Optimization of Pressurized Oxy-Combustion with Flameless Reactor

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ABSTRACT

Pressurized Oxy-Combustion is one of the most promising technologies for utility-scale power generation plants. Benefits include the ability to burn low rank coal and capture $C0_2$. By increasing the flue gas pressure during this process, greater efficiencies are derived from increased quantity and quality of thermal energy recovery. UPA with modeling support from MIT and testing and data verification by Georgia Tech's Research Center designed and built a 100kW system capable of demonstrating pressurized oxy-combustion using a flameless combustor. Wyoming PRB coal was run at 15 and 32 bar. Additional tests were not completed but sampled data demonstrated the viability of the technology over a broader range of operating pressures, Modeling results illustrated a flat efficiency curve over 20 bar, with optimum efficiency achieved at 29 bar. This resulted in a 33% (HHV) efficiency, a 5 points increase in efficiency versus atmospheric oxy-combustion, and a competitive cost of electricity plus greater $C0_2$ avoidance costs then prior study's presented. UPA's operation of the bench-scale system provided evidence that key performance targets were achieved: flue gas sampled at the combustor outlet had non-detectable residual fly ashes, and low levels of SO3 and heavy-metal. These results correspond to prior pressurized oxy-combustion testing completed by Itea-Enel.

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EXECUTIVE SUMMARY

UPA's efforts to demonstrate the viability of pressurized oxy-combustion was successful in operating at 15 and 32 bar. The project did not achieve its stated objectives, but the operating results supported by MIT's modeling and cost estimates suggest that pressurized oxy-combustion with CO_2 sequestration is a superior technology to conventional pulverized coal and supports continued research into this technology.

Efforts to scale the system and design from a 5MW operating system in Italy to the 100 kW bench-scale at Georgia Tech was much more challenging and costly than originally estimated. The new 100 kW design did not achieve the expected engineering parameters , due to of deficiencies and a system failure, however the achieved results were inline with prior testing met by one of UPA's member companies.

The operation of the bench-scale unit on Wyoming PRB coal provided key process performance indicators supporting the benefits of pressurized oxy-combustion. Data taken of flue gas exiting the combustor provided zero residual fly ash, low levels of SO_3 (5 ppm and 19 ppm at 15 and 32 bar respectively), and low heavy metal content.

A significant disappoint was the inability to operate the bench-scale unit across the full pressure spectrum (up to 70 bar). The coal slurry pump did not meet its OEM designed tolerances which caused a project delay until an alternative pump could be identified. Eventually an inferior pump (guaranteed to operate at a maximum of 40 bar) was selected which resulted in a redesign of the project parameters.

In addition to the failure of the coal slurry pump, the system design was hindered by the objective of a perfectly geometrical scale-down of the combustor unit resulting in a less effective design of the fuel injection set which experienced severe damages at transient operating conditions. Furthermore, the decision to control expenses and simplify the system installation and commissioning at Georgia Tech was not effective. The use of a pre-formed refractory was believed to provide a quicker field installation but upon receipt the refractory was not formed to meet the exact parameters of the combustion vessel and the use of felts was not sufficient to provide a gap-free operating surface limiting the effectiveness in the refractory's interior chamber. . Finally, the selection of a mixed manual-automated control system was not effective and allowed for the impact of operator error to halt testing in the remaining scheduled time.

MIT's modeling pictured cycle efficiency dependence on pressure, concluded that pressurized oxy-combustion operates at a higher efficiency (33%) (on HVV basis) yet maintains flat levels (32-33%) over an ample range of pressure from 15 to 50 bar. MIT defined the peak operating value at 29 bar.

The techno-economic analysis defined the first year COE of flameless pressurized Oxy-combustion at $86/MWh, slightly below the DOE target of $80. This was supported by a calculated CO_2 avoided cost of 37 $/ton of CO_2 emissions which is better than values previously reported in DOE baseline studies.

UPA's efforts to build a 100kW pressurized oxy-combustion system did not achieve all its objectives, but it demonstrated operational viability at two pressures and completion of techno-economic analysis supports its operational and economic effectiveness compared with competitive CCS technologies.   This information corresponds to independent studies completed by UPA's member companies supporting further research and operational scaling of this technology.

PROJECT SCOPE

PROJECT BACKGROUND

Unity Power Alliance (UPA), a joint venture between Itea S.p.A. of Gallarate, Italy and ThermoEnergy Corp of Worcester, Massachusetts, was established in March 2012 to demonstrate the commercial feasibility and technical benefits of pressurized oxy-combustion for vertically integrated power utilities and independent power producers.

UPA's intent was to confirm the viability of oxy-combustion by designing, building and testing a bench scale pressurized, flameless reactor with an installed capacity of 100 kW. UPA also proposed to augment the empirical performance data obtained from machine operations with independent analytical models.

Although, already the focus of international efficiency and competitive studies, the pressurized oxygenation of coal lacks a full analysis supported by experimental data and advanced modeling tools for a wider range of promising operating pressures. UPA aimed to fill this gap before embarking on a more ambitious, and costly, larger scale development project.

The mission of the project FOA was to select and promote the development of promising advanced carbon, capture & sequestration (CCS) technologies for coal based power generation. A key goal was to identify a technology which would not exceed the cost of electricity (COE) of non-CCS state-of-the-art coal fired power plants by more than 30%.

UPA's pressurized oxy-combustion project was selected from ten candidates awarded the FOA. The initial proposal was primarily based the Itea's extensive 3-year advancement studies developing a 5 MW flameless, pressurized oxy-combustion pilot (operating at a range of 4-6 bar) with Enel S.p.A., a large Italian electric utility, in Gioia del Colle, Italy. The Itea technology was reinforced by ThermoEnergy's research and patented thermodynamic technology operating oxy-combustion at higher pressures under vacuum (>50 bar).

Prior research by NETL identified key issues concerning atmospheric oxy-combustion[1]:

i. A 12% loss in efficiency due to additional parasitic power consumption (for ASU and CPU);

ii. A large increase of byproducts (specifically fly ash, and SO_2) in the fumes, because of fumes recycling for tempering combustion; and

iii. Impact on COE of fumes purification prior to CO_2 compression/final purification

The Itea-Enel development work responded to these issues; results from this initial testing demonstrated the following:

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[1] NETL – Pulverized Coal Oxy-combustion Power Plants – August 2007, pages 22-23.

i. Zero soot and HPA, close to lower detection limit, low CO and NOx (zero thermal NOx);
ii. Coalescence and melting of the ash with a semi-quantitative segregation at combustor bottom[2];
iii. After heat recovery, clean flue gas exiting the combustor can be recirculated without the need for fume treatment. This gas is used to control the temperature in the combustor and to quench the hot gas prior to its entry into the boiler.[3];
iv. No trace of erosion / corrosion (including acid condensation) in the heat recovered pressurized fumes.[4]

UPA's objective in completing the testing protocol was to derive data to demonstrate operational efficiency at higher pressures while maintaining a control on the impact of COE due to fume purification and the increases in by-products. UPA identified critical data points: monitoring of residual (fly) ash and $S0_3$ in the fumes at the outlet of the combustor, and the presence of soot and heavy metal concentration impacting the $C0_2$ purification process.

STATEMENT OF PROJECT OBJECTIVES

In submitting a response to the funding opportunity announcement, UPA established four primary objectives:

1. Identifying the optimum operating pressure, flue gas recycling method and oxygen purity level for flameless pressurized oxy-combustion to improve overall system efficiency and COE performance;
2. Confirming the fate of toxic pollutants, with the goal of substantially reducing or eliminating emissions of such pollutants[5];
3. Confirming that the system design would achieve net water positive conditions and near zero level discharge of wastewater; and
4. Confirming, through modeling and bench-testing, that a flameless pressurized oxy-combustion system can be built and operated to improve the overall efficiency and achieve greater than 90% capture of carbon dioxide (CO_2) with less than a 35% increase in COE.

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[2] The melted ash resulted in a disposable as vitrified inert, non-leaching, zero-carbon slag

[3] Supporting the elimination of post-treatment of the cool fumes

[4] Based on testing at 4-6 bar with coal containing <2%wt sulfur

[5] During the initial project kick-off meeting (Pittsburgh, PA: June 19, 2012) , it was agreed to modify this objective and remove the analytical goal of testing for sulfur. Itea's experimental data on pressurized oxy-combustion demonstrated that incoming sulfur is present in the combustor as $SO_{2.}$

EXPERIMENTAL METHODS

To facilitate the projects, UPA developed a research team consisting of representatives from Itea and ThermoEnergy supported by the Massachusetts Institute of Technology (MIT) and Georgia Tech's Research Center (GTRC). MIT subcontracted to develop an accurate assessment of the technical, economic and environmental performance of the proposed pressurized oxy-coal combustion process. GTRC was selected to house the operating bench-scale system and provide third party collection of data and validation of results.

TECHNO-ECONOMIC MODELING

The UPA project scope included a modeling track under which Itea developed and provided MIT with a basis for modeling the flameless reactor and heat recovery steam generator (HRSG) inlet conditions. MIT set to develop a techno-economic model capable of assessing the optimal range of flameless, pressurized oxy-combustion while identifying the most efficient process cycle and operational conditions for power generation. This objective considered the maximization of the thermal efficiency of the cycle. With fixed and specified flow-rates, this objective was identified as equivalent to maximizing the net power output.

MIT's study was defined to examine the cycle efficiency dependence on process pressures. A secondary result was the expanded development of an operating cost model supporting a scaled 350 MWe system as designed in cooperation between Itea and Enel, an Italian utility operator. The selection of a generic "mid-west" site was selected with an actual net power output of ~328MWe. $C0_2$ product purity was selected based on characteristics for EOR as EOR is currently the most practically used method of sequestration. Upon conclusion of the capital cost analysis, MIT planned a study on operational COE.[6]

ENGINEERING, DESIGN AND CONSTRUCTION

The 100 kW, 70 bar rating, pilot unit was projected to demonstrate the feasibility of flameless condition/performance capable of running different ranks of coal (Texas lignite, Power River Basin and Illinois #6) up to 70 bar. Besides demonstrating operational success, collection of performance data was intended to complement the parallel modeling work on high-pressure oxy-combustion.

Georgia Tech was selected to host the bench-scale facility at their Atlanta Research Center.

The selection of Itea's flameless combustion technology was key to enabling the designed

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[6] To comply with DOE guidelines, some variables were excluded from the model optimization. Parameters like temperature and pressure of the feed-water entering the turbine, reheat extraction pressures and delivery temperature were also not incorporated. DOE approved methods were defined to be used for all COE projections.

pressurized oxy-combustion process. The designed combustor is a pressurized, slurry-fed coal combustion unit in which the chamber temperature is maintained at uniform levels and controlled by recycling flue-gas. Handling and separation of the resulting vitrified slag is part of the combustion unit package.

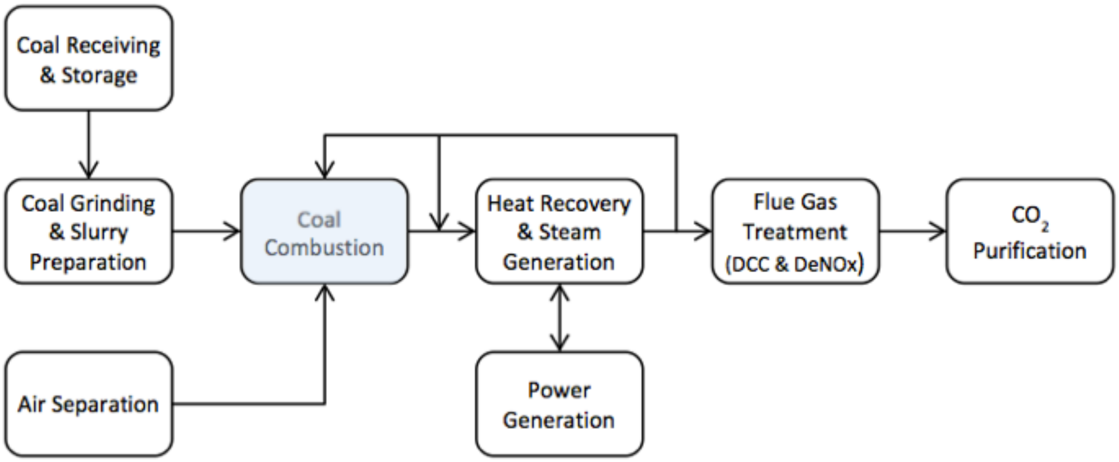

Image 1 | Pressurized oxygen combustion blow flow diagram

The system operates by pumping the coal slurry in with a staged, progressive cavity pump. Oxygen is fed via a separate injection system to the reactor, pre-heated with methane at pressures. Once the system reaches operating levels, the water in the slurry vaporizes and a subsequent reaction of the coal-02-steam mixture takes place.

In operation, preparation of the coal slurry would include grinding and water mixing, but with the bench-scale design, pre-ground coal was selected. High-pressure steam was added to spray the coal slurry using proprietary burners located in the vessel. A small fraction of boiler feed water (BFW) is used in the burner assembly to regulate heat. The proposed oxy-combustion process uses recycled flue-gas to maintain the combustor temperature at 1450°CFinally, the high operating temperature of the combustor, allows for the conversion of the ash content of the coal feedstock into a vitrified slag. This is eventually handled by an inclusive slag handling system. A secondary flue gas recycling stream is used to quench the raw flue gas from a state of combustion to 800°C prior to HRSG unit.

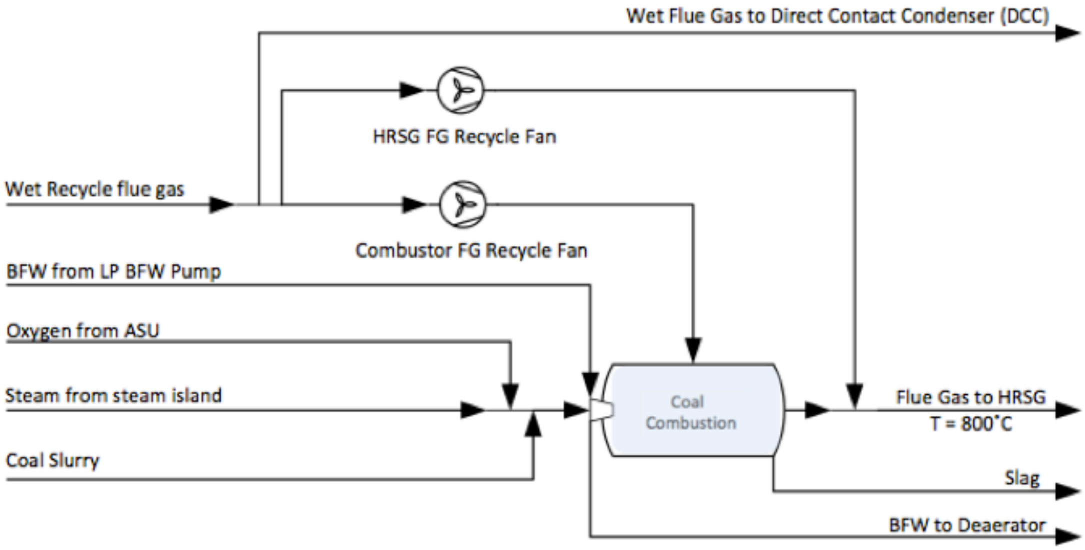

Image 2 | Process flow diagram of pressurized Isotherm oxygen combustion unit

SCALING OF DESIGN

Itea's experience in developing and operating the 5MWth system at Gioia del Colle was instrumental in the design of the 100 kW bench scale system. The initial design concern was whether the bench-scale would remain operational over a very broad range of pressure, from atmospheric to 7o bar, and whether a smaller burner could be successfully designed.

Although the basic design was understood , the engineering and construction objective was to develop a modular design that could facilitate the subcontract of the base unit at ThermoEnergy's Worcester, MA assembly facility, with shipment of acquired components to Georgia Tech. This decision led to the selection of different criteria, including the use of a pre-assembled refractory. Due to the geometric limitations of the smaller vessel design, the decision was made to use a single static lance supplying auxiliary fuel and the coal slurry into the vessel.

For this 100 kW pilot unit, high fumes pressure HRSG was omitted, and recycled fumes (average water vapor concentration ranging from 45 to 70 % vol) was replaced with steam.

Itea undertook the initial effort to develop detailed engineering drawings for building the bench scale pressurized oxy-combustion system. The initial system design was completed during the first half of 2013 and provided to ThermoEnergy., Itea worked closely with ThermoEnergy during the fabrication process to assure full adherence to the design. To further support this effort, Itea relocated an engineer to ThermoEnergy's facility.

Unfortunately scaling constraints continued to be identified after the initial design was released resulting in rework and construction of the bench-scale unit after release of the drawings to

ThermoEnergy. Time and cost delays supported the decision to move all flushing and cooling from automated hardware to manual actuation.

As a non-permanent structure, UPA decided to locally source elements to facilitate a lower cost of construction. These included a 200 kW electrical generator, diesel fuel system, 100 ton cooling tower, compressed air, oxygen, nitrogen, propane, and methane tanks, piping and related supplies. Itea was unable to identify a locally available (Southern US) steam source and UPA was forced to purchase a 4 M BTU Steam Boiler with Electrical Preheat Tank.

Upon completion, the bench-scale system was mobilized and shipped to Georgia Tech. The unit was initially expected to be housed at Georgia Tech's Carbon Neutral Facility but due to other schedules and the size of the bench-scale, the decision was made to locate the facility outside.

PHASE II SUBMISSION

UPA's original intention was to confirm the pressure range and cycle configurations using the bench scale reactor and supplement this work with a Phase II submission scaling the bench scale to a pilot scale facility of 50 MW. In cooperation with Enel, Itea had already initiated engineering of the 50MW design and submission of a Phase II became dependent on funding and the identification of a US partner willing to commit physical and financial resources. UPA's efforts in this area were unsuccessful. Participants demonstrated a willingness to support the technology but no one was willing to provide the financial support or facility access to support the scaling to 50MW. A reiterated cause for this reluctance was the competitive price of natural gas compared to coal.

RESULTS AND DISCUSSIONS

UPA was not able to complete the full experimental scope proposed for this project, but the construction of the bench-scale unit supported by MIT's modeling and GTRC data analysis gives support to the future feasibility of pressurized oxy-combustion.

PROCESS OPTIMIZATION AND TECNO-ECONOMIC MODELING

MIT performed its analysis in two stages: (1) model development and optimization, and (2) techno-economic assessment of the optimal configuration and operating conditions.

MODEL DEVELOPMENT AND OPTIMIZATION

In the first step, a process model[7] was developed using Aspen Process Economic Analyzer (Aspen©) Plus process simulation software to represent the pressurized isotherm oxy-coal combustion technology developed by Itea.

Operating conditions of various units, e.g., oxy-combustor, feed water heaters, deaerator, and the direct contact condenser (DCC), were optimized to maximize the net power output of the plant. The optimization analysis concluded that the wet-recycling configuration offers the highest performance and that the optimized efficiency is relatively flat for the pressure range of 20-60 bar achieving the highest efficiency at 29 bar. Therefore, within this range, other factors such the equipment size and costs will be instrumental in driving the selection and design of a commercial facility.

To better assess the effect of the operating pressures and importance of thermal recovery, a parametric pressure optimization analysis was performance as depicted below. The chart was constructed starting from the optimum pressure MIT found at 29 bar and incrementally varied while testing for all other variables.

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[7] This model is based on DOE guidelines provided in the FOA using Illinois #6 bituminous coal as feedstock and utilizes a supercritical steam cycle with single reheat. The process design assumes that product CO_2 will be used for EOR.

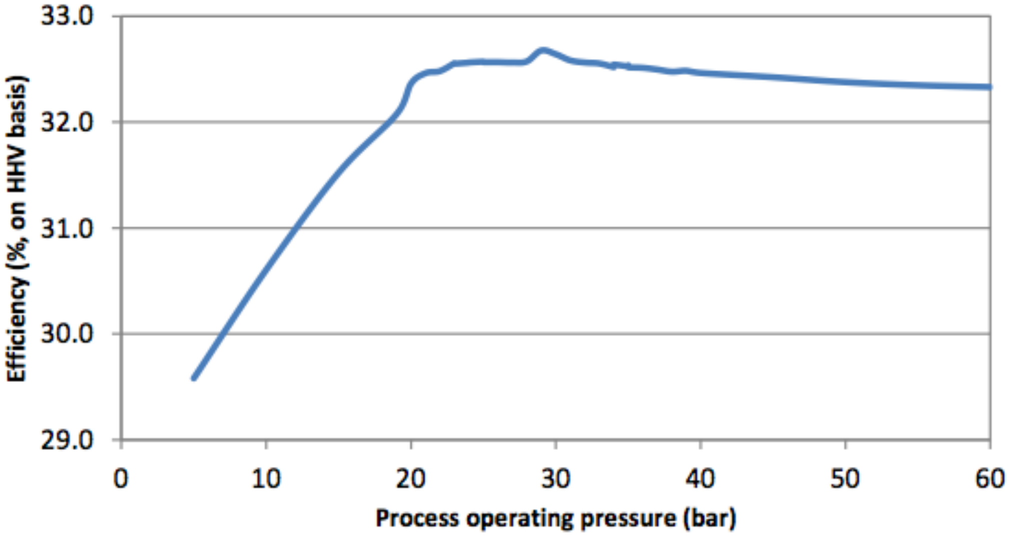

Image 3 | Plant net efficiency (on HHV basis) of the wet recycling configuration as a fluctuation of operating pressure

The level of the efficiency curve from 20 to 60 bar determined that at 20 bar the recovery section reaches its optimum operating level. This allows for the maximum heat recovery from the water vapor condensation, thus minimizing the heat loss in fumes cooling water.

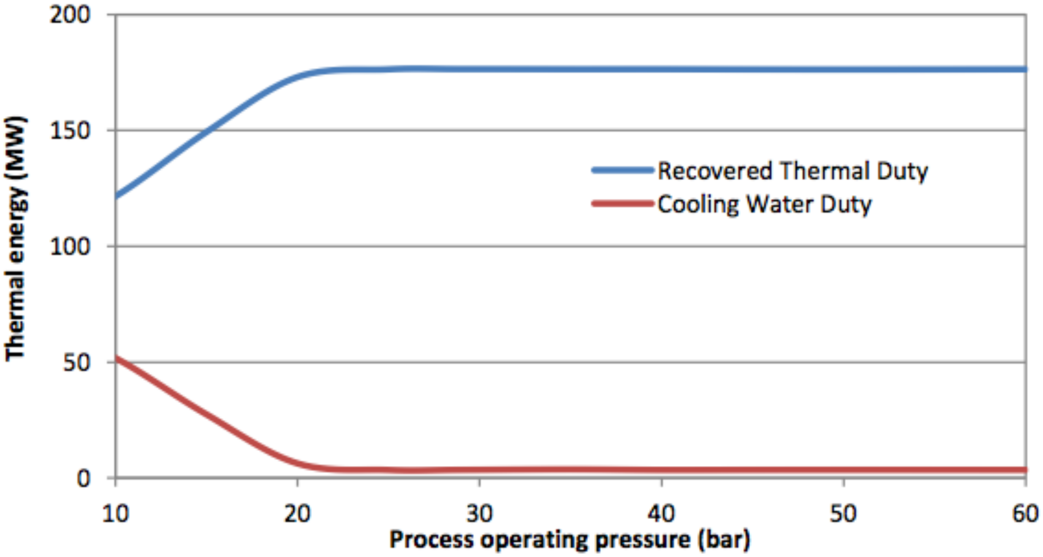

Image 4 | Recovered thermal energy and cooling water duty in the DCC at different operating pressures

MIT's results depict that above 20 bar the thermal recovery achieves a maximum level. Increasing the pressure above the 20 bar range barely increased thermal recovery or thermal losses to cooling water, however above at this range, pressure loss of the flue gas in the HRSG and recycling pipes increased.

Below 20 bar, operating pressures lower than the optimum decrease at a rapidly faster rate. As pressure decreases, the quality of the thermal energy leaving the flue gas as well as the quantity of recovered thermal energy decreases the low temperature of the feed water entering the deaerator resulting in a pressure decrease. Under these conditions, the thermal recovery section can no longer maintain a high temperature of the HRSG feed water leading to lower blended temperature ranges.

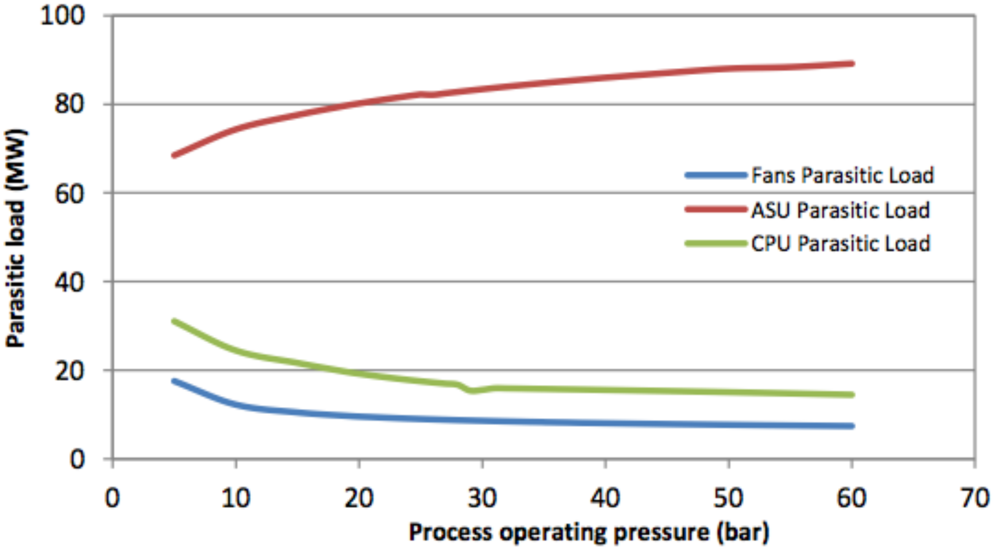

Image 5 | Distribution of parasitic load at different operating pressures

This behavior is also attributable to the operation of the DCC. The optimum recirculating water flow rate is not capable of sufficiently cooling and drying the flue gas at lower pressures; this results in the need for a higher amount of recirculating water. Increasing the flow rate of the recirculating water decreases the temperature of the recirculating water at the DCC-HX inlet. Therefore, the temperature of the feed water exiting the DCC-HX is also lower than optimal. The DCC-HX is not balanced -- the flow rate of the recirculating water is larger than that of the feed water -- thus the recirculating water exiting the DCC-HX has a comparably higher temperature. This unutilized thermal energy in the recirculating water is rejected into the cooling water before entering the contact column. As pressure decreases, the quality of the thermal energy leaving the flue gas decreases, the quantity of recovered thermal energy in the DCC-HX decreases, and the quantity of thermal energy lost to cooling water increases. Since the temperature of the feed water is exiting the DCC and entering the deaerator is relatively low, then the deaerator pressure decreases and/or the deaerator bleed flow rate increases. The thermal recovery section can no longer maintain a high temperature of the feed water entering the HRSG, which causes further reduction in the availability of the high temperature section. These different interactions explain the fast drop in efficiency for smaller operating pressures.

