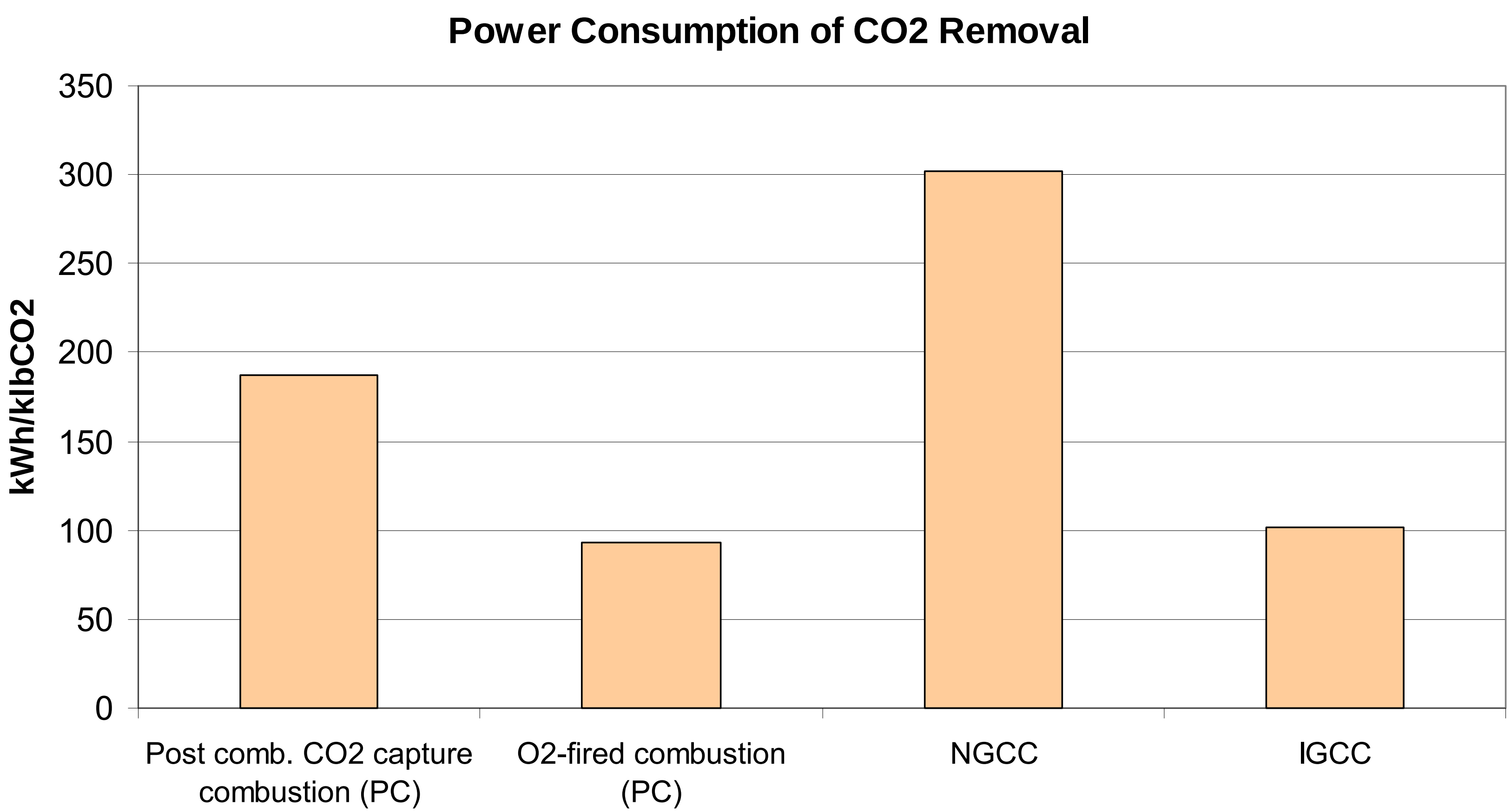
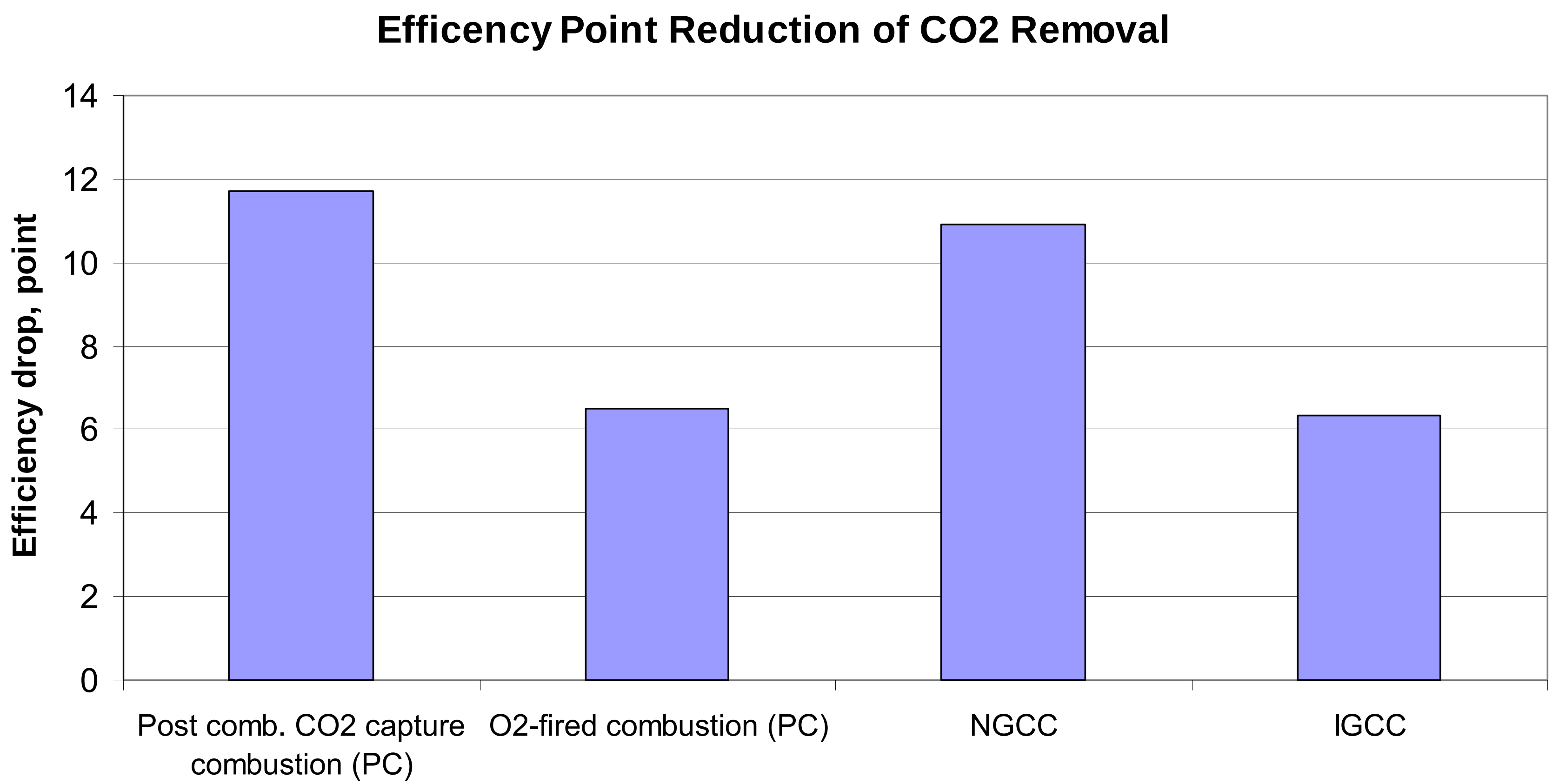