Results of the dry recycling process show a low efficiency compared to the wet recycling scenario, 25.5% compared to 33.5%. This can be explained by the amount of flue gas heading to the recovery section in the dry recycling configuration being significantly greater than that of the wet recycling configuration; therefore, the amount of thermal energy transferred at the high temperature section for the dry recycling is relatively small. The other disadvantage in dry recycling comes from the requirement to have a dry flue gas at the exit of the contact column; this necessitates a much greater flow rate of the recirculating water to reduce the temperature of the flue gas.

The process pressure in dry recycling is favored to be high, above 40 bar, in order to facilitate the condensation of the flue gas in the contact column (the recycled flue gas is dry so the water concentration in the main flue gas stream is low); yet this only slightly enhances the recovery section performance since the required recirculating water flow rate is still relatively large. Also, with higher operating pressures, the pressure losses increase, further affecting the performance of the dry recycling configuration. Potential risks exist from the condensation/corrosion in dry recycle configuration in the heat exchange process between the hot wet flue gas and the cold dry flue gas.

The performance of the dual-recycling configuration is found between its components (wet and dry recycling) and therefore does not offer any advantage over dedicated wet recycling configuration.

OPTIMAL CONFIGURATION

MIT's techno-economic assessment examined the cycle efficiency dependence on process pressure. Capital cost, for unconventional firing equipment and systems, was also estimated[8]. The resulting analysis provided a COE using DOE approved methods.

In this second stage, the technical, economic, and environmental assessment of the optimal configuration with optimized operating conditions (called proposed process or proposed plant hereafter) was performed. Table 1 depicts the overall plant performance results for the proposed process.

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[8] Capital cost was based on supplier data derived from recent Enel-Itea detailed engineering study for a 350 MWe flameless unit.

Parameter	Unit	Value	Notes
Heat rate (HHV basis)	BTU/kWh	10,400	On HHV basis, 30506 kJ/kg dry coal (Illinois #6)
Gross Power	MW_e	454	
Total Parasitic Load	MW_e	126	
- ASU only		83	Excluding coal grinding and cooling tower
- CPU only		15	66% of total parasitic load
			12% of total parasitic load
Net Power	MW_e	328	
Efficiency (on HHV basis)	%	32.7	
Efficiency (on LHV basis)	%	33.9	
CO_2 Capture Rate	%	96	of total carbon contained in the feed coal

Table 1 | Plant Performance Summary

As can be seen, the overall plant efficiency of the proposed process (which includes CO_2 capture and purification to the levels required for EOR application) is superior than pulverized-coal power plants with post-combustion CO_2 capture. The main advantage of this proposed process is that the pressurized operation allows for the recovery and utilization of the latent heat of the moisture included in the flue gas (derived either from feed coal or produced during the combustion process). Conventionally this latent heat leaves the plant as moisture in the flue gas from the stack. The same process can be applied to lower rank coal (lignite and sub-bituminous) with higher moisture content and obtain similar results since the latent heat associated with the moisture content of the feed coal is recovered.

COST ANALYSIS

The capital cost analysis is the first study incorporating cost and performance data from suppliers and engineering firms for the unconventional equipment and systems of the firing sections.[9] For common pieces of equipment, e.g., pumps and shell & tube heat exchangers, items were sized and costs estimated using APEA®. Cost and performance data for common units not included in the process model were taken from available studies, e.g., DOE NETL baseline research. Finally, contingencies were added to cover uncertainty and the cost of requirements for additional components.

The results of the cost estimation[10] effort are summarized in Table 2; different levels of capital cost estimation are presented.

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[9] This assessment was derived from recent Enel-Itea detailed engineering study for a 350 MWe flameless unit.

[10] The guidelines provided by DOE [3] are used for estimating the different levels of the capital cost of the proposed plant.

Parameter	Unit	Value	Notes
Heat rate (HHV basis)	BTU/kWh	10,400	On HHV basis, 30506 kJ/kg dry coal (Illinois #6)
Gross Power	MW$_e$	454	
Total Parasitic Load **- ASU only** **- CPU only**	MW$_e$	126 83 15	Excluding coal grinding and cooling tower 66% of total parasitic load 12% of total parasitic load
Net Power	MW$_e$	328	
Efficiency (on HHV basis)	%	32.7	
Efficiency (on LHV basis)	%	33.9	
CO2 Capture Rate	%	96	of total carbon contained in the feed coal

Table 2 | Capital cost estimation results

The overall plant efficiency of the proposed process (which includes CO_2 capture and purification to the levels required for EOR application) is better than pulverized-coal power plants with post-combustion CO_2 capture. The main advantage of the proposed process is that the pressurized operation allows for the recover and utilization of the latent heat from the moisture present in the flue gas (either from feed coal or produced during the combustion process). In conventional operations, this latent heat leaves the plant as moisture in the flue gas from the stack. It should be noted that the same process can be applied to lower rank coal (lignite and sub-bituminous) with higher moisture content and still obtain similar results since the latent heat associated with the moisture content of the feed coal is recovered.

COE RESULTS

Based on generally accepted cost estimation methodology[11] and modeling[12], the financial assessment of the proposed plant was performed and compared with two other pre-defined scenarios, Conventional Supercritical Pulverized Coal (SCPC) power plant, and SCPC with CCS. Table ES 3 depicts the estimated first year-Cost of Electricity (COE) for the proposed plant and two other scenarios.

Case	First Year COE In base year dollars ($/MWh)
SCPC	59
Pressurized oxy-combustion	86 (DOE target: 80)
SCPC w/CCS	105

Table 3 | COE results of the proposed process in comparison with state-of-the-art technologies

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[11] National Energy Technology Laboratory, "Quality Guidelines for Energy Studies - Cost Estimation Methodology for NETL Assessment of Power Plant Performance," U.S. Department of energy, April 2011.

[12] Power System Modeling Tool (PSFM) version 6.6.4 developed by DOE

Although the estimated first year COE for the proposed process is greater than the COE of the conventional plants by more than 35% (DOE target), it is significantly lower than the COE of current post-combustion capture technology. Furthermore, the CO_2 avoided cost for the proposed pressurized oxy-combustion process is calculated to be $37/ton of CO_2 emissions which is better than values reported in DOE baseline studies for current post-combustion and pre-combustion technologies.

The sensitivity analysis of COE and CO_2 avoided cost for the proposed plant reveals that the most affecting parameters in the order of their effect are plant capacity factor, capital cost, and fuel cost.

Only a small fraction of fumes from the oxy-coal combustion process are released into the atmosphere so it is expected that the environmental performance of the proposed process is far better than those required by current standards and regulations; the amount of emission per MWh of electricity is very small, and because most of sulfur and mercury is removed in the pressurized CO_2 purification unit.

The proposed pressurized oxy-coal combustion process offers a balanced approach to capturing CO_2 emissions from utility scale coal power plants while maintaining the capital costs and COE relatively low when compared to state-of-the-art post-combustion and pre-combustion (IGCC) capture technologies.

PILOT COMMISSIONING AND OPERATION

The objective of the bench-scale reactor at Georgia Tech (GTRC) was to demonstrate the operation of the technology at a steady state over a range of pressures with varying ranks of coal. GTRC's sub-contract was defined to provide a host site for operation and testing, and provide independent validation of the technology by sampling the gas-fume stream taken from the quenching section. In addition to demonstrating operational ability, collection of performance data was intended to support the parallel modeling work on high-pressure oxy-combustion.

The project's experimental matrix was limited to flameless exploration over a broad pressure range (up to 70 bar), and called for testing of three coals with varying rank: Taxes lignite, Powder River Basin (subbituminous) and Illinois #6 (bituminous)[13]; the testing was scheduled to occur over a two-week period .

The pilot was designed by Itea and ThermoEnergy performed the skid-mounted construction.

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[13] The selection of coals was specified in discussion between UPA and NETL

Image 6 | UPA Bench-scale model at GTRC (operator side)

Image 7 | UPA Bench-scale model with gases/tanks at GTRC

Unit commissioning was initially performed cold in September 2013. This operation was unsuccessful. The high-pressure rated coal slurry pump (up to 70 bar) did not operate as specified, inhibiting the delivery of coal to the reactor. The installed pumps failed due to the physical properties of the fluid and could not maintain operation without constant flushing.

This delay forced an investigation into an alternative pump, capable of handling a 40-60% slurry and resulted in the selection and shipment of a replacement pump in late 2013. In sourcing the alternative pump, UPA was unable to identify a unit capable of operating at the original design pressure of 70 bar. The replacement pump was limited to operation under 48 bar, which was below the original bench-scale specification. Conservatively, two pumps were sourced to provide redundancy in operation, but upon restart, a seal failed (suspected due to freezing weather conditions) and the ongoing operation continued with a single pump.

Cold commissioning was reinitiated in January and followed into hot commissioning and operations. The transition to hot commissioning demonstrated operation viability successfully reaching two of the six test points, [14]but revealed a lack of hardware performance and operational control difficulties.

The selected boiler package experienced multiple blocks generated by voltage fluctuations from the rented generator. The designed 20 bar minimum steam supply pressure threshold experienced its own fluctuations resulting in poor steam metering.

To facilitate installation a preassembled refractory was specified but in operation this decision did not provide an adequate “sealing” of the refractory in the combustion chamber. During the

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[14] The initial tests were successful operating at 15 and 32 bar with Wyoming PRB coal

pressurized pre-heating phase, the lack of seals resulted in surface overheating (>300°C) along the top flange and at some nozzles in the shell.

The combustion chamber also experienced insufficient delivery of the feed steam; temperature at the lower combustion zone was below the projected "isotherm" conditions experienced at Itea's 5MW pilot.

Notwithstanding the system faults, preheating was achieved (up to 1380°C). Wyoming PRB coal was prepared in slurry form and fed to the combustor. The lower combustion zone temperature decline several hundred degrees below the expected isothermal condition. The first run was performed at 15 bar. Fume samples were taken to analyze residual ash content and gas composition. The second run was at a range between 30 – 32 bar.

During the switch to Texas lignite, operators failed to close the recycling value prior to re-initiating the slurry feed. This reverse, pressurized flow from the combustor through the feeding line caused the slurry feed pipe to seize and fail. It is unknown if the lance had experienced any prior damage but the low delta pressure between the lance and the combustor suggested the lance tube may have already been corrupt and damaged by the commissioning ramp-up in the first tests.

Following this failure, the test was halted. The feed lance and refractory were damaged and substantial design changes were required. GTRC completed its analysis and the system was disassembled to allow for pre-scheduled use of the GTRC location by another project.

DESIGN OF REFRACTORY AND FEED LANCE

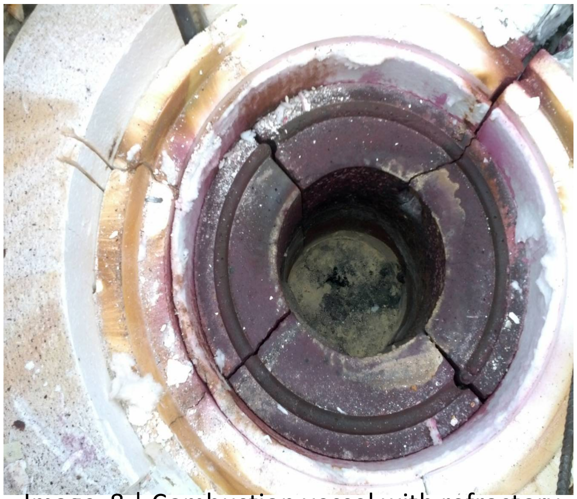

Image 8 | Combustion vessel with refractory installed (post testing)

The benefits achieved by using pre-assembled refractor were off set by the lack of proper seal in the refractory chamber. During system erection step, the technicians noticed the shipped package was oval, instead of cylindrical, in shape. This infield assembly, therefore did not allow for continuous contact of the package with the shell.. The joint gap-openings did not provide the proper refractory chamber confinement, and protect against over-heating of the combustor shell, nor prohibit the presence of condensation on the walls of the unit.. Condensation drained to the bottom of the vessel and visual inspection identified melt damage to the outer, lower temperature rated refractory surfacing the vessel walls.

On the right-top, the break is visible along the combustor axis of both shells. This depicts a path whereby fumes passed through the UNICOR break continuing through the high temperature felt junction, Alumina bricks junction, and outer insulating foam (damaged by the high temperature).

The identified hot fume pathway ran directly to the igniter nozzle which helps explain the overheating (>300°C) of the system.

It appears that overheating occurred at least when the fumes reached the surface walls; this occurrence is visible in the reddish area on the vessel. Other markings are rust caused by water condensation in the vessel chamber. (Image 9)

Image 9 | Condensation in vessel chamber (post testing)

Image 10 | Gap between vessel wall and refractory (post testing)

The gap between the felt and outer wall is also visible. The greatest gap measured 40 mm compared to original 11mm thickness of the felt. (Image 10). A similar, though smaller, gap was found at combustor outlet (located between the refractory inner channel and outer pipe).

The original design called for two felts (11mm and 15mm) to be used and applied surrounding the preassembled refractory, however during installation at the GTRC, the felts and refractory did not align properly into the vessel and the decision was made to eliminate the 15 mm felt wrap. This decision was made in error as the lack of felt left created a greater gap inviting leakage and excessive heat transfer. This caused the effect of both the hot spots and generalized water vapor condensation onto colder walls.

Image 11 | Inspection of bottom of combustion unit (post testing)

By design, a dry air-nitrogen flushing system was installed with the primary purpose of enhancing combustion refractory chamber confinement, and keeping the water condensation away from the walls of the vessel. Unfortunately, the above-identified gap made the flush system ineffective.

The decision to use pre-assembled refractory was made to facilitate the installation and to reduce the expense of bringing a European service technician to Georgia Tech.

Inspection on the combustor bottom revealed rust column patterns and an accumulation of condensed water. Approximately 180 liters was present, but there was no visual evidence of coal in the accumulated water.

SEM EDX analysis of a powdery material collected from the bottom of the combustor was found to be a mixture of unburned coal particle, with zero-Carbon slag granules. This corresponds to the historical [tip] failures characterized by molten slag produced on Itea's 5MW pilot at Gioia del Colle. The similarity between the tests was found to be the repeated performance of the fumes. There was not repetition in the findings between the zero-Carbon molten slag samples.

There was also no evidence of vitrified slag in the slag pot underneath the refractory chamber. Upon inspection, the slag discharge port was clogged by metal granules, which may have been generated by lance metal failure (see below).

The preheating air funnel / lance was severely damaged in operation; the area facing the opening is missing and approximately 40mm of material was no longer present. It is believed that this material melted at a temperature of 1450°C. The condition of the lance feed was even worse with over 200mm of material missing.


Image 12 – Damage observed on slurry feed lance (post testing)

Recorded data taken by Itea's technicians demonstrated the designed steam to oxygen ratio was not achieved during the second phase (higher temperature and pressure) of the pre-heating, because of a multiple steam boiler block (both power supply fluctuations and steam supply below boiler lower flow rate limit). Consequentially, the lance tip was progressively damaged during transients. There were also recorded evidence of insufficient "injection jet speed" by the lowering of combustor bottom temperature versus the isothermal profile.

The failure of the lance tip is apparent from recorded data of the coal slurry feeding run. The temperature unbalance, vs. the expected isothermal profile, was more than 300 °C indicating the fluid-dynamic was likely "a slurry dripping", rather than a "spray". The Combustor chamber's thermal balance calculation indicated that 4-5% of the slurry ,injected in the combustor, was dripping in the lower part of the combustor. The water evaporating from the slurry absorbed thermal energy therefore keeping the zone at a relatively low temperature (800 °C).

Yet, there is physical evidence that the feed system including lance melted onto the top inner wall of the combustor vessel. The grey buildup layer on the ceiling of the refractor is attributable to molten metal.


Image 13 | Top wall of combustion unit (post testing)


Image 14 | Close up to damage on top wall of combustion unit (post testing)

The bright red area (> 1500°C), partially masked by metal layer, is indicative of burning along the opening into refractory top cover. This occurrence further supports the thesis of the failure of the lance tip during pre-heating revealing a state of "metal cooking."

The secondary air funnel surrounding the lance body was also damaged, although it was not exposed to direct infrared radiation in the chamber. This ancillary system was installed to assist the first stage of preheating at atmospheric pressure.

It is still important to recognize that there did not appear to be any residual fly ash/solids in the fumes and there was no evidence of vitrified slag in the combustor bottom slag collecting pot, although the existence of material accumulation covering up to 250 mm was identified on the bottom of the reactor. This material was not uniform, and XRF identified both unburned coal granules and slag granules. Should the lance integrity was compromised, when the coal slurry was fed to the combustor, the combustor would have been fed with a water-slurry dripping into the vessel instead of being sprayed in the downward stream.

SYSTEM AUTOMATION

Since the bench-scale was only designed to operate on a short-term test basis, Itea intentionally designed the combustor with only partially automated ancillary. Instrumentation and manual actuated devices were not readily accessible to the operators and the unit was not designed to allow for visible inspection; there was no observation port with which to monitor/inspect flameless combustion.

Unlike Itea's 5MW system, the pilot was not equipped with automated procedures for stand-by/firing swap. The automated slurry-feed valve (open-close) procedure lacked control, allowing the slurry return line to be engaged at the same time. If correctly operated, this would have avoided the hot fumes return and failure.

ANALYTICAL TESTING AND VALIDATION

GTRC's sampling included collection of fly ash particulate (sampled at 200°C), SO3 vapor (condensed at 130°C), water vapor and all soluble species (condensed at 0°C) and permanent gases. Though limited to a single coal of the projected three coals, GTRC provided a data, which indicated industrial feasibility of high-pressure oxy-combustion.

PARTICULATE

The sampled slipstream fumes contained no fine particulate matter in either coal run. This could be the positive result of the fine particulate bonding with the slag in the combustion chamber; this would be beneficial for gas cleanup in the full-scale design. An alternative cause of this could be because the system design did not provide for an inlet for isokinetic sampling.

SO_3

SO_3 was sampled by a method consistent with EPA method 8A. SO_3 was ultimately found in the hot filter and in the fumes (later condensed) that passed through the filter. For the 15 bar run, 5 ppm of SO_3 was found in the sampled gas (dry gas basis), and 19 ppm was found for the 32 bar experiment. These levels may or may not be acceptable for the full-scale combustor and will have to be analyzed prior to subsequent development of the technology.

HEAVY METALS IN CONDENSATE

The condensate was analyzed by ICP-MS. Many of the metals tested for were either not present or below the detection limit. The metals that were found are summarized in the table 4 below. Concentrations are converted to micrograms per standard liter dry gas.

Metals Found in Condensate			Blank	Feb 17 A	Feb 17 B
Atomic Number	Name of Element	Symbol of Element	Ug/std liter	Ug/std liter	Ug/std liter
5	Boron	B	12.5965	10.1018	25.8463
11	Sodium	Na	1.3937	0.0825	0.0647
12	Magnesium	Mg	0.1196		
13	Aluminum	Al	0.1902		
14	Silicon	Si	0.2983	3.4006	2.1494
16	Sulfur	S	0.2003	2000.3541	335.2081
19	Potassium	K	0.2248	0.1300	
20	Calcium	Ca	1.5277	0.0825	0.1794
24	Chromium	Cr	0.0130	0.0150	0.0088
25	Manganese	Mn	0.2738	0.0775	0.0353
26	Iron	Fe	0.1686	0.1900	0.1999
27	Cobalt	Co	0.0019	0.0030	
28	Nickel	Ni	0.0130	0.0125	
29	Copper	Cu	0.0043	0.0025	
30	Zinc	Zn	0.3156	0.0425	0.0676
47	Silver	Ag	0.0003		
48	Cadmium	Cd	0.0010		
50	Tin	Sn	0.0159	0.0008	0.0282
56	Barium	Ba	0.0039		
58	Cerium	Ce	0.0003		
82	Lead	Pb	00.37	0.0008	

Table 4 – Metal in Condensate

According to UPA's evaluation, the combustor outlet fumes have zero residual particulate, SO_3 at 5.1 and 19.4 ppm dry basis (respectively at 15 and 32 bar) both very far from equilibrium, no soot and negligible heavy metals; these figures that correspond to flameless performance set.

IMPACT ON WATER

Under the SOPO, UPA intended to analyze the impact of pressurized oxy-combustion on the system wastewater. UPA's objective to verify the recyclability of the water condensate was not achieved. This process was not operationally test on the 100 kW bench scale system, however based on the results in Table 4, UPA believes the the removal of metals in the condensate demonstrates the viability of this process.

CONCLUSION

UPA's project promoting pressurized oxy-combustion provided favorable modeling and operational data. Testing results did not meet the stated objectives of the FOA but encourage providing a set of a comprehensive and detailed data supporting this technology.

The comprehensive results project an efficient design concept that is favorable to conventional pulverized coal system and best-in-class technologies. With a 4-5 point improvement in operational efficiency (32-33 % HHV) over an ample pressure range, the opportunity exists for a large degree of freedom in engineering the cost to performance optimization ratios. As currently designed and excluding any such improvements, economics modeling and analysis, now supported by cost/performance data from suppliers for unconventional unit operations, confirms COE figures and CO_2 avoided cost pretty close (+35% and 37 \$/t CO_2) compared to the DOE's started research and targets.

Experimental data from UPA's bench-scale tests using subbituminous Wyoming PRB, confirms "flameless-like" fumes performance over the explored 15 - 32 bar pressure intervals. The resulting analysis and characteristic testing of the fumes complements the feasibility data to process pressure cycle analysis results. Since only a small fraction of fumes from the oxy-coal combustion process are released to the atmosphere, it is expected that the environmental performance of the proposed process is much better than those required by current standards since the amount of emission is very small per MWh of electricity, and because most of sulfur and mercury is removed in the pressurized CO_2 purification unit.

The design inadequacies caused by difficulties in scaling the system from the larger 5MW to 100 kW have been analyzed and proper corrections already identified. Any unit refurbishment initiative must assume the return to the established fabricating procedure, with automated design and the foamed material applied in-field, The material must be"pressed" between the refractory body and felt layers to avoid the escape of hot fumes causing damage to the combustor.

The thermodynamic advantage of the proposed pressurized oxy-combustion process is the ability to recover heat for power generation. With the gas under pressure, the latent heat derived from moisture in the flue gas can be used in the DCC steam cycle for power generation and to increase the overall efficiency of the plant.

Additional testing into differing ranks of coal and operating pressures as originally defined (up to 70 bar) would be instrumental in better understanding the cost to efficiency curves for engineering and construction. This information could then be better correlated with results obtained by Itea and Enel in Italy.

Furthermore, a separate study on the impact of pressurized oxy-combustion on water usage, discharge and recoverability conditions could lead to significant savings and greater efficiencies.

Procedures and process characterizations in this area lack a generally accepted testing protocol. Although, this will not substantially alter the collected set of data, a standardized set of accepted measurements will allow more refined process correlations and optimizations in the future.

Based on MIT and GTRC's finding, the results support further investment into this technology and places UPA's pressurized oxy-combustion above competitively technologies in the industry. Additional testing and larger scale operations is recommended.

GRAPHICAL MATERIALS

LISTING OF FIGURES

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LIST OF ACRONYMS AND ABBREVIATIONS

APEA	Aspen Process Economic Analyzer
ASU	Air Separation Unit
BEC	Bare electric cost
BFW	Boiler feed water
CCS	Carbon, capture & sequestration
COE	Cost of energy
CPU	CO_2 Purification Unit
DCC	Direct contact condenser
DOE	U.S. Department of Energy
EOR	Enriched oil recovery
EPA	Environmental Protection Agency
EPCC	Engineering, procurement and construction cost
FPO	Flameless Pressurized Oxy-combustion
FOA	Funding Opportunity Announcement
GTRC	Georgia Tech Research Center
HHV	High heating value
HRSG	Heat recovery steam generator
HX	Heat exchanger
ICP-MS	Inductively coupled plasma mass spectrometry
IGCC	Integrated gas combined cycle
LHV	Low heating value
MIT	Massachusetts Institute of Technology
NETL	National Energy Technology Laboratory
PRB	Powder River Basin
SCPC	Supercritical pulverized coal
SEM-EDX	Scanning electron microscopy with energy-dispersive X-ray analysis
SOPO	Statement of project objectives
STD	Standard

TASC	Total as-spent cost
TOC	Total overnight cost
TPC	Total plant cost
UG	Microgram
UPA	Unity Power Alliance

APPENDICES

Appendix 1

UPA – Pressurized Isotherm Oxy-coal Combustion – MIT final report

2013

Massachusetts Institute of Technology

Navid Seifkar,
Hussam Zebian,
Randall Field,
Alexander Mitsos

[PRESSURIZED ISOTHERM OXY-COAL COMBUSTION]

This document serves as MIT’s contribution to the final report of the Isotherm Oxy-combustion Process project funded by DOE, under Award Number DE-FE0009478.

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Acknowledgement

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Disclaimer

Executive Summary

The purpose of this study is to present an accurate assessment of the technical, economic and environmental performance of the proposed pressurized oxy-coal combustion process for power generation. This report provides a concise overview of key results of the system analysis, which include benefits of the proposed technology, thermodynamic performance, economic performance, environmental performance, economic sensitivity to cost and performance of the proposed plant.

The analysis was performed in two stages: (1) model development and optimization and (2) techno-economic assessment of the optimal configuration with optimized operating conditions. In the first step a process model was developed using Aspen Plus process simulation software to represent the pressurized isotherm oxy-coal combustion technology developed by ITEA. This model is based on DOE guidelines provided in the FOA [1], Attachment B and uses Illinois #6 bituminous coal (Attachment C in [1]) as feedstock and utilizes a supercritical steam cycle with single reheat (Attachment B in [1]). Also, in the design of the process it is assumed that product CO_2 of the plant is used for EOR application which has the most restricted purity specifications, Attachment D in [1]. Using patent pending methodology and tools developed at MIT [2], structural and continuous optimization of the process were performed. For structural optimization, different flue gas recycling strategies (wet flue gas recycling, dry flue gas recycling, and dual recycling) were examined to find out the best process configuration. During continuous optimization, operating conditions of various units, e.g., oxy-combustor, feed water heaters, deaerator, and Direct Contact Condenser DCC, were optimized to maximize the net power output of the plant. The optimization analysis concluded that the wet-recycling configuration offers the highest performance and that the optimized efficiency is relatively flat for the pressure range of 20-60 bar. Therefore within this pressure range, other factors such the equipment size and costs will dominate the decision when designing a commercial facility. Within the relatively flat efficiency-pressure region, the highest efficiency was observed at 29 bar.

In the second stage of the analysis, the technical, economic, and environmental assessment of the optimal configuration with optimized operating conditions (called proposed process or proposed plant hereafter) was performed.

Using the optimized process model for the wet-recycle scheme, thermodynamic performance of the plant was examined and energy and mass (overall mass, carbon, sulfur, and water) balance results were obtained. Table ES 1 represents overall plant performance results for the proposed process.

Table ES 1 | Plant Performance Summary

Parameter	Unit	Value	Notes
Heat rate (HHV basis)	BTU/kWh	10,400	On HHV basis, 30506 kJ/kg dry coal (Illinois #6)
Gross Power	MW_e	454	
Total Parasitic Load - ASU only - CPU only	MW_e	126 83 15	excluding coal grinding and cooling tower 66% of total parasitic load 12% of total parasitic load
Net Power	MW_e	328	
Efficiency (on HHV basis)	%	32.7	
Efficiency (on LHV basis)	%	33.9	
CO_2 Capture Rate	%	96	of total carbon contained in the feed coal

As can be seen, the overall plant efficiency of the proposed process (which includes CO_2 capture and purification to the levels required for EOR application) is better than pulverized-coal power plants with post-combustion CO_2 capture. The main advantage of the proposed process is that the pressurized operation allows recovering and utilizing the latent heat of the moisture included in the flue gas (either from feed coal or produced during combustion process). Conventionally this latent heat leaves the plant as moisture in the flue gas from the stack. It should be noted that the same process can be applied to lower rank coal (lignite and sub-bituminous) with higher moisture content and obtain similar results since the latent heat associated with the moisture content of the feed coal is recovered.

A combined approach is taken to estimate the capital cost of the proposed plant. For new/advanced technological units, *e.g.*, pressurized oxy-combustor and pressurized HRSG, cost and performance data from vendors and engineering firms are used. For common pieces of equipment, *e.g.,* pumps and shell & tube heat exchangers, items are sized and cost estimated using Aspen Process Economic Analyzer (APEA®). Finally, cost and performance data for common units which are not included in the process model are taken from literature, *e.g.*, DOE NETL baseline studies. The results of the cost estimation effort are summarized in Table ES 2 where different levels of capital cost estimation are presented. The guidelines provided by DOE [3] are used for estimating the different levels of the capital cost of the proposed plant.

Table ES 2 |Capital Cost Estimation Results

Capital Cost Level	Value MM$ (2012Q1)
BEC (bare erected cost)	845
EPCC (eng., proc., and cons. cost)	929
TPC (total plant cost)	1,200
TOC (total overnight cost)	1,215
TASC (total as-spent cost)	1,383

Based on the methodology explained in [3] and by utilizing Power System Modeling Tool (PSFM) version 6.6.4 developed by DOE, the financial assessment of the proposed plant is performed and compared with two other pre-defined scenarios, Conventional Supercritical Pulverized Coal (SCPC) power plant and SCPC with CCS. Table ES 3 shows the estimated first year-Cost of Electricity (COE) for the proposed plant as well as the other two predefined scenarios.

Table ES 3 | COE Results of the Proposed Process in comparison with state-of-the-art technologies

Case	First Year COE in Base Year Dollars ($/MWh)
SCPC	59
Pressurized Oxy-combustion	**86** (DOE target: 80)
SCPC w/ CCS	105

Although the estimated first year COE for the proposed process is greater than the COE of the conventional plants by more than 35% (DOE target), it is significantly lower than COE of current post-combustion capture technology. According to the methodologies defined by DOE [3], the CO_2 avoided cost for the proposed pressurized oxy-combustion process is calculated to be 37 $/tonne of CO_2 emissions which is better than values reported in DOE baseline studies for current post-combustion and pre-combustion technologies.

The sensitivity analysis of COE and CO_2 avoided cost of the proposed plant reveals that the most affecting parameters in the order of their effect are plant capacity factor, capital cost, and fuel cost.

Since only a small fraction of fumes from the oxy-coal combustion process are released to the atmosphere, it is expected that the environmental performance of the proposed process is much better than those required by current standards and regulations since the value of emissions is very small per MWh of power and because most of sulfur and mercury is removed in the pressurized CO_2 purification unit.

In summary, the proposed pressurized oxy-coal combustion process offers a balanced approach to capturing CO_2 emissions from utility scale coal power plants while keeping the capital costs and COE relatively low when compared to state-of-the-are such as post-combustion or pre-combustion (IGCC) capture technologies.

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1. Introduction

The objective of this report (Final Report) is to present the results of the system-level analysis performed by Massachusetts Institute of Technology (MIT) of the proposed pressurized oxy-coal combustion process in partnership with ITEA and UPA. An accurate assessment of the technical, economic and environmental performance of the proposed process is performed. Both methodology and results are presented in this report. This report is organized as follows:

The basis of design for design of the proposed plant is described in Chapter 2 which includes site characteristics, assumptions regarding plant size, purity of CO_2 product, and train and sparing philosophy of the plant as well as plant capacity factor.

Using a patent-pending methodology, structural and continues optimization of various process configurations is performed to identify the best process scheme as well as operating conditions to maximize plant's power output. A brief description of this methodology and the optimization outcomes are presented in Chapter 3.

The discussions and results contained in Chapters 4 to 13 of this report refer to the pressurized oxy-coal combustion process (wet-recycle configuration) and with optimized operating conditions.

A detailed description of the purpose and process operation for each plant areas are presented in Chapter 4. Technology selections for key components and also explanation of key assumptions and design decisions, and a brief discussion of why any other reasonable options, if available, were not selected are included in this chapter. This Chapter also includes process flow diagrams of the main processing units within the plant.

A formatted description of various plant/process components, as per DOE guidelines Attachment A of [1], is included in Chapter 5. The overall process flow diagram and the corresponding stream table is represented in Chapter 6.

Chapters 7, 8, and 9 represent plant's energy and mass balance, thermodynamic performance, and overall technical performance in a format which is consistent with DOE reporting requirements, Attachment A in [1].

The assumptions, methodology, and results of the capital cost estimation of the proposed plant are described in Chapter 10. Assumptions regarding contingencies and owner's costs are stated in this chapter. Also, Chapter 10 describes assumptions used for estimation of operating and maintenance (O&M) costs.

The financial analysis of the proposed plant includes estimation of the first-year cost of electricity (COE) and the current-dollar, thirty-year Levelized COE using the NETL PSFM. Also, cost of avoided CO_2, cost of captured CO_2, and the breakdown of COE into its primary cost components (capital, fuel, variable O&M, and fixed O&M) are included in Chapter 11. Moreover, the results of the sensitivity analysis of key parameters impacting economic metrics (*i.e.*, COE and cost of avoided CO_2) with respect to fuel price, capital cost, and plant capacity factor are shown in Chapter 11.

A brief description of expected environmental performance of the proposed plant is included in Chapter 12.

As part of the project documentation, all final process models developed for this project are delivered with this final report. Also, for economic analysis of the proposed process, the NETL Power Systems Financial Model (PSFM) version 6.6.4 is used. A brief description of these two deliverables (Other Deliverables) is included in Chapter 13.

An analysis of the key findings of this study and recommendations for follow-on work is included in Chapter 14.

1.1. Project Deliverables

The followings were/are submitted to DOE as deliverables of this project:

- Design Basis (previously delivered to DOE according to the project schedule)
- Interim Technical Status Report (previously delivered to DOE in March 2013)
- Final Report (this document)
    - Process Model (delivered to DOE along with the final report)
    - Modified Power System Financial Tool (version 6.6.4) to represent the proposed process as Case G (delivered to DOE along with the final report)

2. Design Basis

This chapter describes assumptions used for the techno-economic evaluation of the proposed pressurized oxy-coal combustion process.

2.1. Site Characteristics

Site selection of for this study is in accordance with specifications for a “mid-west” site at ISO conditions, as detailed in Exhibits 2-1 and 2-2 in the NETL report, “QGESS: Process Modeling Design Parameters” [4] unless otherwise is noted. The values for assumed characteristics are shown in Table 1.

Table 1 | Site characteristics

Site Characteristics	Assumed Value
Topography	Level
Size[†]	17.5 acres (100 acres plant’s footprint including coal yard etc.)
Transportation	Rail or highway
Ash/Slag Disposal	Offsite
Water	50% Municipal and 50% ground water

[†] The considered size (plant footprint) for the proposed pressurized oxy-coal combustion plant is smaller than what is proposed in **[4]** due to higher densities of streams which result in smaller equipment sizes.

Table 2 represents the site conditions assumed for this study in accordance with specifications in the NETL report, “QGESS: Process Modeling Design Parameters” [4].

Table 2 | Site conditions for generic US mid-west location

Site Condition	Value
Elevation, m	0
Barometric Pressure, MPa	0.101
Design Ambient Dry Bulb Temperature, °C	15
Design Ambient Wet Bulb Temperature, °C	10.8
Design Ambient Relative Humidity, %	60
Cooling Water Temperature, °C	15.6
Air composition, wt. %	
H_2O	0.616
AR	1.280
CO_2	0.050
O_2	22.999
N_2	75.055
Total	100.00

2.2. Coal Characteristics

Illinois #6 Bituminous coal is considered as design coal for this study in accordance with specifications as detailed in Exhibit 1-2 in [5]. Table 3 shows different analyses of the design coal.

Table 3 | Analyses of design coal [5]

Parameter	Value	
Rank	HV Bituminous	
Seam	Illinois #6 (Herrin)	
Sample Location	Franklin Co., IL	
Proximate Analysis (weight %)		
	AR	**Dry**
Moisture	11.12	0
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon (BD)	44.19	49.72
HHV, kJ/kg	27,113	30,506
HHV, Btu/lb	11,666	13,126
LHV, kJ/kg	26,151	29,444
LHV, Btu/lb	11,252	12,712

Parameter	Value	
Ultimate Analysis (Weight %)		
	AR	**Dry**
Moisture	11.12	0
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen (BD)	6.88	7.75
Sulfur Analysis (weight %)		
	AR	**Dry**
Pyritic	-	1.14
Sulfate	-	0.22
Organic	-	1.46
Trace Components (ppm)		
Mercury	-	0.150

2.3. Plant Size

The techno-economic assessment performed for this study is based on a plant with net power output of ~328 MWe.

2.4. CO_2 Product Purity

Within the different options for EOR or saline reservoir CCUS presented in Attachment D of the FOA [1], the “Enhanced Oil Recovery”, with most restricted composition limits, is selected as CO_2 product purity limits for this project. The reason behind this decision is that currently EOR is the most practical method of sequestrating CO_2. It should be noted that choosing other options of CCUS with less-restricted purity limits may result in lower parasitic load of the CPU and higher overall process efficiency.

The results from the process model indicate that the impurities in the CO_2 product stream are well below limits specified in Attachment D of FOA package for the EOR case.

Table 4 | CO_2 product composition limits for EOR application

Component	Unit	Limit	Component	Unit	Limit
CO_2	vol% (Min)	95	**NO_X**	ppm$_v$	**100**
H_2O	ppm$_{wt}$	300	**NH_3**	ppm$_v$	**50**
N_2	vol%	1	**COS**	ppm$_v$	**5**
O_2	ppm$_v$	100	**C_2H_6**	vol%	**1**
Ar	vol%	1	**C_3^+**	vol%	**<1**
CH_4	vol%	1	**Particulates**	ppm$_v$	**1**
H_2	vol%	1	**HCl**	ppm$_v$	**N.I.**
CO	ppm$_v$	35	**HF**	ppm$_v$	**N.I.**
H_2S	vol%	0.01	**HCN**	ppm$_v$	**trace**
SO_2	ppm$_v$	100	**Hg**	ppm$_v$	**N.I.**

2.5. Train Philosophy

Figure 1 shows the train arrangement considered for the design of the proposed pressurized oxy-coal combustion power plant.

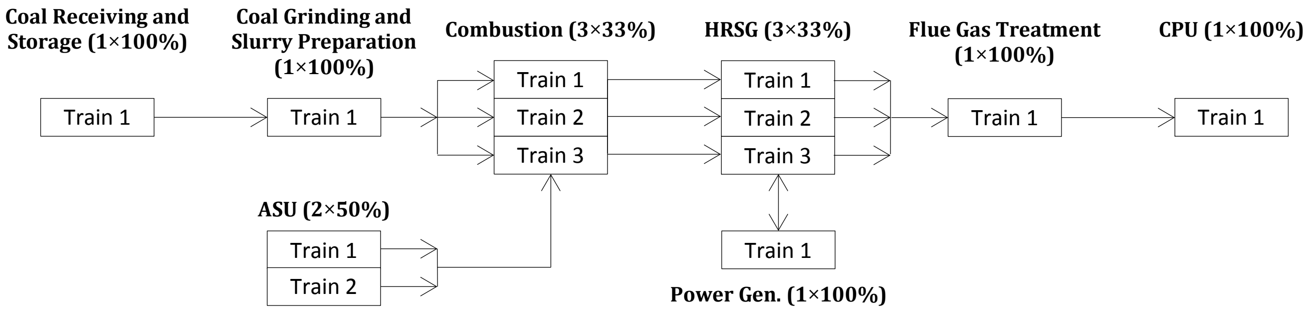

Figure 1 | Train arrangement of the proposed plant

2.6. Plant Capacity Factor

Plant Capacity Factor of 85% is used for discounted cash flow and financial analysis.

2.7. Sparing Philosophy

The only redundancy considered in the design of the plant is related to rotary equipment (1 service + 1 standby) unless otherwise is noted (e.g., HP BFW Pump: 3×50%).

2.8. Selected Case Definition

Three flue gas recycling schemes are considered during process optimization effort of this study: dry recycling scheme, wet recycling scheme, and dual recycling scheme. The optimization analysis concluded that the wet-recycling configuration offers the highest overall plant efficiency and that the optimal operating pressure for this configuration is 29 bar. For more information regarding these schemes refer to Chapter 3, Process Optimization.

This optimal configuration with optimized operating conditions (called proposed process or proposed plant hereafter), was the focus of the technical, economic, and environmental assessment the results of which are presented in this report.

3. Process Optimization

A multi-variable gradient-based optimization is performed for the pressurized Oxy-Coal Combustion (OCC) process with a Direct Contact Column (DCC); three different flue gas recycling configurations are considered, wet recycling, dry recycling, and dual recycling. Section 3.1 includes a brief introduction to oxy-combustion with references to the previous works done by MIT on optimization of pressurized oxy- coal combustion. In Section 3.2, the three different configurations are presented pointing to the characteristics of each. Section 3.3 presents the considered flowsheet and the model description. Then, the optimization formulation is presented in Section 3.4, specifying the objective, the variables, and the constraints. In Section 3.5, the results of the most promising process configuration, wet recycling scheme, are presented in details, and the behavior of the key optimization variables is discussed. The dry recycling and dual recycling configurations are briefly discussed with an explanation of why these cycles are inferior to the wet recycling configuration for the current design basis. Section 3.6, presents the major recent findings regarding process flexibility and opportunities for further heat integration of the OCC process.

3.1. Background

Pressurized Oxy-Coal Combustion (OCC), as one of the most promising technologies for capturing CO_2 from utility-scale power generation plants, has the additional advantage of relying on coal as a low cost feedstock, and compare to other competing technologies requires less alterations to the base process from which it is derived from, the conventional coal fired power plant [6, 7, 8, 9, 10, 11, 12].

In a general OCC the flue gas is predominantly formed of carbon dioxide and water vapor, and the latter can be separated by condensation. The thermal energy from cooling the flue gas and condensing the water vapor can be recovered into the low temperature section of the power cycle [7, 8, 10, 11, 12, 13]. Increasing the flue gas pressure, as done in pressurized OCC, increases the vapor dew point temperature which allows for the condensation to occur at a higher temperature, increasing both the amount of recovered thermal energy and its quality, thus increasing the process efficiency. Pressurizing the combustion process increases the compression requirements of the air separation and oxygen delivery process while reducing those for the carbon capture and purification processes, but also contributes in increasing the pressure losses and irreversibilities within the flue gas loop; the tradeoffs signify a presence of an optimum operation. The optimization effort in this project is based on previous work performed at MIT, a summary of which is included in Appendix A.

3.2. Considered Recycling Configurations

Three different configurations are considered for the recycling of the flue gas for the temperature control processes in the Combustor and the inlet of the HRSG, and are depicted in Figure 2. Optimization and cost analysis are performed only on the most promising process for the design basis, which is the wet recycling.

As the name suggests, wet recycling is recycling of the flue gas before thermal recovery and water condensation. The flue gas is recycled immediately after it exits the HRSG. A large portion of relatively hot flue gas is recycled which demands higher power requirements from the recycling fans compared to the other configurations. The non-recycled flue gas, which has a relatively small flowrate, proceeds to the DCC.

In dry recycling, the flue gas is recycled from downstream of the DCC, after a significant amount of water is condensed. In order to prevent ash agglomeration and undesired water condensation in the recycling pipes, the recycled flue gas temperature is assumed to be increased by 60˚C by heat exchange with the flue gas entering the DCC. This configuration has the largest amount of flue gas entering the DCC, but since the recycled flue gas is relatively cool and has a smaller thermal capacity because it is dry, it requires the least power from the recycling fans. The advantage of this configuration is a decrease in the recycling power requirements. The first disadvantage is having a larger thermal energy transferred at the low temperature section rather than at the high temperature section of the Rankine cycle as explained in Section 3.2.1; this increases exergy loss of the power generation. The second disadvantage is that it requires a larger DCC unit to handle the large amount of flue gas, most of which is recycled back after water condensation. The dry recycling configuration has some concerns regarding the primary recycling stream to the combustor; the Isotherm® combustor requires a significant amount of recycling to insure adequate mixing of the gasses in the combustor to support the flameless combustion process [9].

In dual recycling, the primary recycling to the combustor is in the wet recycling form, while the secondary recycling to the HRSG is in the dry recycling form. This process is a compromise between the above two. It allows for a reduction in the power requirements compared to the wet recycling while respecting the high recycling requirements to the combustor. Compared to the dry recycling, the wet recycling requires a smaller DCC unit, and maintains a larger portion of thermal energy transfer to the Rankine cycle at the high temperature section.

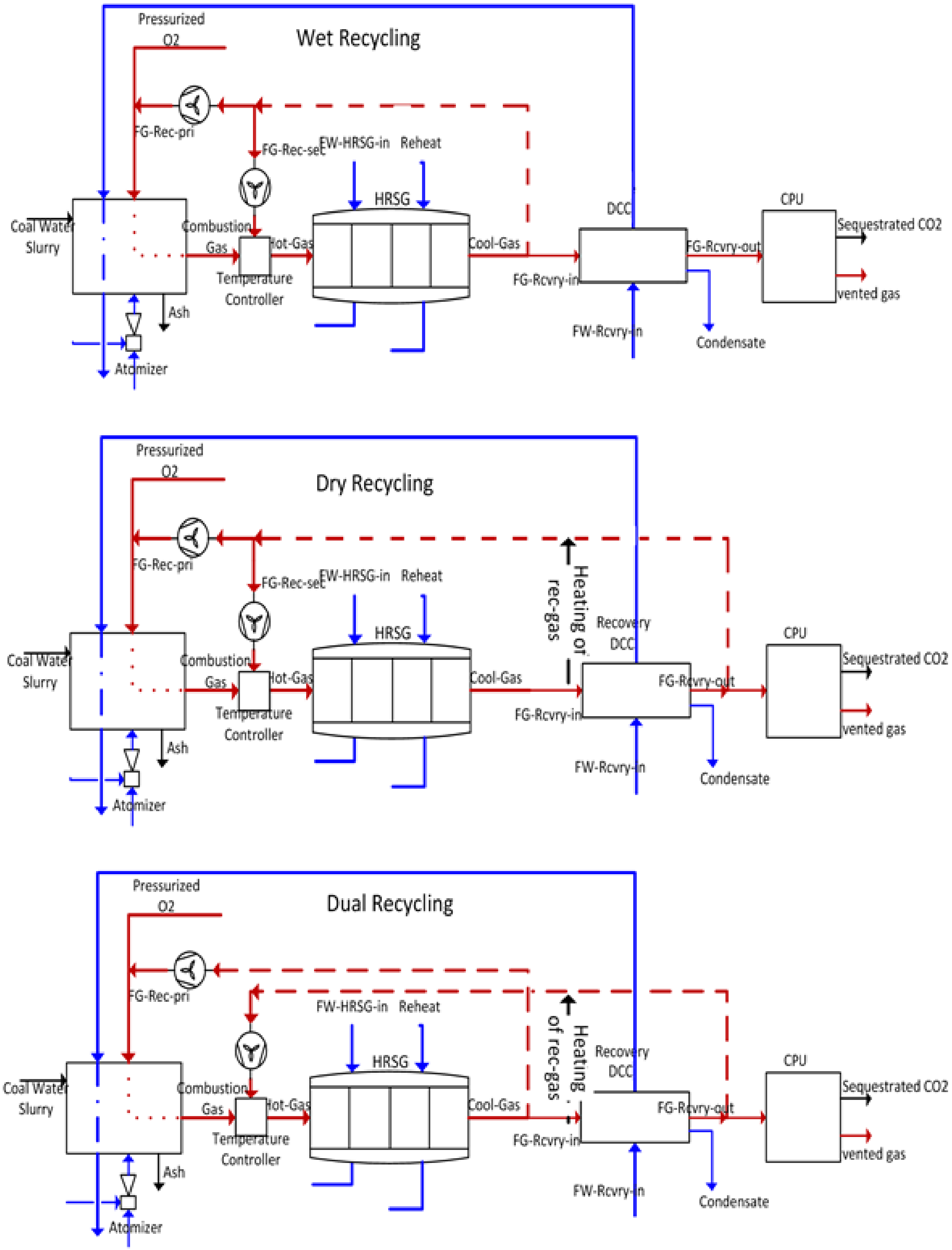

Figure 2 | Wet, dry, and dual recycling configurations

3.2.1. Wet Recycling Is More Attractive

In wet recycling, only a small amount of flue gas proceeds to the DCC for thermal recovery into the low temperature section of the Rankine cycle. The majority of the flue gas is recycled and maintained at the high temperature section, allowing for a high availability of the transferred thermal energy. In dry recycling, all of the flue gas proceeds to the DCC, signifying that a larger portion of the thermal energy is being transferred at the low temperature section of the Rankine cycle instead of the high temperature section. Moreover, it is required to cool and dry the flue gas in the DCC; therefore, the recirculating water of the DCC acquires a relatively large flowrate.

The large flowrate of the recirculating water limits its temperature increase despite the large amount of thermal energy transfer compared to the wet recycling. Therefore, the temperature rise of the boiler feedwater entering the DCC-HX is limited by the recirculating water temperature. Due to this limitation, dry recycling does not take full advantage of the large amount of thermal energy available at the DCC. The deaerator operating pressure and the bleed extraction pressures decrease because the feedwater is not adequately heated. Now lower extraction pressures for the bleeds of the FWHs result in a smaller temperature rise of the feedwater heading to the boiler, which leads to a reduction in the process efficiency by realizing any one of the two following behaviors. First, the lower temperature of the feedwater entering the HRSG results in a lower average temperature of thermal energy transfer at the HRSG increasing exergy destruction. Second in the HRSG, the flue gas of the fixed combustion process is required to heat the feedwater across a larger temperature range, thus the amount of the feedwater allowed to flow in the HRSG decreases (by first law analysis), resulting in a smaller amount of steam generated and expanded through the turbines reducing the power output.

The lower recycling fans power requirements of the dry recycling are not sufficient to overturn the decrease in efficiency due to the decrease in availability of the dry recycling compared to the wet recycling. This is mainly because pressure losses in the modeling and design criteria are considered as a fixed fraction of the operating pressure irrespective of the amount and characteristics of the flue gas, and because in dry recycling the recycled flue gas is heated before entering the recycling fans in order to prevent ash agglomeration and any potential water or acid condensation. Due to the same reasons, the dual recycling does not have an overall advantage over the wet recycling.

The wet recycling process is also more attractive from the cost perspective, by requiring the smallest DCC unit. Therefore, the wet recycle case has been optimized and analyzed in detail in this report. Results of the dry recycling are presented and discussed briefly.

3.3. Wet-recycling Process Configuration

This section describes the process configuration used for the structural and continuous optimization of pressurized oxy-coal combustion. Structural optimization refers to finding the best flue gas recycle configuration and continuous optimization implies finding the optimum operating conditions for each recycling configuration.

Figure 3 shows the process configuration of the wet recycling configuration showing different units, ASU, combustor, the Rankine cycle, DCC, and CPU. The variables and constraints are marked and described further in Section 3.4. Oxygen is separated from air in ASU and provided at an elevated pressure to the combustor as an oxidizer. The combustor is based on the ISOTHERM PWR® technology [9], patented by ITEA [14, 15, 16]. Prior to combustion, the pressurized oxygen is mixed with the primary recycled flue gas stream (FG-Rec-pri) to control the combustion temperature to 1450˚C.