Figure 30 – Comparison of Overall Cycle Efficiency Point of CO_2 Removal of Different Technologies



5.0 Conclusion

To assure continued U.S. power generation from its abundant domestic coal resources, new coal combustion technologies must be developed to meet future emissions standards, especially CO_2 sequestration. Current conventional coal-fired boiler plants burn coal using 15-20% excess air producing a flue gas, which is only approximately 15% CO_2. Consequently, CO_2 sequestration requires non-condensable gases stripping, which is both expensive and highly power-consumptive. Several different technologies for concentrating the CO_2 by removing the non-condensable gases have been proposed including amine-based absorption and membrane gas absorption. However, these techniques require substantial energy, typically from low-pressure steam.

A new boiler is presented where the combustion air is separated into O_2 and N_2 and the boiler uses the O_2, mixed with recycled flue gas, to combust the coal. The products of combustion are thus only CO_2 and water vapor. The water vapor is easily condensed, yielding a pure CO_2 stream ready for sequestration. The CO_2 effluent is in a liquid form and is piped from the plant to the sequestration site. The combustion facility is thus truly a zero emission stackless plant.

The efficiency and cost-effectiveness of carbon sequestration in oxygen-firing boilers can rival competing gasification plants by specifically tailoring boiler design by appropriate surface location, combustion system design, material selection, furnace layout, and water/steam circuitry. Boiler efficiencies of near 100% can be achieved by recovery of virtually all of the flue gas exhaust sensible and latent heat. Boiler size can be drastically reduced due to higher radiative properties of O_2-combustion versus air-combustion. Furthermore, a wider range of fuels can be burned due to the high oxygen content of the combustion gas and potential for high coal preheat.

A conceptual design of a CO_2 sequestration-ready oxygen-based 460 Mwe PC boiler plant was developed. The selected O_2-fired design case has a system efficiency of 30.1% compared to the air-fired system efficiency of 36.7%. The design O_2-fired case requires T91 waterwall material and has a waterwall surface area of only 44% of the air-fired reference case. Compared to other CO_2 sequestration technologies, the O_2-fired PC is substantially better than both natural gas combined cycles and post CO_2 removal PCs and is slightly better than integrated gasification combined cycles. Furthermore, of the CO_2 sequestration-ready technologies, the O_2-fired PC is the simplest, requires the least modification of existing proven designs, and requires no special chemicals for CO_2 separation.

Thus CO_2 sequestration with an oxygen-fired combustion plant can be performed in a proven reliable technology while maintaining a low-cost high-efficiency power plant. As new lower power-consuming air separation techniques, such as membrane separation, become commercially available for large scale operation in O_2-fired plants, the CO_2 removal power consumption and efficiency reduction will continue to decline.

This study will continue with the following subsequent tasks:

Task 3: Combustion System Design and Analysis
Task 4: Furnace and HRA Design and Analysis
Task 5: Cost Estimate

6.0 References

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7.0 Bibliography

N/A

8.0 List of Acronyms and Abbreviations

A	Absorptivity
ε	Emissivity
σ	Stefan-Boltzmann constant
ASU	Air separation unit
CFD	Computational fluid dynamics
FD	Forced draft
FGD	Flue gas de-sulfurization reactor
HHV	Higher heating value
HRA	Heat recovery area
ID	Induced draft
IGCC	Integrated gasification combined cycle
LHV	Lower heating value
NGCC	Natural gas combined cycle
NOx	Nitrogen oxides
PC	Pulverized coal
Q/A	Heat Flux
SCR	Selective catalytic reactor
T	Temperature
TEG	Triethyleneglycol
U	Heat transfer coefficient
UBC	Unburned carbon loss