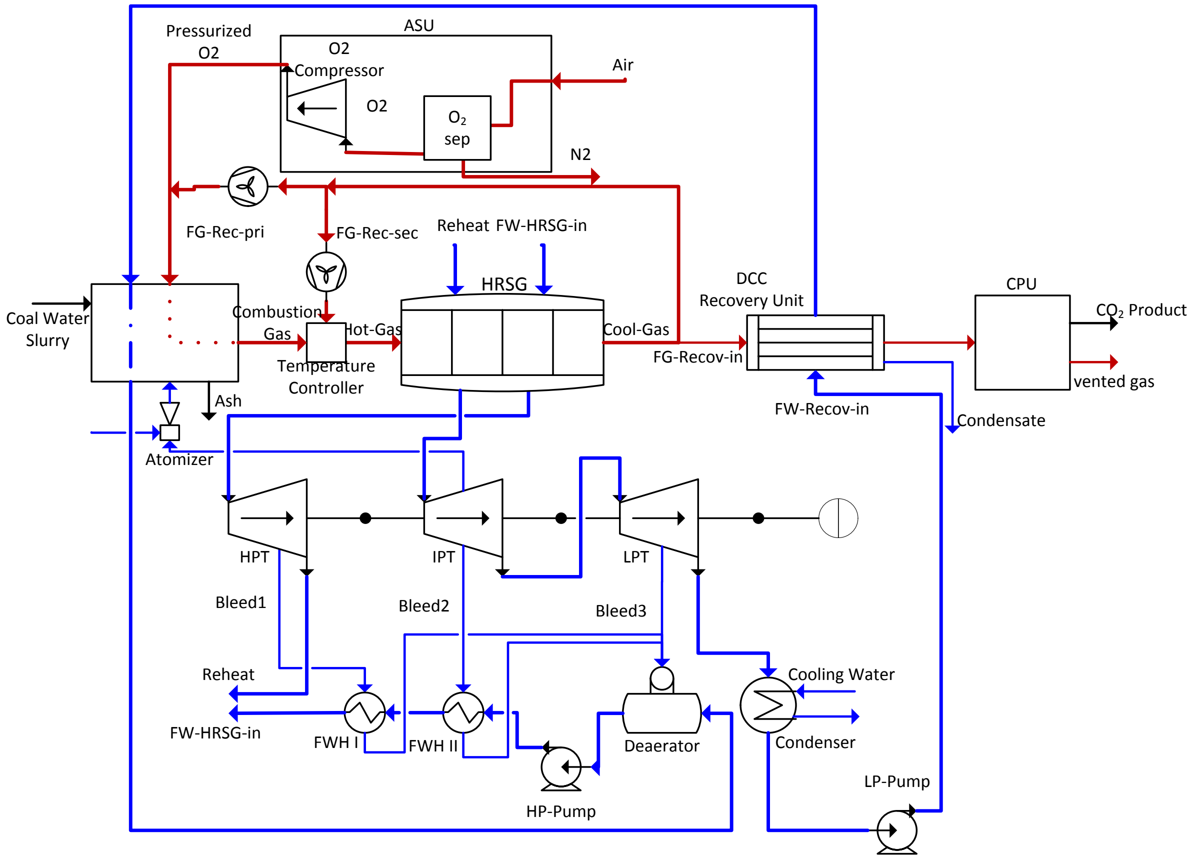

Figure 3 | Wet recycle configuration flowsheet

Combustion gas (Comb-Gas) exiting the Combustor is mixed with a secondary recycling stream (FG-Rec-sec), forming Hot-Gas, before entering the Heat Recovery Steam Generator (HRSG) to maintain a temperature of 800˚C specified due to constraints posed by materials of construction of the HRSG. The HRSG is based on a proprietary ITEA/Ansaldo Caldaie design, developed with the support of ENEL. The HRSG is the main means of heat transfer from the flue gas to the Rankine cycle’s working fluid (water).

Upstream of the DCC, the flue gas temperature must remain above the acid condensation temperature. Sulfur and nitric oxides resulting from the combustion of coal cause damage (corrosion) to the material and components if they condense outside the DCC. Therefore, constraints on the combustion inlet gas, Cool-Gas, and on the feedwater entering the HRSG, FW-HRSG-in, are placed with safety margins to avoid droplet condensation in the flue gas and condensation on the feedwater tubes.

In the depicted wet recycling configuration, a large fraction of the Cool-Gas exiting the HRSG is recirculated for the temperature control processes utilizing fans that compensate for the flue gas pressure drops. The flue gas pressure losses occur mainly in the HRSG and the recirculating pipes. The non-recycled fraction of Cool-Gas, in the wet recycling configuration, proceeds to the DCC where moisture and some of the acid components are condensed from the flue gas before entering the CPU. Moreover, low-quality heat (latent heat at approximately 40˚C to 220C˚ and sensible heat at up to 310˚C), is recovered by cooling the flue gas and condensing the water vapor. FG-DCSC-out exits the DCC and proceeds to the CPU where it is further purified, and compressed and then pumped to 150 bar.

The power cycle considered for this study is a supercritical, single reheat, regenerative Rankine cycle. Only high pressure feedwater heaters (FWHs) are utilized for regeneration; the thermal recovery from the DCC and the thermal energy absorbed from the combustors (by cooling burner assembly) reduce the benefits of adding low pressure FWHs preceding the deaerator. Since the flue gas exiting the DCC is required to have low moisture content then its temperature has to reach a relatively low value (41˚C is assumed). Therefore, the quality of the recovered thermal energy is that of a relatively cool source. The fact that the thermal energy quality is relatively low necessitates placing the DCC before the deaerator and the FWHs. The steam expansion line is accurately represented by three turbine stages and includes the stages' steam leaks and connections' pressure losses. Four extractions are required from the expansion line, one for each bleed (two closed FWHs and one open FWH or deaerator) and one for the combustor's atomizer stream.

Figure 4 shows the DCC flowsheet with the variables and constraints marked to be described in Section 3.4. Flue gas entering the Contact Column is cooled by the recirculating water sprayed from the top of the column, RW-Sep-in. The recirculating water exits the Contact Column at an elevated temperature after absorbing the condensed water and acids. Lime is added to the RW-Sep-out to control the liquid's pH. The pump is used to compensate for pressure losses in the process condensate recycle loop. In DCC-HX, thermal energy is transferred from the process condensate to the feedwater of the Rankine cycle; this thermal energy is the recovered thermal energy from the flue gas. Then, the splitter is used to reject the extra process condensate. Cooling water is utilized to decrease the temperature of the recirculating water during some operation, making it more effective in decreasing the temperature of and drying the flue gas in the Contact Column.

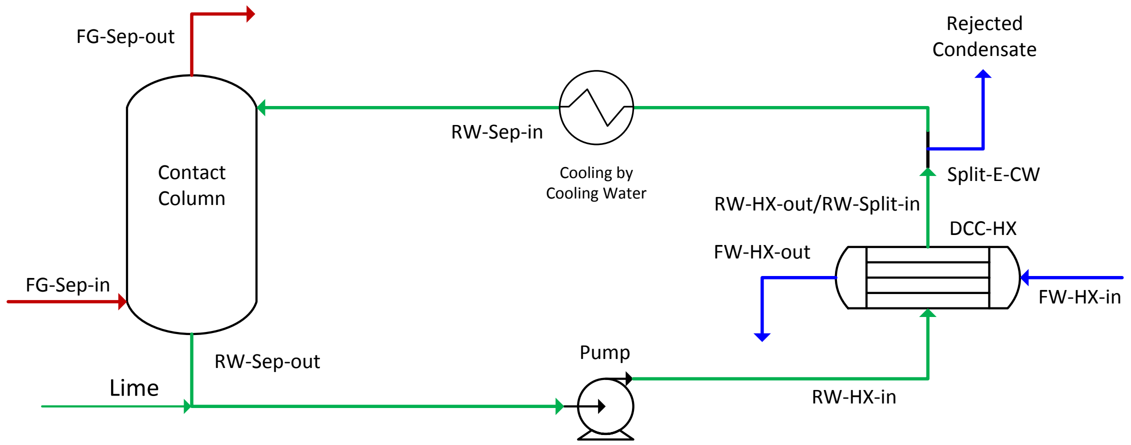

Figure 4 | Direct contact condenser unit flowsheet

3.4. Formulation of Optimization

The importance of simultaneous multi-variable optimization is illustrated in [11, 12], where significant improvements are obtained compared to a single variable sensitivity analysis. Therefore, for the DCC pressurized OCC cycle a similar methodology is applied. Optimization is performed within Aspen Plus using the built-in SQP optimizer. Multi-start, similar to that in [11, 12] is performed here to increase the probability of finding the global optimum, and disregard local-sub-optimal solutions.

3.4.1. Objective Function

The objective considered is to maximize the thermal efficiency of the cycle. The fuel flowrate and specifications are fixed; therefore, the objective is equivalent to maximizing the net power output. The net power is equal to the total power produced from the Rankine cycle minus the power consumption of the pumps, the recirculating fans, the ASU, and the CPU.

3.4.2. Optimization Variables

The decision variables used for optimization are presented in Table 5. The combustion operating pressure, P_{comb}, specifies the flue gas pressure and directly affects the amount of thermal recovery and the pressure losses. $\dot{m}_{FW-main}$, is the amount of feedwater recirculating in the Rankine cycle.

The bleeds extraction pressures, $P_{BLD\#}$, and the bleeds flowrates, $\dot{m}_{BLD\#}$, and the FHWs_I\&II duties, $\dot{Q}_{FWHI\&2}$, are variables associated with the Rankine cycle's regeneration section, and specify the amount and range of the feedwater preheating prior to its entry to the HRSG. Moreover, although not represented in the table, the optimization decides whether each bleed should extraction steam from the high pressure, intermediate pressure or low pressure turbines, and these are integer valued variables.

Table 5 | Optimization decision variables

Decision Variable	Notes
Combustor pressure	Optimized (range 2-70 bar)
Boiler feed water flowrate, $\dot{m}_{FW-main}$	$MITA_{HRSG}$ controlled
FWH I bleed pressure, P_{BLD1}	
FWH II bleed pressure, P_{BLD2}	
Deaerator bleed pressure, P_{BLD3}	
Deaerator bleed flowrate, $\dot{m}_{BLD3}$	
FWH I bleed flowrate, $\dot{m}_{BLD1}$	
FWH II bleed flowrate, $\dot{m}_{BLD2}$	
FWH I duty, $\dot{Q}_{FWH1}$	
FWH II duty, $\dot{Q}_{FWH2}$	
Recirculating water flowrate in DCC, $\dot{m}_{REC-Sep-in}$	
DCC-HX duty, $\dot{Q}_{DCC-HX}$	

Within the DCC, there are two variables: thermal energy transferred in DCC-HX, $\dot{Q}_{DCC-HX}$ and recirculating water flowrate, $\dot{m}_{RW-Sep-in}$. These two variables essentially determine how much thermal energy is transferred from the flue gas into the recirculating water stream and eventually into the working fluid of the Rankine cycle. Table 5 shows the optimization variables as they appear in Figure 3 and Figure 4.

The optimization variables are all independent, for example the bleeds' extraction pressures, $P_{BLD\#}$, and flowrates, $\dot{m}_{BLD\#}$, can be manipulated separately. In contrast, the temperature of the bleed is not independent of the extraction pressure and therefore not considered as a variable. Similarly, the duty transfer in the feedwater heaters, $\dot{Q}_{FWHs}$, are independent variables that define the regeneration from the bleeds to the feedwater. The recirculating water flowrate in the DCC, $\dot{m}_{Rec-Sep-in}$, and the duty transfer in the DCC-HX, $\dot{Q}_{DCC-HX}$, are optimization variables but not the temperature of the recirculating water because it is defined by the recirculation flowrate, the flue gas conditions entering the contact column especially the flue gas pressure, and the amount of thermal energy transferred in the DCC-HX.

3.4.3. Optimization Constraints

Constraints define the allowable limits of operation. The limits are dictated by physical, practical, or economic considerations. Table 6 states the constraints applied during optimization as they appear in Figure 3 and Figure 4.

Table 6 | Optimization constraints

Constraint	Value	Notes
Temperature approach of HRSG	≥5.56 ˚C	
Adiabatic operation of deaerator (Q_{DA})	0	Vapor fraction of 0.5% in deaerator is specified to purge inert gases
Temperature approach of FWH I	≥5.56˚C	
Temperature approach of FWH II	≥5.56˚C	
Flue gas temperature at DCC exit	≤ 41˚C	Flue gas temperature is controlled to ensure that relatively dry and cool flue gas leaves DCC
Temperature approach in DCC heat exchanger	≥5.56 ˚C	
Combustion Temperature	1,450˚C	
HRSG inlet temperature	800 ˚C	
Flue gas temperature at HRSG exit[†]		At least 20˚C above acid condensation temperature
Combustor gas feed temperature[†]		At least 20˚C above acid condensation temperature
Boiler feed water temperature on HRSG tube side[†]		At least 5˚C above acid condensation temperature

[†] These constraints are inactive since the high-pressure acid dew point cannot be reliably and consistently calculated. For the final optimized solution the existing pilot plant data (at lower pressures) in combination with engineering judgment were used to ensure flue gas temperature remains above acid dew point except in DCC.

3.4.4. Constant Parameters

To comply with the guidelines provided with DOE, some variables were excluded from optimization. Parameters like the temperature and pressure of the feedwater exiting the HRSG/entering the turbine, and more importantly, the reheat extraction pressure and the reheat delivery temperature are not incorporated as optimization variables. Preliminary results show that the current guidelines for the reheat conditions are for a fact suboptimal for the pressurized OCC process and potential advantages are expected with optimizing the reheat specification. Moreover, the purity of the oxygen stream resulting from the ASU affects efficiency and capital cost significantly and thus would be interesting to optimize.

Active Constraint Optimization

This study utilizes recent methodological proposals in [11, 17, 2] where it is proven analytically and numerically that optimal operating conditions of the cycle are obtained at some active constraints. More specifically, [11] proves that heat exchangers need to operate at the MITA specification for optimal performance. A more dedicated proof for the optimum operation of regenerative Rankine cycles is presented in [17, 2] along with elaborate numerical case studies. The optimum regeneration necessitates the existence of a double-pinch, *i.e.*, MITA encountered at the onset of the bleed condensation and simultaneously at the drain outlet. Therefore, variables can be manipulated at the simulation level to achieve the desired value of the constraint. The advantages are numerous including reducing violations and fatal errors in the simulation, constraint violations in the optimization, and the size of the optimization problem. More importantly, the procedure partially avoids convergence to suboptimal local optima or saddle points by guaranteeing that the manipulated variables are set to the values that are obtained at the global optima. Moreover, the procedure developed is based on explicit equations and assignments eliminating the need for a spatially distributed model further reducing computational expense [17, 2].

The variables and constraints coupled in this study are:

- $MITA_{DCC-HX}$ / $\dot{Q}_{DCC-HX}$: The allowed minimum internal temperature approach on the DCC-HX is achieved by the amount of thermal energy transfer in the DCC-HX
- $MITA_{HRSG}/\dot{m}_{FW-main}$
- Double-pinch_FWH(1 & 2) / $\dot{Q}_{FWH1\&2}$ and $\dot{m}_{BLD1\&2}$: The double pinch condition is made up of two simultaneous pinch occurrences requiring the manipulation of two variables. Therefore, both the duty transfer within each closed FWH and the flowrate of the respective bleeds are defined in terms of the bleed extraction pressure according to the following equations [17, 2]:
    - $\dot{m}_{BLD} = \dot{m}_{FW} \frac{h^l\left(T^{sat}(P_{BLD}) - \Delta_{MITA_{FWH}}T, P_{FW}\right) - h^l(T_{FW,i}, P_{FW})}{h^{g,sat}(P_{BLD}) - h^l\left(T_{FW,i} + \Delta_{MITA_{FWH}}T, P_{BLD}\right)}$
    - $\dot{Q}_{FWH} = \dot{m}_{BLD}\left(h_T(P_{BLD}) - h\left(T_{FW,i} + \Delta_{MITA_{FWH}}T, P_{BLD}\right)\right)$

  Note that since the expressions are explicit in $\dot{m}_{BLD}$ and $\dot{Q}_{FWH}$ there is a unique double pinch for a given extraction pressure P_{BLD}.
- $q_{Dearator}$ / $\dot{m}_{BLD3}$: The constraint of 0.5% saturated vapor in the deaerator is satisfied by the deaerator bleed flowrate as proven in [17, 2].

3.5. Optimization Results

Simultaneous multi-variable optimization of the DCC pressurized OCC process with a wet recycling configuration is performed similar to [11, 12]. To better assess the effect of the operating pressure and the importance of thermal recovery, a pressure parametric optimization is performed and results are plotted in Figure 5. The figure is constructed in the following manner, starting from the optimum found at 29 bar, operating pressure is incrementally varied while optimizing for all variables except pressure.

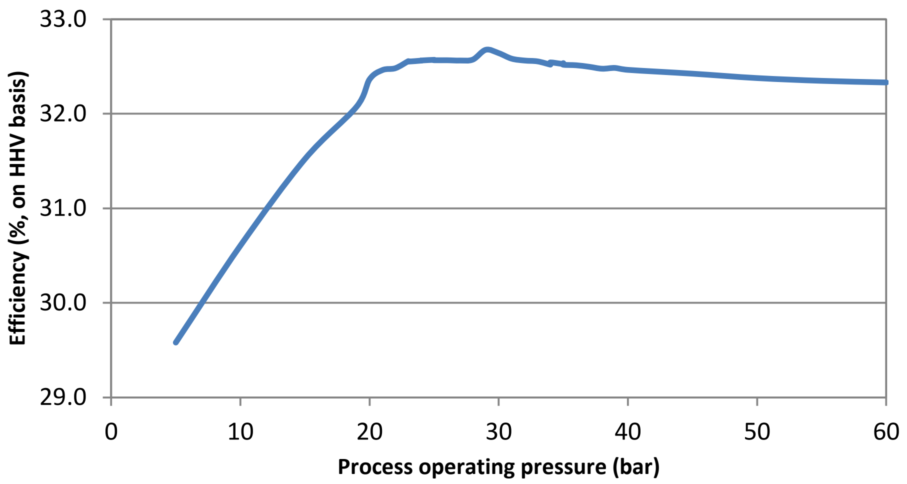

Figure 5 | Plant net efficiency (on HHV basis) of the wet recycling configuration as a function of operating pressure

The efficiency plateau from 20 to 60 bar seen in Figure 5 is quite interesting and can be explained by the results shown in Figure 6, which plots the thermal energy transferred to the Rankine cycle feedwater at the DCC-HX, and the thermal energy that is lost form the recirculating water to the cooling water before the former enters the DCC. In the range of 20 bar and higher, the recovery section can operate at its optimum; the theoretical optimum flowrate of the recirculating water, which is the flowrate that results in a balanced DCC-HX [12], is capable of sufficiently cooling and drying the flue gas to the maximum constrained temperature of 41˚C. This results in a relatively high increase in temperature of the recycling water as it exits the DCC and enters the DCC-HX. Now since for this range of operating pressure, the DCC-HX is balanced, the feedwater temperature can increase significantly, very close to the recirculating water inlet temperature, and the recirculating water temperature can decrease significantly, very close to the cooling water inlet temperature; the cool recirculating water exiting the DCC-HX ensures that very little amount of thermal energy is transferred/lost with the exit stream or the cooling water.

It is clear from Figure 6 that above 20 bar the thermal recovery is maximum while the cooling water duty is minimum. Increasing the pressure above 20 bar barely increases thermal recovery or decreases thermal losses to cooling water, but increases pressure losses of the flue gas in the HRSG and the recycling pipes; therefore, there is a small decrease in efficiency at pressures higher than 29 bar.

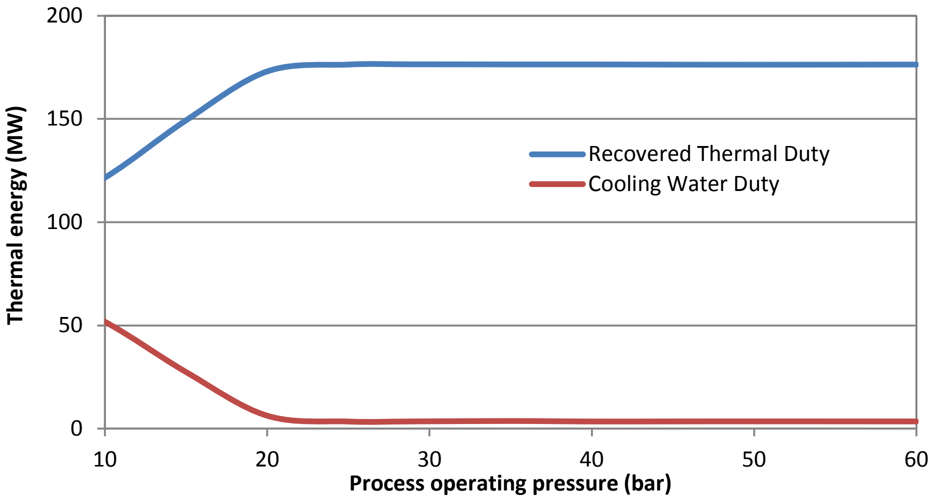

Figure 6 | Recovered thermal energy and cooling water duty in the DCC at different operating pressures

As mentioned at pressure above 20 bar, there is very little increase in the recovered thermal energy. However, the power requirement of the fans, ASU, and CPU play a role in determining the efficiency profile. As pressure increases above 29 bar, the ASU power requirements increase at a larger rate than the power decrease for the CPU and the fans (Figure 7); this also contributes to the slightly decreasing efficiency profile above 29 bar operating conditions.

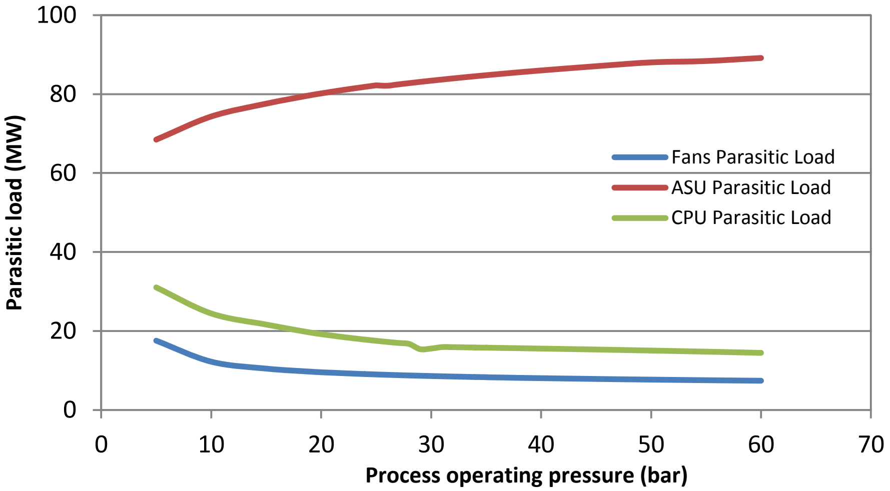

Figure 7 | Distribution of parasitic load at different operation pressures

At pressures lower than the optimum the efficiency decreases at a faster rate and falls rapidly below 20 bar. This behavior is also attributed to the operation of the DCC. At low pressures, the optimum recirculating water flowrate is not capable of sufficiently cooling and drying the flue gas (41˚C); therefore, a higher amount of recirculating water is required. Increasing the flowrate of the recirculating water results in a decreasing the temperature of the recirculating water at the exit of the contact column/inlet of the DCC-HX. This means that the temperature of the feedwater exiting the DCC-HX is also lower than that of the optimum. The DCC-HX is not balanced, where the flowrate of the recirculating water is larger than that of the feedwater, thus the recirculating water exiting the DCC-HX has a relatively high temperature. This unutilized thermal energy in the recirculating water is rejected into the cooling water before entering the contact column. Further decrease in operating pressure requires further increase in the recirculating water which further decreases the recovered thermal energy and increases the cooling water duty, which is exactly what Figure 6 depicts. In conclusion, as pressure decreases, the quality of the thermal energy leaving the flue gas decreases, the quantity of recovered thermal energy in the DCC-HX decreases, and the quantity of thermal energy lost to cooling water increases. Moreover, since the temperature of the feedwater is exiting the DCC and entering the deaerator is relatively low, then the deaerator pressure decreases and/or the deaerator bleed flowrate increases. The thermal recovery section can no longer maintain a high temperature of the feedwater entering the HRSG, which causes further reduction in the availability of the high temperature section. These different interactions explain the fast drop in efficiency for smaller operating pressures.

Results of the dry recycling process show a low efficiency compared to the wet recycling scenario, 25.5% compared to 33.5%. This can be explained as follows. First, the amount of flue gas heading to the recovery section in the dry recycling configuration is significantly larger than that of the wet recycling configuration; therefore, the amount of thermal energy transferred at the high temperature section for the dry recycling is relatively small. The other disadvantage in dry recycling comes from the requirement to have a dry flue gas at the exit of the contact column; this necessitates a much larger flowrate of the recirculating water to reduce the temperature of the flue gas. Due to its large flowrate, the temperature rise of the recirculating water exiting the contact column is relatively low, resulting in a relatively small rise in the temperature of the feedwater at the DCC-HX. Here again, the DCC-HX is not balanced, where the feedwater temperature rise is relatively low, and the recirculating water temperature at the exit is relatively high. The unutilized thermal energy maintained in the recirculating water at the exit of the DCC-HX is eventually lost by the rejected stream and to cooling water. Here too, since the feedwater temperature rise is relatively small, regeneration is incapable of delivering the feedwater at a high temperature to the HRSG, resulting in further reduction of availability of the high temperature section and an overall decrease in efficiency.

It is worth mentioning that the process pressure in dry recycling is favored to be high, above 40 bar, in order to facilitate the condensation of the flue gas in the contact column (the recycled flue gas is dry so the water concentration in the main flue gas stream is low); yet this only slightly enhances the recovery section performance since the required recirculating water flowrate is still relatively large. Also, with higher operating pressures, the pressure losses increase, further affecting the performance of the dry recycling configuration. Moreover, there are potential risks due to condensation/corrosion in dry recycle configuration in the heat exchange process between the hot wet flue gas the cold dry flue gas.

The performance of the dual recycling configuration is in between those of the wet recycling and dry recycling and does not offer any efficiency advantage over wet recycling configuration.

3.6. Other Advantages

3.6.1. Ideal Flexibility to Uncertainties

Recent research studies are performed on the pressurized OCC process in an attempt to characterize its behavior and show its advantages over conventional Rankine cycles and other advances power generation technologies. Thermal power generation faces several uncertainties during operation. For example, coal type and specifications vary significantly from one source to another, or even from different batches of the same source. A process that is optimized for a given coal type would be unattractive if its performance deteriorates or suffers when that specific coal is not economically attainable. With the change of coal type, other input specifications and parameters need to change accordingly, like the oxidizer flowrate, the flowrate of slurry water responsible for transferring the coal into the combustor, and the atomizer stream flowrate, which contribute to the alteration of the cycle's behavior. Other forms of uncertainty are due to the ambient conditions, in particular that the cooling temperature may vary significantly.

In [18] the OCC process is shown to have ideal flexibility to the uncertainties in the input specifications mentioned above; more specifically, changes in inputs can be accommodated without any compromise in the processes' performance compared to a process specifically designed for the new values of the inputs. The study concludes that stochastic optimization is not needed for designing the flexible power plant. In essence, the uncertainties in input conditions and parameters need not to be quantified but merely the range of input conditions needs to be taken into account during design.

Another important disturbance to the process is the variation and uncertainty in the thermal load which is investigated in [18], with an accurate representation of unit operations particularly the turbine expansion line. The turbines operate at constant volumetric fluid flowrate profile which requires changing the turbine inlet pressures and pressure ranges with the change in load. In [19] results show that the process is ideally flexible for variable load due to the characteristics of the thermal recovery section. The performance of the nominal load design when operating at a given part-load matches the maximum performance of the process designed specifically for that part-load. When designing the process specific to a part-load, the turbines are maintained at the nominal load design in order to allow for a full range of load operations. The ideally flexible behavior is owed to the thermal recovery section, which is not affected by the reduction of the pressure ranges of the turbine expansion line with decreasing thermal load. The recovery section always provides adequate preheating to the working fluid. In particular, a unit flowrate of working fluid always receives the same preheating from the flue gas at the thermal recovery section independent of the operating load. This signifies a relatively larger preheating duty with the decrease in the turbine pressure ranges; the recovery section compensates the decrease in the effectiveness of the inflexible regeneration section, and therefore, the OCC process is ideally flexible. This flexibility is in contrast to Rankine cycles without pressurized recovery, wherein the performance significantly deteriorates compared to the nominal operation. Moreover, as the thermal load and the working fluid flowrate decrease, the required turbine pressures fall below the critical pressure, and the working fluid in the HRSG pass through the saturation region. However, this transition occurs at a larger load for Rankine processes without pressurized recovery compared to the pressurized OCC processes which are able to maintain supercritical conditions at smaller thermal loads.

3.6.2. Further heat integration possibilities

There is another possible advantage to the pressurized OCC process by preheating the cold input combustion streams using the recovery section. In the current study, the recovery section is integrated with the low temperature section of the Rankine cycle and therefore providing a low quality thermal energy. However, this process allows for a minor modification where in the DCC a heat exchanger can be utilized to transfer thermal energy into the cooler streams entering the combustor, in particular the coal water slurry mixture. The coal water slurry mixture has a low temperature, around 30˚C. During the combustion process, the coal water slurry, which eventually becomes a part of the flue gas, experience an increase in temperature by means of a high quality source. The modification is to preheat the coal water slurry using the DCC, thus saving on high quality thermal energy and increasing the overall efficiency. This modification is a topic of future work.

4. Plant Description

This chapter provides a description of the entire plant from a process engineering perspective. The narrative features an analysis that explains why key assumptions and design decisions were made in light of the options available.

All materials presented hereafter in this report refer to the optimized plant based on wet-recycle configuration.

This section shows main process/utility areas of the proposed pressurized oxy-coal combustion plant and may be different from actual areas considered in the cost estimation of this report. Process Flow diagram of each processing unit is presented in corresponding sub-section.

Figure 8 depicts the overall block flow diagram of the optimized process based on wet-flue gas recycling configuration.

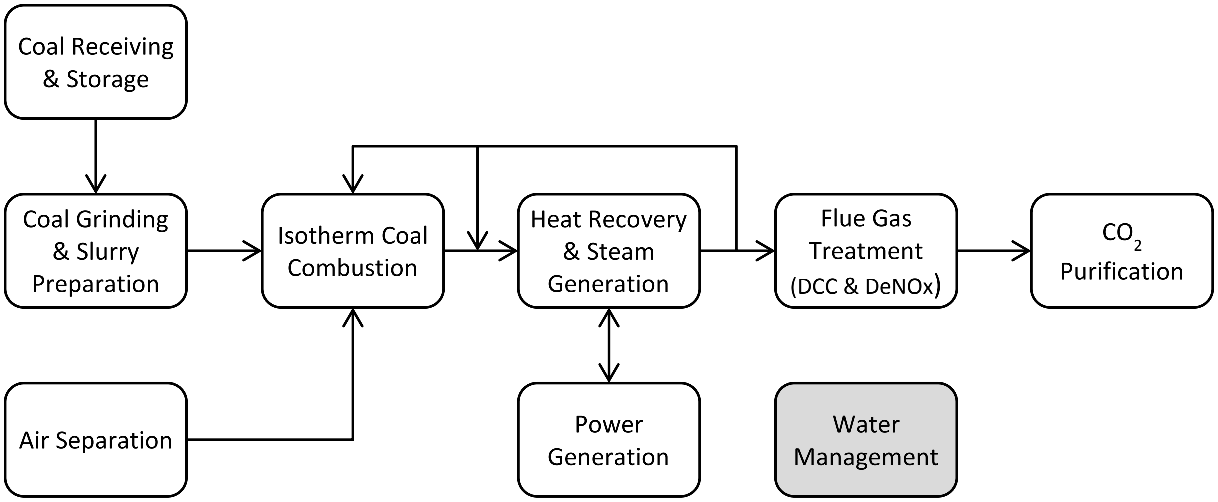

Figure 8 | Plant block flow diagram

4.1. Coal Receiving and Storage Unit

The function of the Coal Receiving and Storage system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to and including the slide gate valves at the outlet of the coal storage silos [20]. The coal receiving and storage unit is not included in the developed process model and assumed to be similar to what is described in the DOE NETL baseline study [20].

4.1.1. Process Description

“The unloading is done by a trestle bottom dumper, which unloads the coal into two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The coal from the feeder is discharged onto a belt conveyor. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile. The reclaimer loads the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a belt conveyor that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm x 0 (1¼" x 0) by the crusher. A conveyor then transfers the coal to a transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of three silos. Two sampling systems are supplied: the as-received sampling system and the as-fired sampling system” [20].

4.2. Air Separation Unit

The purpose of ASU is to provide required oxygen for the Isotherm oxy-combustors. The proposed plant requires two ASU trains (with common cold-box) each producing 3560 tonne/day of contained oxygen at ~30 bar. The cycle chosen for this duty is one in which gaseous oxygen (GOX) is produced by pumping liquid oxygen (LOX) to the required 30 bar pressure and allowing this stream to be boiled against condensing high pressure air. There is no requirement for an oxygen compressor in this configuration. A low purity pumped LOX cycle was chosen, which produces 95% by volume oxygen purity. Other studies [21] have shown that for oxy-combustion plants this is the optimum purity. Even with no air in-leakage, the increase in power required for the ASU to produce 99.5% purity oxygen is greater than the decrease in CPU power consumption achieved by reducing O_2 and raw CO_2 purity [22].

4.2.1. Process Description

To minimize the ASU power requirement, a cycle has been chosen that uses three columns. Figure 9 depicts the assumed process design for ASU unit [22]. This design has similar specific work (kWh/tonne of O_2) as traditional ASUs but provides a safe operation by avoiding O_2 compression unit. Also, availability of oxygen supply can be improved by storing liquid oxygen. No integration with process is considered at this stage (for simplicity).

The proposed air separation unit consists of:

- Compression system;
- Adsorption front-end air purification system; and
- Cold-box containing the separation and the heat exchanger equipment.

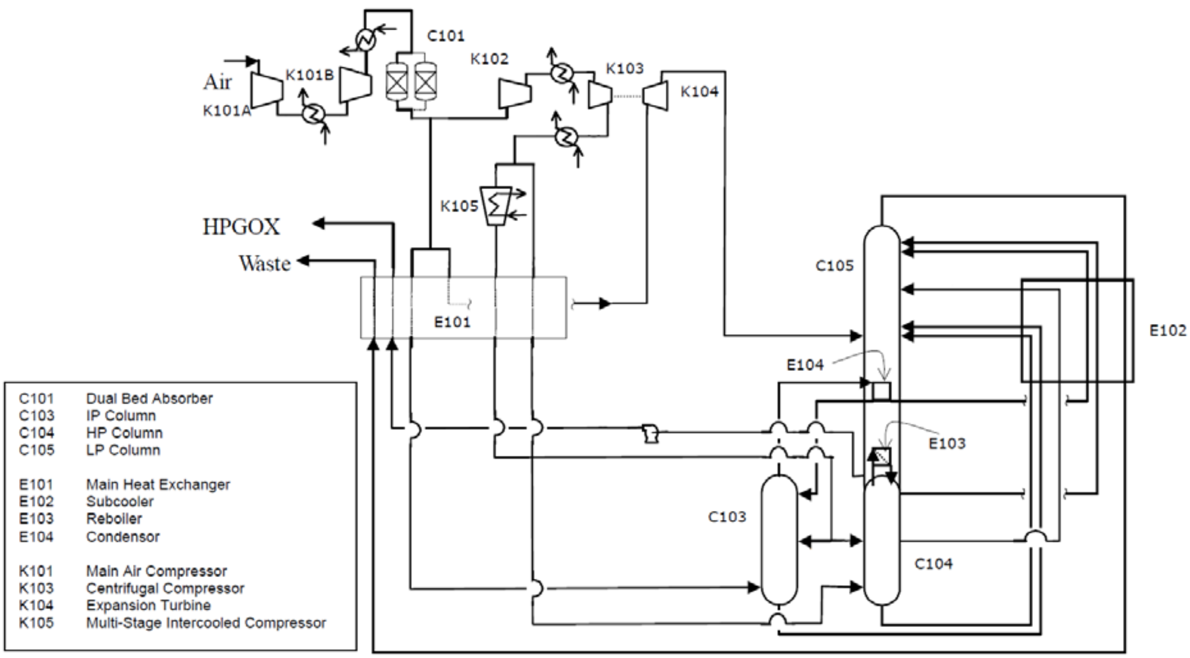

Figure 9 | Process flow diagram of Air separation unit [22]

“The standard double column cycle has a low pressure column (C105) with its reboiler (E103) integrated with the condenser of a high pressure column (C104). The column pressures are set to give a temperature driving force in the reboiler/condenser E103. In this cycle an extra column is added operating at an intermediate pressure (C103). The condenser (E104) for this column also integrates with a reboiler in the low pressure column but at a lower temperature, boiling a liquid stream higher up within the low pressure column which has a lower boiling point. This arrangement minimizes the amount of feed air that must be compressed to the higher pressure of C104, leading to the low power requirement of this process cycle” [22].

This process offers the benefits of high reliability, low maintenance cost and it is simple to install and operate [22]. The detailed description of ASU process considered in this study can be found here [22].

4.3. Coal Grinding and Slurry Preparation Unit

This unit prepares the received and crushed coal according to the feeding requirements of the Isotherm combustors.

4.3.1. Process Description

“The receiving and handling unit ends at the coal silo. Coal is then fed onto a conveyor by vibratory feeders located below each silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. Each hopper outlet discharges onto a weigh feeder, which in turn feeds a rod mill. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is discharged through a trommel screen into the rod mill discharge tank, and then the slurry is pumped to the slurry storage tanks” [20]. The solids concentration of the final slurry is ~63 wt.%.

“The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required depends on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended” [20].

“The equipment in the coal grinding and slurry preparation system is fabricated of materials appropriate for the abrasive environment present in the system. The tanks and agitators are rubber lined. The pumps are either rubber-lined or hardened metal to minimize erosion” [20].

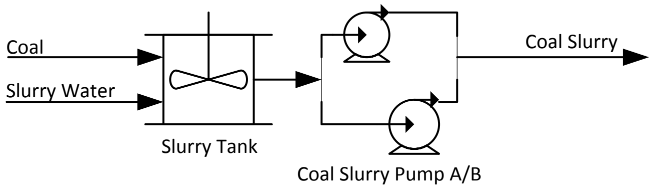

Figure 10 | Process flow diagram of coal grinding and slurry preparation unit

4.4. Isotherm Combustion Unit

The combustion technology considered for this project, flameless Isotherm Combustion, developed by ITEA, is the key enabling technology of the designed pressurized oxy-combustion process. Isotherm combustor is a pressurized, slurry-fed coal combustion unit in which combustion chamber temperature is uniform and controlled by recycling flue-gas.

Vitrified slag separation and handling is part of the combustion unit packaged and delivered by the technology provider.

4.4.1. Process Description

The prepared coal slurry from slurry preparation unit is sent to combustor units. Also, high pressure steam is used to atomize the coal slurry using proprietary burners. A small fraction of BFW is used in burner assemblies for heat management purposes. The proposed process utilizes the ITEA's proprietary Isotherm oxy-combustion reactor in which enough flue-gas is recycled to the combustor to maintain the combustor temperature at 1450˚C. The ash content of coal feedstock is converted to vitrified slag due to high operating temperature of the combustor and extracted by the slag handling system. Another flue gas recycle stream is used to quench the raw flue gas from combustor to 800˚C prior to HRSG unit.

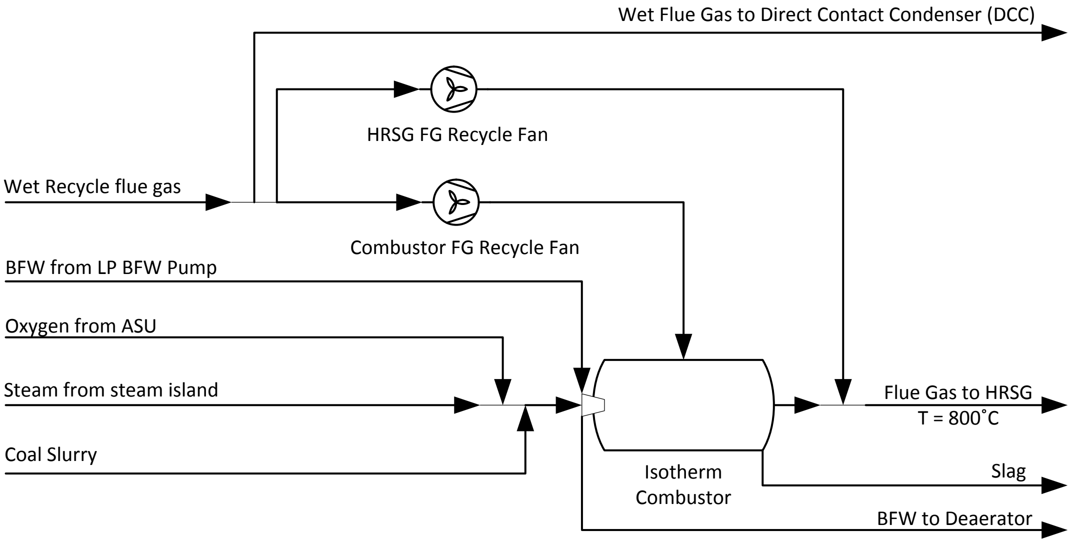

Figure 11 | Process flow diagram of pressurized Isotherm oxy-coal combustion unit

4.5. Heat Recovery and Steam Generation Unit

The steam cycle is designed based on guidelines provided by DOE for supercritical steam cycle in the FOA package. The design also includes a reheat unit to reheat the HP turbine exhaust before it is sent to IP turbine. A simple design of HRSG unit and power island is considered for this study to showcase the advantages of the proposed pressurized oxy-combustion technology over the conventional PC power plants as well as atmospheric oxy-combustion processes. The design of steam cycle and steam turbines can be optimized based on the pressured oxy-combustion characteristics for further efficiency enhancement of the overall power plant.

ENEL has contributed an advance design for a heat recovery steam generator capable of handling inlet gas at 800˚C.

4.5.1. Process Description

Preheated BFW from DCC enters the deaerator. A small flow of this stream is sent to burner assembly of the isotherm combustion units to protect the burner assemblies by removing excess heat from the assembly. Unlike, conventional steam cycles, deaerator in this design operates at relatively high pressures due to the fact that the BFW entering deaerator is already preheated by recovering heat from recycle loop of DCC. IP steam extracted from IP turbine is used in deaerator. The considered steam cycle includes two BFW heaters which preheat the BFW stream from deaerator before being directed to the boiler. The operating conditions of deaerator as well as the BFW heaters are optimized using methods described in the Process Optimization section of this report.

Within the heat recovery sub-unit heat is recovered from flue gas to generate HP steam at 242 bar. The unit also includes a reheat loop where the exhaust from HP turbine is reheated in HRSG to 599˚C.

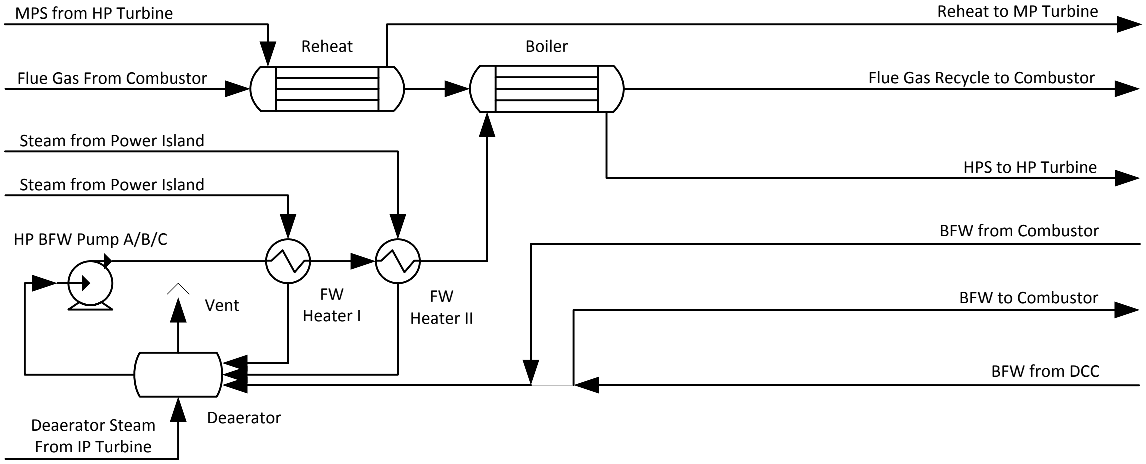

Figure 12 | Process flow diagram of heat recovery and steam generation unit

4.6. Power Island

Power island which is comprised of three steam turbines, HP, IP, and LP, converts the steam generated in HRSG unit to electric power. The power island also provides steam streams at required pressures for operation of other unit operations within the plant such as atomization steam used in Isotherm combustors.

Specifications and assumptions (e.g., isentropic efficiency of turbines) for the steam island are according to guidelines provided by DOE where applicable. The designed system also accounts for losses due to leakages of steam turbines based on DOE guidelines.

The designed power island is a simplified design to present advantages of the proposed oxy-combustion technology and can be improved during the detailed engineering of a project.

4.6.1. Process Description

High pressure steam at 242 bar and 599˚C from HP boiler enters the HP turbine and leaves at 49 bar. The steam leaving the HP turbine is sent to HRSG unit where it is reheated to 599˚C and directed to the IP turbine. Steam leaving the IP turbine at 5.2 bar is sent to the LP turbine which is a condensing turbine. The exhaust from the LP turbine is sent to the water-cooled stream condenser and then pumped and sent to the DCC unit for pre-heating.

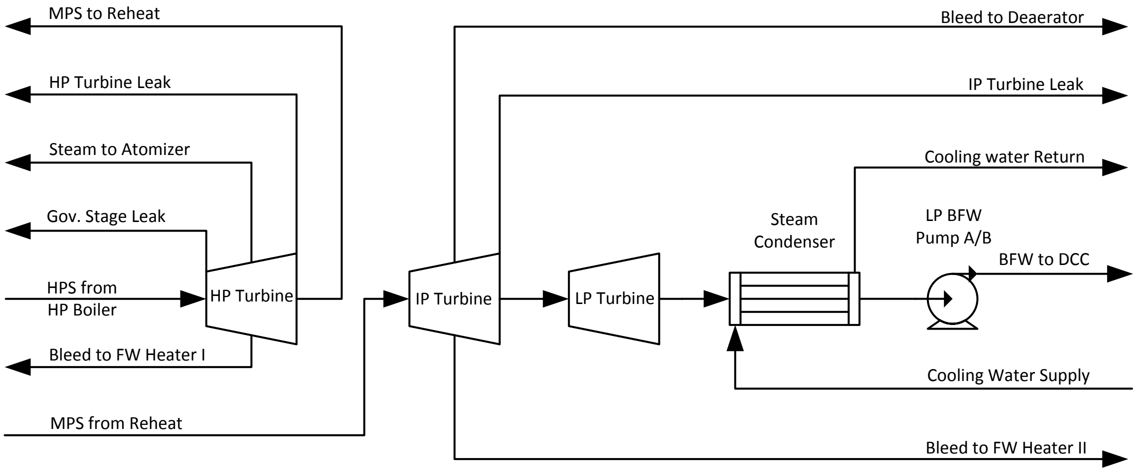

Figure 13 | Process flow diagram of power island

4.7. Flue Gas Treatment Unit

The purpose of the flue gas treatment (FGT) unit is to:

- Recover the low-grade heat remaining in the flue gas after HRSG unit
- Remove particulate matter from the flue gas
- Remove SOx from flue gas by reacting it with lime solution and removal as gypsum.
- Remove of remaining of SOx and NOx components from flue gas.

The designed process of this unit is a combination of different flue gas treatment technologies which individually are commercially available (usually at low pressure) and marketed by various technology providers and engineering firms. These technologies integrated together or even individually at high pressure have no commercial scale experience. The flue gas treatment unit comprises two processing units: Direct Contact Condenser (DCC) and DeNOx unit.

The DCC operates similar to a traditional wet FGD system to put the pressurized flue gas in contact with the lime solution to remove SO_2 from flue gas. During this process the remaining heat (low-grade) as well as particulate matter is removed from flue gas. Design and operation of this unit is quite important since removal of SO_2, particulate and heat from flue gas take place simultaneously in this unit.

4.7.1. Process Description

The remaining of flue gas from HRSG unit (which is not recycled back to combustor/HRSG units) is sent to DCC where it comes in direct contact with lime solution. It's assumed that lime solution removes 98% of SOx components in the flue gas based on the DOE guidelines for FGD. At the same time, lime solution recovers the remaining sensible heat and latent heat from flue gas. The process is designed such that the flue gas leaves DCC at ~40˚C (by manipulating the recycle rate of lime solution); at this temperature and operating pressure of the system the moisture content of the leaving flue gas stream is <1 mol%. The hot lime solution leaving the DCC enters the residence tank with enough residence time to allow for formation of gypsum. Also, fresh lime solution is added such that the pH of the system is maintained in the 8-10 range. Solids, including gypsum, other salts and fly ash settle in the residence tank and are removed for waste treatment. This configuration has been developed for conceptual process evaluation purposes; a detailed design of the required equipment has not been executed as part of this project. The hot lime solution is pumped and sent to a heat exchanger where it exchanges heat with BFW. After heat recovery, a fraction of the lime solution is further cooled using cooling water in process condensate cooler and sent to DCC. The remaining of the solution is sent to water treatment unit for neutralization, recovery and reuse.

The flue gas leaving DCC is sent to DeNOx column where the remaining of SOx along with NOx components is removed in a process similar to the traditional lead-chamber process. To increase the reaction rate for removal of acid components (SOx and NOx) the DeNOx column operates at low temperature which is provided using cooling water cooler. The recycle loop pH is controlled by manipulating the recycle as well as water make-up rate. The bleed stream from recycle loop is sent to water treatment system for neutralization and water recovery.

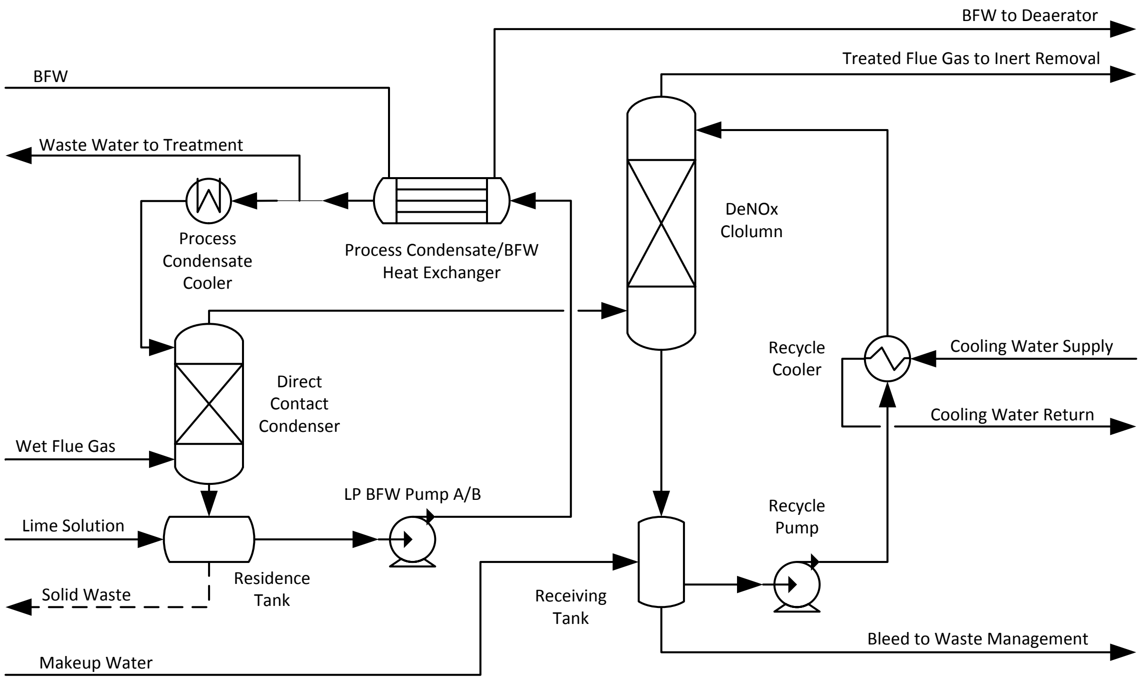

Figure 14 | Process flow diagram of flue gas treatment unit

4.8. CO_2 Purification Unit

Treated flue gas, which is mostly CO_2, is send to CO_2 Purification Unit (CPU). The CPU removes inert gases such as N_2, O_2, and Ar from the CO_2 stream using cryogenic separation techniques and delivers a high purity CO_2 product which meets the purity requirement for EOR applications. The considered process configuration delivers CO_2 with purity > 99.8 mol% and oxygen content of < 100 ppmv. For producing CO_2 product with higher/lower purity level other process configurations can be used which can have different parasitic loads. Cryogenic purification of CO_2 rich streams is a commercial technology which is currently marketed by technology providers such as Air Products and Chemicals.

4.8.1. Process Description

The flue gas leaving DeNOx column is send to regenerative dryer beds to remove the remaining moisture in the gas. The consumed dryer-bed is regenerated using steam while the other bed is in use.

Then, flue gas is sent to an activated-carbon bed to remove the remaining mercury from the gas before it is send to cryogenic separation section. The flue gas is cooled against the product CO_2 in a heat exchanger which acts as the reboiler of the distillation column downstream. The chilled flue gas is mixed with the vapor product from the distillation column and sent to a series of flash units to create refrigeration and separate various inert components from the flue gas. The heat integration of this process has to be carefully examined to achieve high efficiency of the CO_2 purification process (Figure 15).

The high-purity CO_2 streams (liquid products of Flash I & II) are vaporized to provide the required refrigeration duty for the process and then are compressed in a multi-stage internally-cooled compressor. The compressed high-purity CO_2 stream is then cooled and sent to the cryogenic distillation column where most of the remaining of the inert components, are removed to produce high-purity CO_2 stream. The CO_2 product is subsequently pumped to the required delivery pressure of 150 bar.

The vapor stream from the last flash unit is preheated and sent to a turbo expander to generate electricity. Alternatively, since a considerable fraction of this stream is O_2, it can be pressurized and recycled back to the combustor units; this option has not been considered for this study.

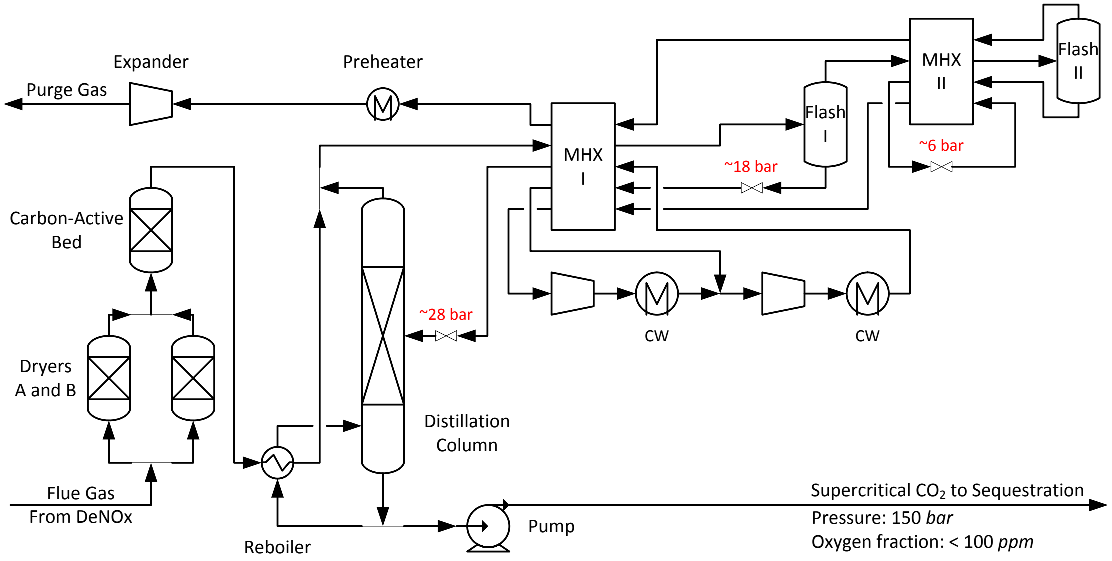

Figure 15 | Process flow diagram of CO_2 purification unit

5. Components Descriptions

Table 7 provides the information required by the DOE in the FOA, Attachment A of [1], for each major component or subsystem of the proposed plant.

Table 7 | Description of key plant units/components

Component or Subsystem	Technology Type	Basis for Design and Performance			Operating Condition				Assumed or specified Performance	Calculated Performance	Contaminant removed	Assumptions regarding Anticipated Application Issues	Technology Readiness
					Inlet		Outlet						
		Self-defined or Generic	Vendor Data /Commercial Design	Vendor Data / Future Design	Temperature (˚C)	Pressure (bar)	Temperature (˚C)	Pressure (bar)					
Coal Receiving and Storage	Solids Handling		X		Ambient	Atmospheric	Ambient	Atmospheric					Mature/ Commercial
Coal Receiving and Storage	Solids Handling		X		Ambient	Atmospheric	Ambient	Atmospheric					Mature/ Commercial
Air Separation Unit	Cryogenic Distillation		X		Ambient	Atmospheric	0	29		281 kwh/tonne O_2	Mainly N_2/Ar (O_2 purity: 95 vol%)	None	Mature/ Commercial
Isotherm Combustor	Pressurized Oxy-Combustion			X (ITEA)	~200	29	1,450	29	100% carbon conversion	Chemical equilibrium	-	- Complete combustion	Pilot
HRSG	Pressurized Heat Recovery and Steam Generation			X	800	29	320	28.1	Heat Loss = 0.5% of duty (due to smaller size)		-	- Resistant to minor acid condensation - resistant to Fouling	New Design

Component or Subsystem	Technology Type	Basis for Design and Performance			Operating Condition				Assumed or specified Performance	Calculated Performance	Contaminant removed	Assumptions regarding Anticipated Application Issues	Technology Readiness
					Inlet		Outlet						
		Self-defined or Generic	Vendor Data /Commercial Design	Vendor Data / Future Design	Temperature (˚C)	Pressure (bar)	Temperature (˚C)	Pressure (bar)					
Power Island	Steam Expansion	X			599	242	~39	0.068	Turbine efficiencies as per FOA Attachment B	Gross Power output: 454 MWe	-	None	Mature/ Commercial
Direct Contact Condenser	Pressurized Direct Contact Heat Recovery + Pressurized wet FGD			X	320	28.1	40	27.5	98% SO$_2$ removal	L/G ratio (wt) of ~1.5:1	- 98% SO$_2$ removal	Resistant to fouling	New Design
DeNOx Unit	Lead-chamber Process		X (Air Products and Chemicals)		40	27.5	21	27.1	Based on lead-chamber kinetic model	NO$_X$ ppmv: 46 SO$_X$ ppmv: trace	NO$_X$ and SO$_X$	-	New Design
CPU (Inert Removal)	Cryogenic Distillation		X (Air Products and Chemicals)		21	27.1	4	150	Process parameters as per Attachment B	CO$_2$ purity: 99.95 mol% O$_2$ ppmv: 100 N$_2$: trace Ar ppmv: 52 NO$_X$ ppmv: 32 SO$_X$ ppmv: trace	Mainly O$_2$/N$_2$/Ar	-	Mature
Water Management System	Water Treatment		X		Various	Various	Various	Various					Mature/ Commercial

6. Block Flow Diagram and Stream Table

This sectioned includes the overall plant PFD and the associated Stream Table. The stream numbers in the PFD (Figure 16) and stream table (Table 8) correspond to each other.

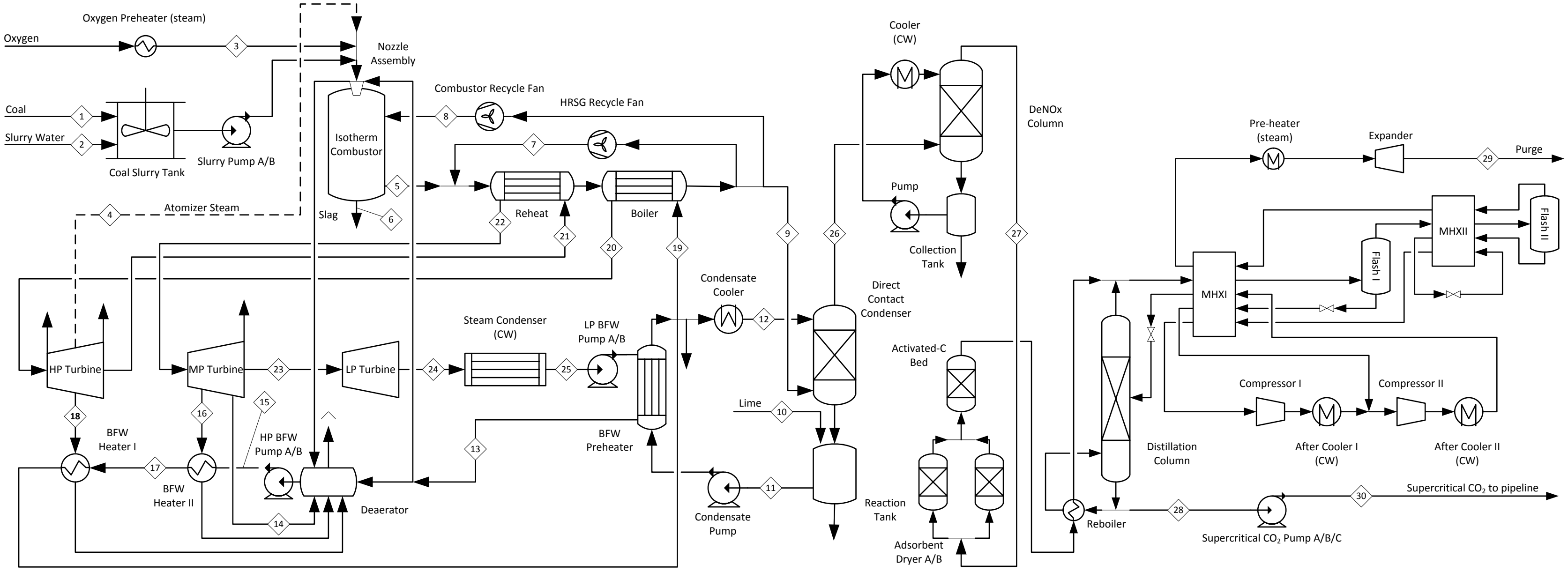

Figure 16 | Plant process flow diagram (stream numbers correspond to streams in Table 8)

Table 8 | Stream table (stream numbers correspond to streams in Figure 16)

Steam No.	1	2	3	4	5	6	7	8	9	10
Stream Name	**Coal**	**Slurry Water**	**Oxygen**	**Steam**	**Flue Gas**	**Slag (dry)**	**HRSG Recycle**	**Combustor Recycle**	**FG to DCC**	**Lime (dry)**
State	Solid	Liquid	Gas	Gas	Gas	Solid	Gas	Gas	Gas	Solid
Temperature (°C)	15	15	300	400	1450	1450	327	327	321	15
Pressure (bar)	1.0	1.0	29.0	58.0	29.0	1.0*	29.0	29.0	28.1	1.0*
Mass Flowrate (tonne/h)	133	73	297	7	1,745	13	2,691	1,248	497	6
Mole Flowrate (kmol/h)	-	4,067	9,230	370	57,343	-	88,412	41,015	16,328	107
Vapor Fraction	0.00	0.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00	0.00
Mole fraction (liquid and vapor streams)										
O_2			0.955		0.027		0.027	0.027	0.027	
N2			0.017		0.013		0.013	0.013	0.013	
Ar			0.028		0.016		0.016	0.016	0.016	
CO_2					0.433		0.433	0.433	0.433	
H_2O		1.000		1.000	0.504		0.504	0.504	0.504	
SO_x					0.006376		0.006376	0.006376	0.006376	
NO_x					0.000158		0.000158	0.000158	0.000158	
Gypsum										
Mass Fraction (solid only streams)										
Coal	1.00									
Ash/Slag						1.00				
Lime										1.00

Table 8 continued…

Steam No.	11	12	13	14	15	16	17	18	19	20
Stream Name	**Process Condensate**	**Process Cond. Rec.**	**Preheated BFW**	**Deaerator Bleed**	**HP BFW**	**FWH II Bleed**	**HP BFW**	**FWH I Bleed**	**HP BFW**	**HP Steam**
State	Liquid	Liquid	Liquid	Gas	Liquid	Gas	Liquid	Gas	Liquid	Gas (SC)
Temperature (˚C)	200	40	182	488	224	594	262	450	315	599
Pressure (bar)	28.1	30.5	23.0	22.2	289.0	43.7	288.7	96.1	288.3	242.0
Mass Flowrate (tonne /h)	928	766	928	44	1,196	79	1,196	151	1,196	1,196
Mole Flowrate (kmol/h)	46,443	38,353	51,528	2,467	66,408	4,362	66,408	8,385	66,408	66,408
Vapor Fraction	0.00	0.00	0.00	1.00	0.00	1.00	0.00	1.00	0.00	-
Mole fraction (liquid and vapor streams)										
O_2										
N_2										
Ar										
CO_2										
H_2O	0.985	0.985	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000
SO_x	0.000002	0.000002								
NO_x	0.000000	0.000000								
Gypsum	0.012	0.012								
Mass Fraction (solid only streams)										
Coal										
Ash/Slag										
Lime										

Table 8 continued…

Steam No.	21	22	23	24	25	26	27	28	29	30
Stream Name	**HP Turbine Exhaust**	**Superheated IP Steam**	**IP Turbine Exhaust**	**LP Turbine Exhaust**	**Steam Condensate**	**DCC Exhaust**	**Treated Flue Gas**	**Purified CO_2**	**Purge**	**Supercritical CO_2 Product**
State	Gas	Gas	Gas	Liquid/Gas	Liquid	Gas	Gas	Liquid	Gas	Liquid (SC)
Temperature (°C)	358	599	298	38	38	40	21	-8	7	4
Pressure (bar)	49.0	45.0	5.2	0.068	0.068	27.5	27.1	28.0	1.0	150.0
Mass Flowrate (tonne /h)	1,020	1,020	897	897	928	341	340	298	42	298
Mole Flowrate (kmol/h)	56,639	56,639	49,776	49,777	51,528	7,971	7,945	6,761	1,174	6,761
Vapor Fraction	1.00	1.00	1.00	0.91	0.00	1.00	1.00	0.00	1.00	-
Mole fraction (liquid and vapor streams)										
O_2						0.049	0.049	0.000	0.329	0.000100
N_2						0.027	0.028	0.000	0.186	0.000
Ar						0.032	0.032	0.000	0.219	0.000
CO_2						0.887	0.900	1.000	0.264	1.000
H_2O	1.000	1.000	1.000	1.000	1.000	0.004	0.001	0.000	0.000	0.000
SO_x						0.000256	0.000000	0.000000	0.000000	0.000000
NO_x						0.000323	0.000046	0.000032	0.000127	0.000032
Gypsum										
Mass Fraction (solid only streams)										
Coal										
Ash/Slag										
Lime										

7. Energy and Mass Balances

This section includes tables that quantitatively demonstrate that heat and mass streams are balanced for the overall system. For each balance, the boundary of the system or subsystem is defined.

7.1. Plant Energy Balance

The energy balance of the proposed plant is shown in Table 9. The energy balance includes main processing units but excludes energy contribution of some unit operations such as power consumption associated with cooling tower operation and coal grinding and handling.

Table 9 | Plant energy balance

Stream / Form of Energy	HHV	Sensible/Latent	Power	Total	Notes
		HEAT IN (MW)			
Coal	1,003	0		1,003	On HHV basis (30,506 kJ/kg dry coal)
Water inputs		~ 0		0	Includes slurry water, makeup water (DeNOx)
Air		~ 0		0	
Total (MW)	**1,003**	**2**	**0**	**1,003**	
		HEAT OUT (MW)			
Slag	0	6		6	
Cooling tower duty		670		670	Excluding cooling tower power demand
Electric motor losses		8		8	Consumers and generators combined
Ambient losses		2		2	Combustor, boiler, superheater (assumed value)
Turbine leaks combined		17		17	Includes governing stage, HP turbine and IP turbine leaks (currently being vented)
Product CO_2		-25		-25	In the form of compression (pressure) energy and sensible heat
ASU purge		0		0	
Deaerator vent		5		5	
DCC waste water		5		5	
$DeNO_x$ waste water		~0		~0	
Inert removal purge		~0		~0	
Net produced power			327	327	
Total (MW)	**0.0**	**688**	**329**	**1,017**	
Energy Imbalance (MW)				**-14**	

7.2. Plant Mass Balances

Table 10 represents the mass balance around the plant. The mass balance includes all the unit operations included in the process model which accounts for all process units. Other utility and supporting facilities, e.g., water loss in cooling tower, loss of coal during handling and storage operations, are not included in the mass balance.

Table 10 | Plant overall mass balance

Material Stream	Flow	Notes
MASS IN (tonne/h)		
Coal	133	
ASU Air	1,255	Excluding moisture
Slurry Water	73	
Makeup water (steam Island)	32	
Lime	6	Solid
Makeup water ($DeNO_X$ unit)	5	
Total (tonne/h)	**1,504**	
MASS OUT (tonne/h)		
Slag	13	Dry
ASU purge	959	Excluding moisture
Deaerator vent	6	
DCC waste water	162	Including solids
Turbine leaks	18	
$DeNO_X$ waste water	6	
Removed water in inert removal unit	0	Adsorbed in adsorbent bed
Inert removal purge	43	
Product CO_2	298	
Total (tonne/h)	**1,504**	
Mass Imbalance (tonne/h)	**0**	

7.2.1. Plant Carbon Balance

Table 11 depicts the carbon balance around the pressurized oxy-combustion plant designed in this study. The carbon balance excludes any loss of coal during coal handling and feed preparations unit operations within the plant.

Table 11 | Plant carbon balance

Stream / Form of Carbon	Fixed Carbon	CO_2	Total	Notes
CARBON IN (kmol/h)				
Coal	7,076		7,076	
Total (kmol/h)	**7,076**	**0**	**7,076**	
CARBON OUT (kmol/h)				
Slag	0		0	Based on 100% conversion assumption
DCC waste water		0	0	
DeNOx waste water		0	0	
Inert removal purge		311	311	
CO_2 product		6,758	6,758	
Total (kmol/h)	**0**	**7,069**	**7,069**	
Carbon Imbalance (kmol/h)			**+7**	

7.2.2. Plant Sulfur Balance

The sulfur balance of the plant is shown in Table 12.

Table 12 | Plant sulfur balance

Stream / Form of Sulfur	Pyritic /Sulfate /Organic	Gypsum	Sulfuric Acid	SO_X	Total
SULFUR IN (kmol/h)					
Coal	104				104
Total (kmol/h)	**104**	**0**	**0**	**0**	**104**
SULFUR OUT (kmol/h)					
Ash	0				0
DCC waste water		100	2	0	102
DeNOx waste water			2	0	2
CPU Purge				~0	0
CO_2 product				~0	0
Total (kmol/h)	**0**	**100**	**4**	**0.0**	**104**
Sulfur Imbalance (kmol/h)					**0**

7.2.3. Plant Water Balance

Table 13 shows the water balance around the plant boundary limit. The water balance table includes main process water streams only. The followings are excluded: water streams related to cooling tower, boiler blowdown(s), and water treatment units.

Table 13 | Plant water balance

Stream / Form of Water	Water	Moisture /Hydrates	Steam	Converted from/to Other Chemicals	Total	Notes
WATER IN (kmol/h)						
Coal		823			823	Moisture in AR coal
Slurry water	4,067				4,067	
Combustion product water	2,966				2,966	Produced during combustion
Steam island makeup water	1,752				1,752	
DeNO$_X$ makeup water	278				278	
Total (kmol/h)	**9,062**	**823**	**0**	**0**	**9,885**	
WATER OUT (kmol/h)						
Slag		0			0	
Deaerator vent			334		334	
Turbine Leakages			1,048		1,048	All leaks combined
DCC waste water	7,991	201 (in gypsum)		10	8,201	Basic solution Includes solids (gypsum and fly-ash)
DeNOx waste water	290			7	297	As acids
Inert Removal Adsorber Purge		10			10	Adsorbed on adsorbent bed (excluding regeneration steam)
CPU Purge		trace			0	
CO$_2$ product		trace			0	
Total (kmol/h)	**8,281**	**211**	**1,382**	**16**	**9,890**	
Water Imbalance (kmol/h)					**+5**	

8. Thermodynamic Performance

A key advantage of the proposed pressurized oxy-combustion process relative to atmospheric oxy-combustion is that it provides the possibility to recover more heat from the flue gas for power generation. Since the flue gas is at pressure (in this case ~29 bar) it is possible to recover the latent heat of moisture in the flue gas at high temperatures such that it can be used in the steam cycle for power generation and to increase the overall efficiency of the plant.

Figure 17 depicts the profile of recovered heat from flue gas in the pressurized oxy-combustion process. Unlike traditional steam cycles in which flue gas is at near atmospheric pressure, in this process the latent heat of the flue gas can be recovered and used in the steam cycle (DCC unit). It should be noted that the profile depicted in Figure 17 is for illustration purposes only and the actual TQ diagram of the plant is different with respect to the order of the heat removal stages, *i.e.*, boiler and reheat.

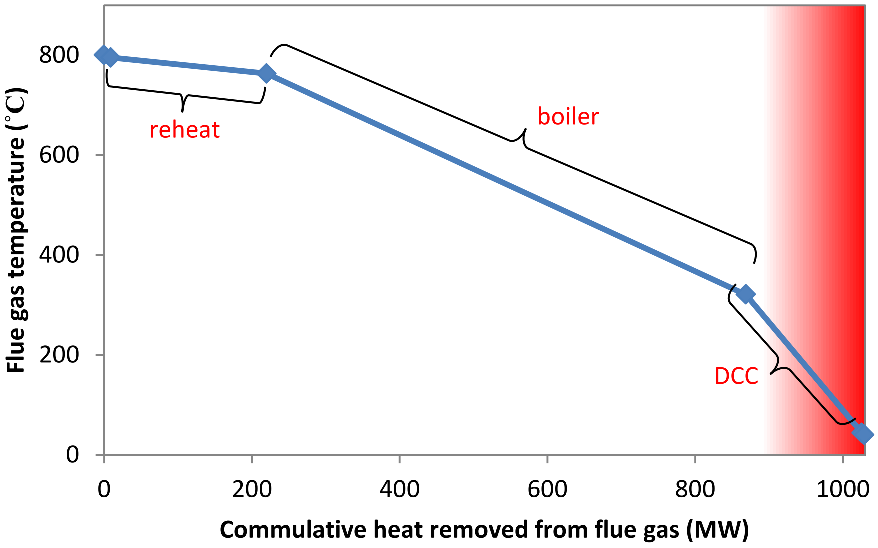

Figure 17 | Heat recovery in pressurized oxy-coal combustion process with direct contact condensation

In this study Illinois #6 bituminous coal is assumed as feedstock of the plant and all the reported figures represent plant performance for this coal. Based on the reason discussed above, it is expected that the proposed process has significant advantage over traditional coal power plants when it is used for low rank coals, e.g., Wilcox Group Lignite or PRB sub-bituminous [1], Attachment C. This is due to the fact that the latent heat related to the moisture in the low rank coals can be recovered in DCC and no longer leaves the plant as moisture in flue gas. Using the proposed pressurized oxy-combustion process for high moisture coal will affect the distribution of heat recovery in HRSG and DCC units and optimized operation of the plant will require change of the process operating conditions.

9. Technical Performance Summary

Thermodynamic performance of the proposed plant is summarized in this chapter. Table 14 shows the plant performance summary of the proposed pressurized oxy-combustion process.

Table 14 | Plant performance summary

Parameter	Unit	Value	Notes
Heat rate (HHV basis)	BTU/kWh	10,400	On HHV basis, 30506 kJ/kg dry coal (Illinois #6)
Gross Power	MW_e	454[†]	
Parasitic Load - ASU only - CPU only	MW_e	126 83 15	Total excluding cooling tower 66% of total parasitic load 12% of total parasitic load
Net Power	MW_e	328	
Efficiency (on HHV basis)	%	32.7	
Efficiency (on LHV basis)	%	33.9	
CO_2 Capture Rate	%	96	On basis of total carbon in feed coal

[†] The estimated gross power is based on the DOE guidelines for steam cycle configuration (Supercritical with one reheat) and steam turbine efficiencies. According to information received from vendors, current steam turbine technology should enable overall plant efficiency to be 1.5% higher than the value presented above.

10. Cost Estimation

The methodology used for estimation of capital costs and operations and maintenance costs are described below. The finance structure, basis for the discounted cash flow analysis, and first-year COE cost calculations are documented in Chapter 11. The capital costs associated with CO_2 transportation, sequestration, and monitoring are excluded from this study.

10.1. Capital Cost Estimation Methodology

The capital cost is built up by each plant area, including the process contingency applied to each. The estimation basis for each cost component is provided, e.g., a factored analysis based on a similar system, or vendor estimates for commercial equipment or vendor projections for conceptual equipment. Capital costs are reported at the following levels, consistent with the NETL report, “QGESS:Cost Estimation Methodology for NETL Assessments of Power Plant Performance” [3]: bare erected cost (BEC), total plant cost (TPC), total overnight cost (TOC), and total as-spent cost (TASC). Owner’s costs are broken down into the following categories: prepaid royalties, pre-production costs, inventory capital, initial cost for catalyst and chemicals, land, financing costs, and “other” but not all owner’s cost components are included in the cash flow and first-year COE cost calculations because no entries for these components exist in DOE’s PSFM ver. 6.6.4 for pre-defined cases.

The cost estimates carry an accuracy of -30%/+50%, consistent with a “feasibility study” level of design engineering applied to the various cases in this study [3]. The value of the study lies not in the absolute accuracy of the individual case results (proposed process case and the baseline case) but all cases are evaluated under the same set of assumptions. This consistency of approach allows meaningful comparisons among the cases.

Each bare erected cost (BEC) account was evaluated against the level of estimate detail and field experience to determine process contingency. Process contingency is intended to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology. Higher process contingencies were added for process areas that were deemed to be first-of-a-kind or posed significant risk due to lack of operating experience.

Project contingencies were added to the BEC and EPC fees and process contingencies to cover project uncertainty and the cost of any additional equipment that would result from a detailed design. The contingencies represent costs that are expected to occur. A project contingency of 15% is applied to the sum of BEC, EPC fees and process contingencies. Table 15 show the guidelines used for specifying process contingency for different plant areas [3].

Currency exchange factor of 1.3 is used to convert Euro to USD throughout this analysis.

Table 15 | Guidelines for applying process contingencies [3]

Technology Status	Process Contingency (% BEC)
New concept with limited data	40+
Concept with bench-scale data	30-70
Small pilot plant data	20-35
Full-sized modules have been operated	5-20
Process is used commercially	0-10

10.2. Capital Cost Estimation Results

Table 16 represents the breakdown of bare erected cost of the proposed plant along with the process contingencies considered for each plant area based on maturity and commercial status of each technology.

Table 16 | Breakdown of bare erected cost for different plant areas and assumed process contingencies

PLANT AREAS	BEC MM$ (2012Q1)	Process Contingency	Process Contingency MM$ (2012Q1)	Cost Basis
Coal Receiving and Storage	18	10%	2	[20]
Coal Grinding and Slurry Preparation	21	10%	2	[23]
ASU	208	5%	10	[24]
Combustion[†]	112	20%	22	[23]
HRSG	197	25%	49	[23]
Power Island (including steam condenser)	141	10%	14	APEA
Flue Gas Treatment (includes DCC and DeNOx units)	37	20%	7	APEA
CPU (cryogenic inert removal)	21	15%	3	APEA
Water Treatment[‡]	87	10%	9	[24]
Balance of Plant	3	10%	0	APEA
TOTAL	**845**	**14%**	**120**	

[†] Part of the combustion area bare erected cost includes field related costs such as inspection, commissioning and supervision; Adjustment is made in calculating other capital cost levels, e.g., EPCC and TPC.

[‡] The cost figure of the water treatment area is for a “zero discharge” design including crystallization and total water recovery which is a strict approach in water treatment design. In addition, the report includes the water treatment requirement of the DCC unit (~MM$18.5). The water treatment area cost may vary depending on plant site.

The Engineering, Procurement, Construction and Management (EPCM) contractor services are estimated to be 10 percent of BEC. These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes construction management, resident engineer(s), scheduler(s), and personnel for project controls, document control, materials management, site safety, and field inspection [3].

The TPC for each technology was determined through a combination of vendor quotes, scaled estimates from previous designed/built projects, and Aspentech APEA software. TPC includes all major equipment, materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project). Escalation and interest on debt during the capital expenditure period were estimated and added to the TOC to provide the Total As-Spent Cost (TASC). The Total Overnight Cost (TOC) of the plant is calculated by adding owner's costs to the Total Plant Cost (TPC).

Different levels of the plant's capital cost are shown in Table 17. The BEC, EPCC, TPC and TOC figures are on un-escalated cost basis but TASC figure is report on escalated cost basis.

Table 17 | Capital cost of the proposed plant calculated by PSFM

Capital Cost Level	Value MM$ (2012Q1)	Notes
BEC (bare erected cost)	845	- process equipment, supporting facilities, direct and indirect labor - Un-escalated cost
EPCC (eng., proc., and cons. cost)	929	- BEC + EPC fees (10% of BEC) - Un-escalated cost
TPC (total plant cost)	1,200	- EPCC + process contingencies (average of 14% of EPCC) and project contingency (15% of EPCC and process contingencies) - Un-escalated cost
TOC (total overnight cost)[†]	1,215	- TPC + financing costs - Un-escalated cost
TASC (total as-spent cost)	1,383	- Interest and escalation during construction - Escalated cost

[†] PSFM version 6.6.4 does not take into account any other component of owner's cost.

10.3. Operations and Maintenance Costs

A factor base method is used to calculate the operations and maintenance costs of the entire plant for this study. The factors are shown in Table 18. The presented variable O&M factor is for 100% plant capacity factor.

Table 18 | O&M costs factors

Parameter	Value	Notes
Variable O&M Costs Factor	1.5% of EPCC	Default value in PSFM version 6.4.4 (for 100% capacity factor)
Fixed O&M Costs Factor	3.5% of EPCC	Default value in PSFM version 6.4.4

The operating costs associated with CO_2 transportation, sequestration, and monitoring are excluded from this study.

11. Financial Analysis

The cost metric used in this study is the Cost of Electricity (COE), which is the required revenue per net megawatt-hour during the power plant's first year of operation, assuming that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate. To calculate the COE, the Power Systems Financial Model (PSFM) version 6.6.4 was used to determine a "base-year" (2015) COE that, when escalated at an assumed nominal annual general inflation rate of 3 percent, provided the stipulated internal rate of return on equity over the entire economic analysis period (capital expenditure period plus thirty years of operation).

The economic performance evaluation performed is based on the global economic assumptions listed in Table 19 which is based from the NETL report, "Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance" [3].

Table 19 | Global economic assumptions

Parameter	Value
TAXES	
Income Tax Rate	38% Effective (34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 years
CONTRACTING AND FINANCING TERMS	
Contracting Strategy	Engineering Procurement Construction Management (EPCM) owner assumes project risks for performance, schedule and cost
Type of Debt Financing	Non-Recourse (collateral that secures debt is limited to the real assets of the project)
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	0 years
Debt Reserve Fund	None
ANALYSIS TIME PERIODS	
Capital Expenditure Period	5 Years
Operational Period	30 years
Economic Analysis Period (used for IRROE)	35 Years (capital expenditure period plus operational period)
TREATMENT OF CAPITAL COSTS	

Parameter	Value
Capital Cost Escalation During Capital Expenditure Period (nominal annual rate)	3.6%
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	5-Year Period: 10%, 30%, 25%, 20%, 15%
Working Capital	zero for all parameters
% of Total Overnight Capital that is Depreciated	100% *(this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)*
ESCALATION OF OPERATING REVENUES AND COSTS	
Escalation of COE (revenue), O&M Costs, Fuel Costs (nominal annual rate)	3.0%

11.1. Finance Structures

The economic performance evaluation performed for this study is in accordance with the high-risk, Investor Owned Utility (IOU) finance structure with a 5 year capital expenditure period as outlined in the NETL report, "Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance" [3]. Table 20 shows the financial structure assumed for this study.

Table 20 | Assumed financial structure

Type of Security	% of Total Investment	Current (Nominal) Dollar Cost	Weighted Current (Nominal) Cost
HIGH RISK INVESTOR OWNED UTILITY			
Debt	45	5.5%	2.475%
Equity	55	12%	6.6%
Total			9.075%

All other cases which this case is compared against are also assumed to be high-risk and have the same distribution of investment (debt vs. equity). Although according to [3] PC plants with CO_2 capture would be considered high risk but non-capture PC case should be considered low risk. Table 21 describes the financial structure assumed for this study.

11.2. COE Results

Supercritical Pulverized Coal (SCPC) is considered as analogues technology for comparison purposes in the results section.

Table 21 summarizes the COE results of the proposed process in comparison with typical SCPC with and without capture calculated predefined in PSFM tool version 6.4.4. As mentioned above, a similar financial structure is assumed for all three cases.

Table 21 | Calculated LCOE using PSFM tool

Case	First Year COE in Base Year Dollars ($/MWh)	Levelized COE in Base Year Dollars ($/MWh)
SCPC†	59	75
Pressurized Oxy-combustion	**86** (DOE target: 80)	108
SCPC w/ CCS†	105	133

† Predefined cases in PSFM tool, version 6.4.4 with the case scenario input parameters.

As illustrated in this report, the proposed pressurized oxy-combustion process captures >95% of carbon and delivers it in the form of supercritical pure CO_2 steam while it poses reasonable heat rate (10,400 BTU/kWh) and overall net plant efficiency of ~33% (on HHV basis). In addition, the proposed plant is capable of producing power at a COE of 86$/MWh in the first year of its operation which is slightly above the DOE target of 80$/MWh.

11.3. COE Breakdown

Figure 18 depicts the breakdown of the first-year COE in base-year dollars. In calculating the COE breakdowns for the proposed plant a capital charge factor of 0.133 (calculated using PSFM) and capacity factor of 85% are assumed [3].

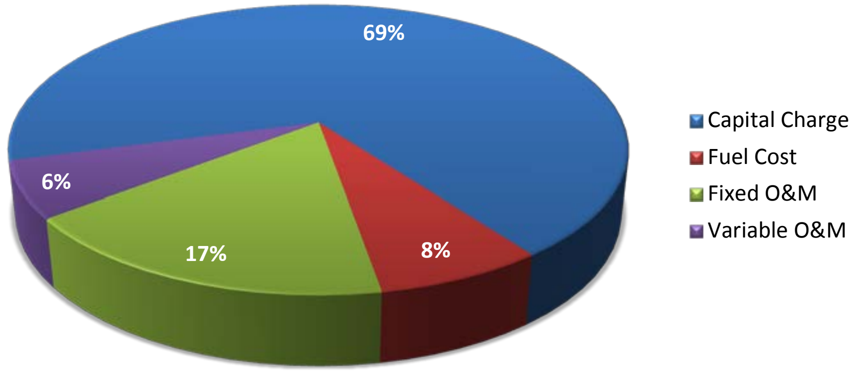

Figure 18 | Breakdown of first-year COE

11.4. Cost of Avoided CO₂

The cost of avoided carbon dioxide (CO_2) emissions is calculated as:

$$CO_2\ \ Avoided\ \ \ Cost = \left(\frac{COE_{Capture} - COE_{NoCapture}}{CO_2Emissions_{NoCapture} - CO_2Emissions_{Capture}} \right) = \frac{86 - 59}{0.802 - 0.042}$$

The calculated CO_2 avoided cost for the proposed process is 36 $/tonne of CO_2 emissions. The No Capture value of $59/MWh in the above equation is calculated using PSFM for pre-defined SCPC case. The No Capture value of 0.802 tonne CO_2/MWh in the above equation corresponds to the super-critical PC with capture case (Case 11) in [20].

According to DOE guidelines, cost of captured carbon dioxide (CO_2) emissions is to be calculated as:

$$CO_2\ Capture\ Cost = \left(\frac{COE_{Capture} - COE_{NoCapture}}{CO_2 Captured_{PerNetOutput}} \right) = \frac{86-59}{298/328}$$

The cost of CO_2 capture cost for the proposed is plant 30 $/captured tonne of CO_2.

11.5. Sensitivity Analysis

This section represents the results of the sensitivity analysis of key parameters impacting economic metrics (*i.e.*, first year COE in base-year dollars and Cost of Avoided CO_2).

11.5.1. Sensitivity to Fuel Cost

Sensitivity of first year COE and CO_2 avoided cost of the proposed plant to cost of fuel (coal) is presented in Figure 19. The coal price for the financial assessment of the proposed plant is based on fuel price forecast values included in PSFM tool, version 6.6.4. While performing the sensitivity of CO_2 avoided cost, the fuel cost changes only for the proposed plant, not for the SCPC case (SCPC case in PSFM and Case 11 in [20]).

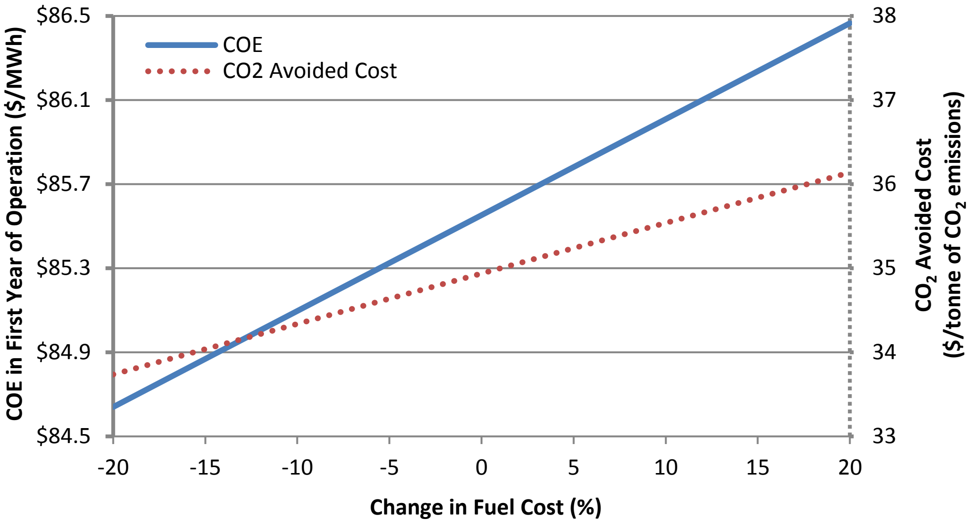

Figure 19 | Sensitivity of COE and cost of avoided CO_2 to fuel cost

11.5.2. Sensitivity to Capital Cost

Figure 20 depicts the sensitivity of first-year COE and CO_2 avoided cost of the proposed oxy-combustion plant to capital (EPCC) cost based on the results from PSFM tool. In COE and CO_2 avoided cost calculations the capital cost of the predefined baseline case (SCPC in PSFM tool) is kept constant and only EPCC of the proposed process is varied within the range of -20% to +20% of its estimated value. EPCC is considered as the plant's capital cost because it is the level of capital cost input to PSFM tool.

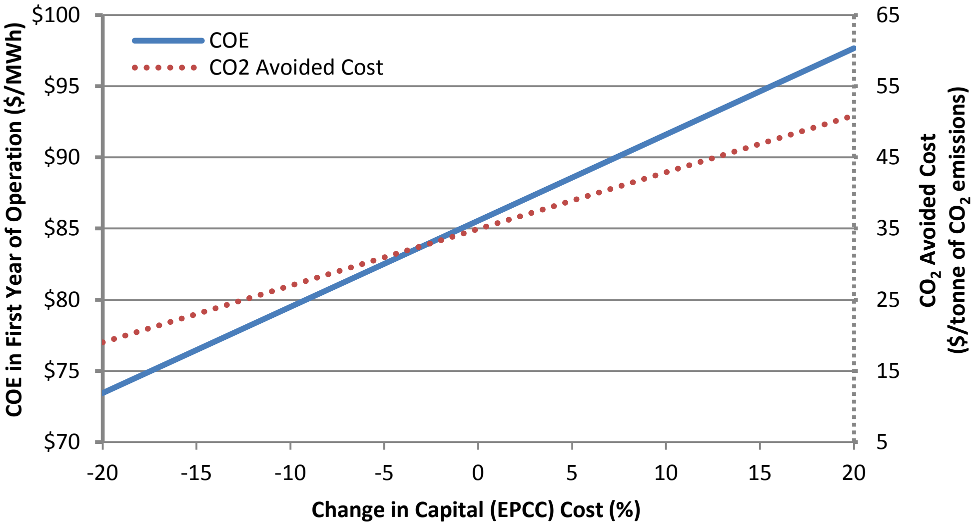

Figure 20 | Sensitivity of COE and cost of avoided CO_2 to capital cost

11.5.3. Sensitivity to Capacity Factor

Figure 21 shows the sensitivity of first-year COE and CO_2 avoided cost of the proposed oxy-combustion plant to plant capacity factor based on results from PSFM tool. The default capacity factor is assumed to be 85% for both the proposed case as well as the no capture SCPC case (reference [20], Case 11). While performing sensitivity, the capacity factor for base (SCPC w/o capture) case is kept constant 85% and only the capacity factor of the pressurized oxy-combustion case is changed within the range of -20% to +20% of the default value.

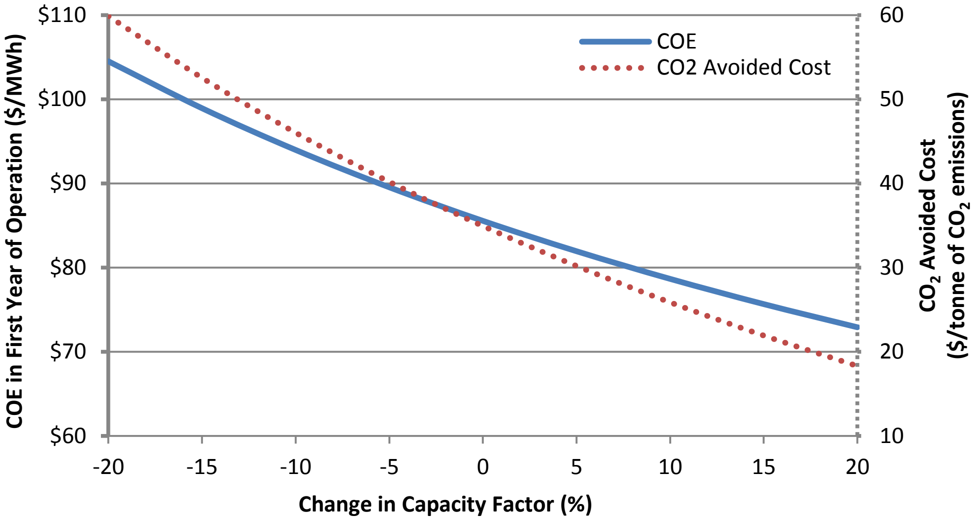

Figure 21 | Sensitivity of COE and cost of avoided CO_2 to plant capacity factor

As can be seen in Figure 19, Figure 20, and Figure 21, the sensitivity ranking order of COE and CO_2 avoided cost of the proposed process is Capacity Factor, Capital Cost and Fuel Cost.

12. Environmental Performance

The standards specifying environmental targets of power plants are usually stated on gross generated power in the plant [20]. Unlike conventional coal power plants, the proposed oxy-combustion process does not emit its fumes into atmosphere with the exception of a small steam of CO_2 and inert components which leaves the inert removal unit as tail gas (stream 29 in Figure 16 and Table 8). As a result, generally the emissions of the proposed plant are significantly lower than limits specified by various standards and regulations.

13. Other Deliverables

As part of the project documentation, economic spreadsheets/models/tools (including the completed NETL Power Systems Financial Model) developed as part of this project are delivered to DOE. The supporting material of the project include the developed process model in Aspen Plus as well as DOE's Power Systems Financial Model used for financial assessment of the proposed plant with a defined Case (Case G) representing the proposed pressurized oxy-combustion process.

13.1. Process Model

The process model is developed using Aspentech Aspen Plus® process simulation software version v8.0.

Process model package includes the following files which all need to be in the same folder to run the process model:

- Aspen Plus process model file (Pressurized Isotherm Oxy-combustion Process Model.bkp);
- Link file (kinetics.opt)
- Dynamic Link Library file (nitric.dll)

13.1.1. Modeling Approach

Each process unit (block) shown in Chapter 4 is included in a hierarchy block in the process model developed in Aspen Plus®. All process units with the exception of ASU are modeled to represent all the main pieces of process equipment. The ASU sub-model is not as rigorous as other parts of the process model since it's a mature technology and is only used to represent the ASU power demand with a good accuracy when compared to vendor-reported performance figures.

In addition, coal receiving and handling unit is not included in the process model since it does not have major impact on plant's overall mass and energy balance.

13.1.2. Property Methods

Table 22 shows different property methods used for various process areas of the proposed plant. Moreover, in order to improve the prediction of some of the used property methods, some of the property parameters are altered based on available experimental data in literature.

Table 22 | Property methods used in the Aspen Plus process model

Section	Property Method
Coal receiving and Storage	Not modeled
Coal Grinding and Slurry Preparation	Partially modeled
Air Separation Unit (ASU)	Peng-Robinson (PENG-ROB)
Isotherm Oxy-combustion	PENG-ROB
HRSG	- Steam tables (STEAM-NBS)
Power Island	- Steam tables (STEAM-NBS)
Direct Contact Condenser (pressurized wet FGD)	- ELECNRTL
DeNOx Unit (lead-chamber process)	ELECNRTL (selected to enable modeling of electrolytes chemistry in the process)
CPU (cryogenic inert removal)	RK-ASPEN

13.1.3. Process Parameters

Process Parameters used for process modeling are taken from Attachment B of the FOA [1] where applicable.

13.1.4. Model Chemistry

Two chemistry packages defined in the process model are Wet FGD and Lead-chamber. These chemistry packages are used in DCC and $DeNO_x$ units, respectively. Both chemistry packages (sulfuric and nitric), which were originally developed by Aspentech, were modified to have a better representation of the chemistry taking place at the operating conditions. Table 23 lists reactions included in Wet FGD chemistry package used in the DCC unit. Table 24 lists reactions included in lead-chamber chemistry package used in the DeNOx unit.

Table 23 | Wet FGD chemistry package used in DCC unit

No.	Reaction	Rxn Type	Notes
1	$2H_2O$ <--> $H3O^+$ + OH^-	Equilibrium	Water Dissociation
2	$2SO_2$ + O_2 --> $2SO_3$	-	For process simulation purposes 98% SO_2 conversion is specified
3	SO_3+ $2H_2O$ <--> HSO_4^- + H_3O^+	Equilibrium	
4	HSO_4^- + H_2O <--> SO_4^{2-} + H_3O^+	Equilibrium	
5	H_3O+ + SO_4^{2-} <--> H_2SO_4 + H_2O	Equilibrium	
6	CaO + H_2O --> $Ca(OH)^+$ + OH^-	Dissociation	Salt Dissociation (1st step)
7	$Ca(OH)^+$ <--> CA^{2+} + OH^-	Equilibrium	Salt Dissociation (2nd step)
8	Ca^{++} + SO_4^{2-} + $2H_2O$ <--> $CaSO_4.2H_2O$	Equilibrium	Salt Formation

Table 24 | Lead-chamber (SOx/NOx removal) chemistry package used in DeNOx unit

No.	Reaction	Rxn Type	Notes
1	2NO + O_2 --> $2NO_2$	Kinetic	Limiting reaction
2	$2NO_2$ --> N_2O_4	Equilibrium	
3	$3N_2O_4$ + H_2O --> 4 HNO_3 + 2NO	Kinetic	Represented as kinetic reaction; in reality it is a mass transfer-limited reaction
4	HNO_3 + H_2O <--> H_3O^+ + NO_3^-	Equilibrium	
5	SO_2 + NO_2 <--> SO_3 + NO	Equilibrium	
6	SO_3+ $2H_2O$ <--> HSO_4^- + H_3O^+	Equilibrium	
7	HSO_4^- + H_2O <--> SO_4^{2-} + H_3O^+	Equilibrium	
8	H_3O+ + SO_4^{2-} <--> H_2SO_4 + H_2O	Equilibrium	

13.2. Financial Model

Version 6.6.4 of the Power Systems Financial Model (PSFM) published by DOE is used to perform discounted cash flow and financial assessment of the proposed plant. The proposed pressurized oxy-coal combustion plant is added as POCC (Case G) to PSFM.

14. Conclusions and Recommendations

An analysis of key findings and recommendations for follow-on work is included in this chapter.

14.1. Key Findings

The following conclusions can be drawn from this system-level study:

- Between the recycling schemes investigated in this project, wet recycling demonstrates the highest performance.
- The optimized efficiency is relatively flat for the pressure range 20-60 bar. Therefore within this pressure range, other factors such the equipment size and costs will dominate the decision when designing a commercial facility. Within the relatively flat efficiency-pressure region, the plant efficiency was maximum at 29 bar.
- CO_2 capture rate of greater than 95% is possible in the proposed pressurized oxy-combustion capture.
- The selected CO_2 purification technology for the proposed process is capable of producing CO_2 product with the most stringent purity limits to meet EOR. Using less stringent CO_2 purity levels, for example for storage of CO_2 in saline reservoirs, can decrease the overall plant parasitic load, hence, increase overall efficiency.
- Due to higher densities in the pressurized oxy-coal combustion process, unit operations are smaller and the plant has a small footprint relative to conventional power generation plants with the same generation capacity.
- Due to smaller equipment sizes as well as plant footprint, it is estimated that the capital cost of the plant is lower than the conventional technologies. To reveal the exact cost reduction magnitude with better accuracy, detailed engineering of the proposed plant (at equipment level) is required.
- Although the increase in COE of the proposed pressurized oxy-combustion process is greater than 35% (DOE target), when compared to conventional power generation technology without CO_2 capture (SCPC w/o CC), it is significantly lower than that of state-of-the-art pre- and post-combustion capture technologies.
- The proposed pressurized oxy-combustion process is expected to offer similar advantages over ambient pressure power generation technologies when used for lower rank coals (i.e., lignite and sub-bituminous). This is due to the fact that the proposed process is capable of recovering the latent heat of the moisture in the combustion fumes and utilizing it in the low-temperature section of the steam cycle.

14.2. Recommendations for Follow up Work

Based on the conclusions drawn above, the proposed pressurized oxy-combustion process offers significant advantages in terms of high CO_2 capture rate (>95% of carbon contained in fuel) within reasonable increase in cost of electricity compared to the state-of-the-art power generation technologies. Quantifying advantages of the proposed process, in terms of technical performance and economic viability, requires further investigation. In particular investigating the followings can result in even greater performance improvements and cost savings as well as reducing the technical risks associated with deployment of the proposed technology:

- Experimental investigation of high pressure oxy-combustion of coal in the range of pressure suggested in this report. The assumptions used in modeling of the proposed process in this study are based on experiences gained from experimental work of the pilot plant operation at ~5 bar. The following items are of interest in high-pressure oxy-coal combustion:
    - Carbon conversion
    - Concentration of SOx and NOx components in combustion fumes
    - Concentration of fly-ash (particulate matter) in combustion fumes
- Detailed engineering design of the proposed process
- Examination of various components (equipment) of the plant by a multi-disciplinary team to validate reasonability of cost, operation and availability

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Appendix A – Previous Work

Simultaneous multi-variable optimization, as introduced in [11, 12], is required to obtain the optimum operation and achieve an attractive cycle performance. Optimization performed in [11], where thermal recovery is achieved by means of a surface heat exchanger, contributes in significant efficiency increase, 0.75% points over the literature proposal of 10 bar combustor pressure [8], while simultaneously reducing the combustor's operating pressure, to the range of 7bar, thus making the process more attractive and practical. Efficiency is 3.1% points higher than that of the atmospheric operation. Results also show the importance of the 15 other optimization variables in obtaining such efficiency improvements. However, this type of heat exchanger is subjected to considerable corrosion from condensation of sulfuric acid from the SOx in the flue gas. While the surface heat exchanger is thermodynamically more efficient, it the costs of acid tolerant materials of construction are prohibitively expensive.

A more practical and economic alternative for thermal heat recovery which is less susceptible to corrosion is a direct contact condensation with neutralization as discussed in [12]. There it is demonstrated that the optimum performance is still highly attractive and comparable to the surface heat exchanger recovery process despite the use of the Direct Contact Condenser (DCC) while operating at a slightly higher pressure of around 15 bar. Direct contact columns are used in various engineering and chemical processes [25] and in fact are used in commercial CO_2 purification process in the OCC cycle, where the dry flue gas is purified from nitrogen and sulfur oxides and other contaminants. In general, the direct contact process is performed by having two streams in a vertical counter flow arrangement where undesired substances in the gas phase is washed down and removed by a liquid stream.

In this study, similar to [12], a DCC unit is used to condense water vapor from the flue gas and recover some of the latent and sensible energy, but with a few modification to major components of the process to better match the current industrial equipment and operating units in particular the DCC, the heat recovery steam generator (HRSG), and the flue gas recycling loop; moreover, an updated and more accurate designs of the air separation unit (ASU) and the CO_2 Purification Unit (CPU) are implemented. It is important to mention the main changes that contribute in the difference of the results of this current study compared to those of [12]:

a) The coal utilized in this study is a common USA benchmark coal, Illinois #6, as opposed to North African coal used in [12].
b) The DCC configuration, operating strategy and modeling assumptions in this study differ the how the system was modeled in [12].
c) The pressure losses inside the HRSG and the flue gas recycling pipes are considered to be a fixed fraction of the flue gas pressure, as opposed to calculating the pressure losses based on similarity analysis for a fixed design.
d) The ASU is more accurately represented.

e) The CPU is modeled with higher accuracy and is designed to deliver a higher CO_2 purity product suitable for enhanced oil recovery applications.
f) The turbine expansion line specifications follow the DOE guidelines in Attachment D of [1].
g) The combustor temperature in this study is 1450˚C versus 1550˚C as specified in [12].

These are the major changes in the design, and change the cycle's performance substantially, which demands a comprehensive optimization study.

Appendix 2

UPA – GTRC final report to UPA

SubAwardee Final Report to Prime Awardee.

Project: DE-FOA-0000636
Title: OPTIMIZATION OF PRESSURIZED OXY-COMBUSTION WITH FLAMELESS REACTOR
Prime Awardee: Unity Power Alliance (an LLC formed by ThermoEnergy and ITEA)
SubAwardee: Georgia Tech Research Corp (GTRC)
SubAward PI: Dr. Scott Sinquefield

Background

This project involves building a pressurized coal combustion pilot plant and testing it. The plant is designed to demonstrate coal combustion in a flameless regime of operation. The Italian firm ITEA, who is a collaborator on this project, owns the technology, and has two full scale operating plants. These plants operate at about 4-5 bar pressure and produce high pressure steam for electricity generation. A key goal of this project was to test the flameless operation at much higher pressures, up to 40 bar. The flameless combustion zone is created by recycling CO_2 and water vapor from the stack gas back into the combustor. This slows the kinetics enough to prevent a flame front from forming in the combustion chamber, and reduce it to a hot gas-phase reaction. The purpose of the pilot was to demonstrate the combustor could operate at steady state over a range of pressures and with coals of varying rank (i.e. heating value), and produce high carbon conversion efficiency with minimal excess oxygen required. According to ITEA, a crucial issue is keeping the SO_3 gas concentration low in the gas leaving the quenching section before the heat transfer section.

GTRC was contracted on this project to provide a host site to operate the reactor safely and run a parametric series of tests, and to sample a hot gas stream taken from the quenching section and provide chemical analysis. ITEA and UPA agreed upon an experimental matrix consisting of three coals of varying rank, and two operating pressures. GTRC was to construct a sampling system and collect raw gas samples from a side stream leaving the reactor, and condition the sample so as to separate and collect different chemical species of interest at different points in the sampling train. Details and results are presented in the following sections

Experimental

A matrix of two pressures and three coals were specified by ITEA and UPA:

1. Pressures: 25 bar, and 40 bar
2. Coals: a) Texas lignite, b) Powder River Basin (subbituminous), and c) Illinois #6 (bituminous).

The sample port was built into the reactor and placed near the exit of the combustion chamber just after a point where steam is injected to cool the gases to a point where they can enter a steam superheater. Steam is injected inject in this pilot reactor to simulate exhaust gas recirculation in the envisioned full scale unit. The data to be collected by GTRC included:

1. Fly ash particulates; sampled at 200C
2. SO_3 vapor; to be condensed out at 130C
3. Water vapor and all soluble species; to be condensed at 0C
4. Permanent gases.

Figure 1. The UPA/ITEA pilot plant.

The pilot is shown in Figure 1. At the extreme right of the picture, half of the coal slurry feed tank can be seen. Coal is pumped in with a staged progressive cavity pump in slurry form. Oxygen is fed via a separate injection system. The reactor is pre-heated with methane at low pressure. When hot enough, it can be switched over to coal and the pressure increased to the set point. The heat contained in the reactor vaporizes the water in the slurry and the coal/O_2/steam mixture reacts. The reaction chamber is located inside the largest pressure vessel visible on the right hand side of Figure 1. The slag falls through a refractory funnel in the reaction chamber and then into a crucible at the bottom of the vessel. The slag accumulates and must be removed manually by unbolting and lowering the bottom dome. The hot (1600C) product gases exit the reactor to the side through the horizontal section visible in Figure 1. The quenching steam is injected at the entrance to the horizontal section. Quenching drops the gas temperature to 700C. It is near the center of this horizontal section where the gas sample was located. It was a true side port in that is did not extend into the middle of gas flow. No attempt was made to do isokinetic sampling. UPA/ITEA was not concerned about this as they were interested only in ultrafine aerosols and not is sampling a representative particle size distribution. At the sample port on the pressure vessel, there were two large needle valves placed in series. By adjusting these valves, it was possible to step down the reactor pressure to a low value for the sampling train (i.e. about 10-20 psig). It was after these valves that GTRC's gas conditioning and sampling began.

The entire sampling train is shown in Figure 2. The raw gas sample was carried through a ¼" stainless heated sample line to an oven held at 200C. A tee fitting was used to vent a portion of the sample through a precision metering valve. This allowed for fine tuning of the inlet gas pressure which was monitored by a gauge. A stainless steel filter housing with glass fiber filters was placed in the oven to collect fine particulates. The heat-traced line then carried the gas to a second oven held at 120C, containing a Teflon filter housing with coalescing element. A 10-foot length of Teflon tubing led into the filter to act as a heat transfer means to ensure the gas temperature dropped to the set oven temperature before entering the filter. This filter captured condensed H_2SO_4 vapor but allowed water to remain in the vapor phase and proceed to the chilled impingers. Following the H_2SO_4 filter was a small chest-type freezer which served to keep an ice-bath cold. Two 2-liter sized Erlenmeyer flasks in series were immersed in the ice-water bath to condense all water vapor. Following these condensers was a packed bed of silica gel to ensure the gas was completely dry. Following the silica was a volumetric gas flow totalizer similar to the meters used to measure residential natural gas usage. This provided a measure of the total volume of gas sampled over the course of a test run. An electronic gas mass flow meter (MFM) was also placed in line to dial in the gas sampling rate to a desired level but the MFM reading was not needed as data.

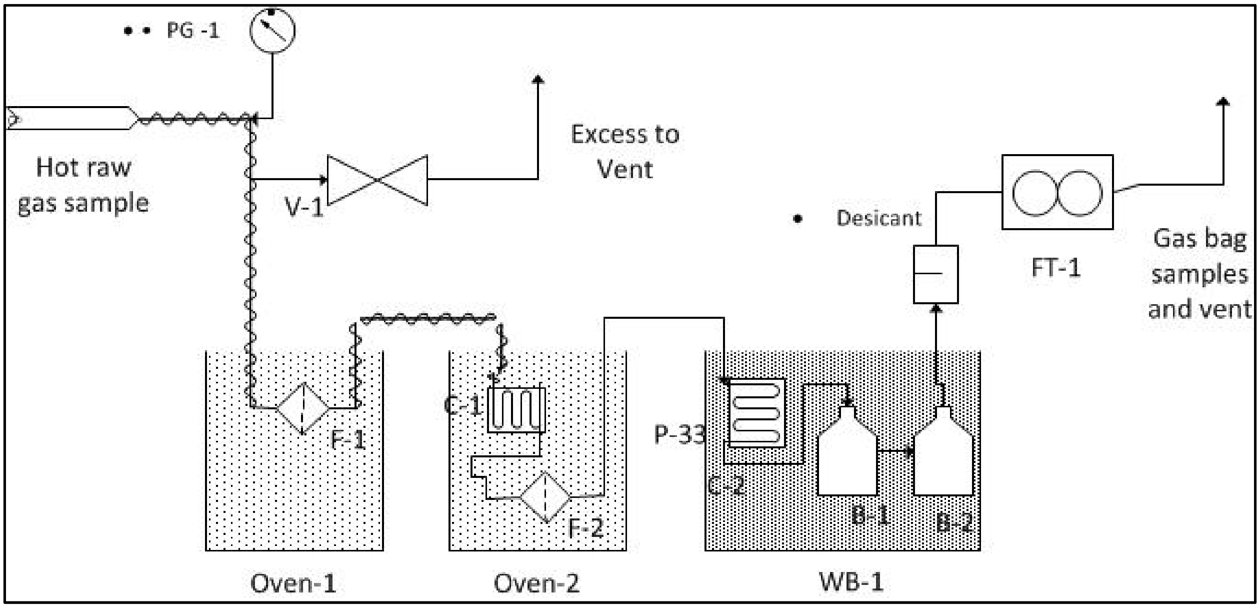

Figure 2. Diagram of the gas sampling system.

Table 1

Sampling system component details of Figure 2

PG-1	Pressure gauge used with V-1 to adjust line pressure to achieve 1-2 SLM dry gas through FT-1
V-1	Swagelok SS-4BW metering valve
Oven-1	Lab oven rated to 240C
F-1	Headline 136 SS Filter housing, with 25-64-50 glass microfiber cartridge filters
Oven-2	Precision™ lab oven rated to 230C
C-1	10-ft coil of ¼" TFE tubing used as heat exchanger
F-2	Headline 132P Teflon filter housing, with 25-64-50C microfiber coalescing filters
WB-1	Ice water bath placed in chest-type freezer.
C-2	10-foot coil of ½" stainless tubing, used as chiller/condenser
B-1, B-2	2-liter Erlenmeyer flasks used to collect condensed water vapor
Desiccant	A plexiglass flow tube filled with indicating desiccant
FT-1	CleanAir Technologies S275 DGM, mechanical gas volumetric totalizer

Sampling Details

ITEA had indicated that determining the SO_3 gas concentration was the most important measurement. They informed us that the raw gas sample would be 95% water vapor (due to the quenching steam addition) and 5% other gases. They also expected SO3 concentrations possibly as low as 10ppm, but that was based on data from their low pressure coal combustors. Thus the best method would be one where SO_3 condenses to H_2SO_4 and accumulates so as to get a larger quantity to analyze. A sampling system was built in accordance with EPA method 8A for SO_3. The method uses a quartz glass filter held at a temperature slightly above the boiling point of water. H2SO4 vapor pressure is lower than water and will condense on the quartz. ITEA's analytical chemist specified 120C as being adequate. The H_2SO_4 sample was collected by rinsing the Teflon tube and filter housing with DI water, and placing the coalescing element in the jar with the rinsate. [It was planned to also use an FT-IR with heated gas cell as a parallel analytical tool. However the FT-IR developed problems just before testing was to start. There was not enough time to have it repaired.] Not knowing the SO_3 concentrations to expect, we made conservative calculations and determined that we should sample 4000 wet standard liters of raw gas. At 95% water vapor, this would yield 200 standard liters of permanent gas. We estimated this would require about 2 hours of sampling depending on the actual steam concentration.

Gas sampling was initiated when the ITEA personnel operating the combustor inform us the reactor was at steady state and we can start. All heat tracing and ovens were brought to temperature, and impingers placed in the ice water bath. The raw gas flow was initiated and allowed to vent for a few minutes. Then the valve V-1 was adjusted to provide 1-2 SLM in the gas totalizer. This was found to be a pressure of about 5 psig at V-1. Periodic adjustment was required throughout the run. Oven and heat trace temperatures were monitored with thermocouples throughout the run. The analyses performed are listed in Table 2.

Table 2,
Chemical Analysis Plan

Slag from combustion chamber.	Slag will be collected when all testing is completed and the lower dome is removed. ITEA will collect and analyze slag samples as they decide.
Fine particulates on heated filter	Particulate will be analyzed for fixed carbon and metals by ICP-MS (inductively coupled plasma emission – mass spectrometry).
Rinsate from H2SO4 collection filter	Analyze for sulfate, sulfite, and total sulfur, by IC (ion chromatography).
Steam condensate (ice bath)	Analyze for metals by ICP-MS, and sulfate/sulfite, nitrate/nitrite by IC
Permanent gases	Analyze for O2, N2, CO2, CO, CH4, H2, and SO2 by gas chromatography.

Results

The testing suffered numerous delays due to operational problems (note this is a brand new reactor), and delays due to severe weather conditions (note the reactor was located outdoors). In the end, only two of planned six parametric cases were tested, along with one blank run using pure methane instead of coal. After the two successful test cases were run, the steam-cooled fuel feed lance that feeds the coal slurry into the reactor failed catastrophically causing a rapid loss of pressure in the reactor. ITEA personnel decided that it would not be prudent to use a space lance until the root cause of failure could be determined. It was decided to terminate the testing at this point and make arrangements to move the plant to storage site until a more permanent site could be found. The two test cases did prove to be good experimental runs and the data are presented here. The two cases are for the sub-bituminous coal at 15 and 32 bars pressure.

Fine Particulates

No particulates were found in the filters for the two coal runs. The cartridge elements were clean and white with zero net mass gain. For blank run a 47mm flat round filter was used. It proved to be very fragile and was damaged while opening the filter holder. It did show some red-brown discoloration which was attributed to traces of rust dust from inside the pressure vessel sampling port.

SO_3/H_2SO_4

Table 3 shows some of the analytical results along with corrections for dilution and flow rates. The results for SO_3 are expressed as ppmv in the both the wet and dry sampled gas. The coal feed rate was not provided so we cannot provide results for SO_3 formed per unit mass of coal burned. Since the sulfite numbers are zero, we can compare the measured SO_3 with the value calculated using the total sulfur. The agreement is good: 5.1 ppm (dry gas) versus 6.3 ppm for the 15 bar case, and 19.4 ppm versus 22.1 ppm for the 32 bar case. This indicates that virtually all sulfur collected in the coalescing filter is from SO_3.

Unfortunately not all of the SO_3 in the raw gas was captured in the Teflon filter. Table 4 shows the results of sulfur and nitrogen ions found in the steam condensate. The numbers are converted from aqueous concentrations to what the gas phase concentration (in ppm) would have been prior to absorption in the condensate. Sulfite ions would have come from dissolved SO_2 gas so that is not of concern. However sulfate ions would have come from SO_3 gas, which is a concern. For the 15 and 31 bar cases the SO_3 captured by the condensate represents 31 ppm and 130 ppm (dry gas basis) respectively. Clearly more of the SO_3 gas made it through the coalescing filter than was captured in it. In the planning stages ITEA expressed concern that capturing SO_3 gas in steam condensate could lead to erroneous results due to the reaction with nitrates as shown in Equation 1.

$$NO_3^- + SO_3^{2-} \leftrightarrow NO_2^- + SO_4^{2-} \qquad (1)$$

The reaction occurs after absorption in water, but the end result is that the presence of NO_2 gas (the precursor to aqueous NO_3^-) can cause the formation of SO_4^{2-} ions in solution and thus lead to a higher measure of SO_3 gas. However note that no nitrate (NO_2^-) was found in the condensate. Reaction 1 would have produce nitrate. The absence of it suggests that Reaction 1 did not influence the sulfate level and thus did not affect the measurement of SO_3 gas. (continued on page 7)

Table 3. Dilution Calculations and Data Workup
(key findings are shown in bold)

Sample date, time	15-Feb	17-Feb	17-Feb
Start time	09:55 am	12:15 pm	3:35 pm
Coal	Blank	PRB	PRB
Reactor pressure, bar	7	15	32
Flow totalizer start, liters	1285.8	1466	1745.5
Flow totalizer end, liters	1466	1745	1936
Net dry gas sampled, actual liters	180.2	279	190.5
Ambient temp, C	4	10	15
Gas sampled, std liters @25C	**193.86**	**293.79**	**197.11**
Steam condensate, g	**279.4**	**734.6**	**579.6**
Condensate as std liters gas	379.26	997.16	786.76
Gas + steam, std liters	573.12	1290.95	983.87
% water vapor in sample	**66.17%**	**77.24%**	**79.97%**
Coalescing filter tare, g	7.566	7.6252	8.0957
Filter plus DI water rinse	100.2	100.3	100.2
DI rinse water, g	92.634	92.6748	92.1043
Particulate filter tare, g	0.0924	3.8243	3.4598
Particulate filter plus aerosols, g	damaged	3.8243	3.4598
Net particulates collected, g	0	0	0
Huffman Labs Analysis Results			
Sulfate, mg/L, as SO_4--	0	64	163
Sulfite, mg/L, as SO_3--	0	0	0
Total S, mg/L, as S	0	26	62
Sulfate mass collected, mg	0	5.9311872	15.013001
Sulfate mass, g		0.0059312	0.015013
SO_3 gas in std liters	0	0.0015116	0.0038262
SO_3 in dry gas, ppmv	**0**	**5.1**	**19.4**
SO_3 in wet gas, ppmv	**0**	**1.2**	**3.9**
Alternately if we take the numbers for total sulfur and convert to SO3(g)			
sulfur mass collected, mg		2.4095448	5.7104666
sulfur mass collect, g		0.0024095	0.0057105
sulfur as SO_3 gas, std liters		0.0018413	0.0043638
SO_3 in dry gas, ppmv		**6.3**	**22.1**
SO_3 in wet gas, ppmv		**1.4**	**4.4**

Table 4. Analysis of Steam Condensate for NOx and SOx ions.

These results are for the sulfates and nitrates found in the final steam condensate (condensed at zero celcius)				
*units are mg of analyte per liter solution				
Huffman Sample Number :		119814-01	119814-02	119814-03
Client Sample Number :		GA Tech Condensate #1	GA Tech Condensate #2	GA Tech Condensate #3
		Blank Run, Feb 15	First Coal Run, Feb 17A	Second Coal Run, Feb 17B
		Sample	Sample	Sample
		Conc	Conc	Conc
	ANALYTE	mg/L*	mg/L*	mg/L*
	Nitrate	67	5	26
	Nitrite	1	<1	<1
	Sulfate	1	49	174
	Sulfite	<1	76	40
Condensate collected, liters (Table 3)		0.2794	0.7346	0.5796
Total dry gas sampled, std L (Table 3)		193.86	293.79	197.11
Convert above data to ppm of corresponding gas phase species				
Nitrate, NO_3^-	**from NO_2**	**38.09**	**4.93**	**30.15**
Nitrite, NO_2^-	**from NO**	**0.77**	**X**	**X**
Sulfate, SO_4^{2-}	**from SO_2**	**0.37**	**31.21**	**130.33**
Sulfite, SO_3^{2-}	**from SO_2**	**X**	**58.09**	**35.95**

Therefore we feel the sulfate found in the condensate is from SO_3 in the sampled gas and we must include this number with the sulfate found in the coalescing filter. Therefore the final results for SO_3 (the precursor to sulfuric acid gas H_2SO_4) in the raw sampled gas are shown in Table 5.

Table 5. SO_3 gas concentrations in the raw syngas

Sample date, time	15-Feb	17-Feb	17-Feb
Start time	09:55 am	12:15 pm	3:35 pm
Coal	Blank	PRB	PRB
Reactor pressure, bar	7	15	32
SO_3 from filter, ppmv in dry gas	0	5.1	19.4
SO_3 from condensate, ppmv in dry gas	0.37	31.21	130.33
Total SO3, ppmv in dry raw gas	**0.37**	**36.3**	**149.7**

Steam Condensate

In addition to the absorbed SOx and NOx, the condensate was analyzed by ICP-MS for 60 elements, and weighed to determine the steam concentration in the raw gas. ITEA reported that we should expect 95% water vapor in the sampled gas. We found 66% for the blank case, and 77% and 80% for the 15 bar and 32 bar cases respectively, as shown in Table 3. ITEA would have to verify the steam and coal flow rates to check the overall mass balance. GTRC was not given feed rate data. 80% water vapor is drastically different from 95%. One explanation could be that the sampling port was located too close to the steam injection point and thus the streams had not mixed completely prior to being sampled.

The condensate was analyzed by ICP-MS. The raw data are shown in a table in Appendix A. Many of the metals tested for were either not present or below the detection limit. The metals that were found are summarized in Table 6 below. Concentrations are corrected to micrograms per std liter dry gas.

Table 6. Metals found in steam condensate.

IC-MS for metals found in the final steam condensate (condensed at zero Celsius)					
Client Sample Number :			GA Tech Cond #1	GA Tech Cond #2	GA Tech Cond #3
			Blank	Feb 17 A	Feb 17 B
ATOMIC	ELEMENT	ELEMENT	<<<Converted to micrograms per dry std liter gas>>>		
NUMBER	NAME	SYMBOL	ug/std liter	ug/std liter	ug/std liter
5	Boron	B	12.5964	10.1018	25.8463
11	Sodium	Na	1.3937	0.0825	0.0647
12	Magnesium	Mg	0.1196		
13	Aluminum	Al	0.1902		
14	Silicon	Si	0.2983	3.4006	2.1494
16	Sulfur	S	0.2003	2000.3541	335.2081
19	Potassium	K	0.2248	0.1300	
20	Calcium	Ca	1.5277	0.0825	0.1794
24	Chromium	Cr	0.0130	0.0150	0.0088
25	Manganese	Mn	0.2738	0.0775	0.0353
26	Iron	Fe	0.1686	0.1900	0.1999
27	Cobalt	Co	0.0019	0.0030	
28	Nickel	Ni	0.0130	0.0125	
29	Copper	Cu	0.0043	0.0025	
30	Zinc	Zn	0.3156	0.0425	0.0676
47	Silver	Ag	0.0003		
48	Cadmium	Cd	0.0010		
50	Tin	Sn	0.0159	0.0008	0.0282
56	Barium	Ba	0.0039		
58	Cerium	Ce	0.0003		
82	Lead	Pb	0.0037	0.0008	

The most obvious observation is that most of the metals in Table 6 are present in no greater amounts in the coal runs than for the blank run. This would suggest that the majority of the inorganic species present in the coal were retained in the slag. Or it could be the result of non-isokinetic sampling. ITEA will have to report those findings based on the coal and slag compositions.

Permanent Gases

The gas leaving the ice-water condenser was analyzed by an Agilent micro-GC for H_2, O_2, N_2, CO, CO_2, CH_4, and SO_2. These results are shown below in Table 7. The most obvious numbers are for oxygen. For both the 15 and 32 bar cases, the O_2 levels are in the 20-30% range. This is far too much oxygen for a real combustor. One would typically expect to find a few percent excess O_2 in a stack gas. These high oxygen levels may have contributed to the SO_3 concentration. Some sophisticated combustion modeling would be required to address this question. Also the amount of nitrogen is far too high for a real oxygen-blown combustor. But this can be explained as ITEA was using it as a diluting gas somewhere in the feed system.

Table 7. Dry Gas Composition

Sample date	Sample time	Pressure bar	**Feedstock**	**Label**	H_2 %	O_2 %	N_2 %	CO_2 %
15-Feb		7	**Blank**	**blank_001**	0.007	5.191	83.747	7.956
17-Feb	12:45	15	**Wyoming**	**12_45_15bar**	0.007	23.965	36.267	38.081
17-Feb	1:10	15	**Wyoming**	**1_10_15bar**	0.007	24.724	34.922	38.964
17-Feb	1:33	15	**Wyoming**	**1_33_15 bar**	0.005	25.841	32.422	38.604
17-Feb	4:06	31	**Wyoming**	**4_6_32bar**	0.006	34.332	26.789	36.515
17-Feb	4:48	31	**Wyoming**	**16_48_31bar**	0.008	29.485	28.225	40.245

Table 7 (continued)

Feedstock	**Label**	CO_2 %	CO %	CH_4 %	SO_2 %	total	Unaccounted gas (mol%)
Blank	**blank_001**	7.956	BDL	BDL	0.106	97.007	2.993
Wyoming	**12_45_15bar**	38.081	BDL	0.002	0.501	98.821	1.179
Wyoming	**1_10_15bar**	38.964	0.001	0.002	0.517	99.137	0.863
Wyoming	**1_33_15 bar**	38.604	0.001	0.003	0.541	97.416	2.584
Wyoming	**4_6_32bar**	36.515	0.001	0.004	0.719	98.367	1.633
Wyoming	**16_48_31bar**	40.245	0.001	0.005	0.617	98.586	1.414

Conclusions

GTRC was contracted to 1) provide a host site for the UPA combustor which met all environmental and safety requirements of the U.S. DOE., and 2) to sample and analyze hot combustion gases while UPA operated the combustion under a variety of conditions. Six conditions (3 coals and 2 pressures) were selected for a total of six experimental runs, plus any required blank runs. The reactor was set up, commissioned and operated. Operational and weather-related difficulties, and ultimately a failed fuel injection lance limited the experiments to two runs; a sub-bituminous coal (Powder River Basin) at two pressures.

The slipstream gas sample contained no fine particulate matter in either coal run. This could be the result of the fine particulate bonding with the slag in the combustion chamber, which would be a very good thing in terms of gas clean-up in the full scale design. Or it could be the result of the design of the sampling inlet which did not provide for an isokinetic sample.

SO_3 was sampled by a method consistent with EPA method 8A. However SO3 was ultimately found in the hot filter and in the gas (later condensed) that passed through the filter. For the 15 bar run, 36 ppm of SO_3 was found in the sampled gas (dry gas basis), and 150 ppm was found for the 32 bar experiment. These levels are likely too high for the full scale combustor but since this is a new type of reactor, it may be an acceptable level. ITEA, who owns the technology, will have to make this determination.

The raw gas was purported to consist of 95% water vapor, but only 80% water vapor was found. Assuming the flow of coal and steam were correct, the discrepancy could be due to incomplete mixing of the steam with the hot gas leaving the combustion chamber. The sampling port was located just downstream of the steam injection point.

Between 97 and 99% of the permanent gas was accounted for in the GC analysis. Oxygen levels were found to be very high (20-30%) in the dry gas. It would be reasonable to expect lower SO_3 levels at lower oxygen feed rates (i.e. lower oxygen/fuel ratio).

June 1, 2014
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Appendix A. Raw Analytical Results

Table A1. Analysis of Steam Condensate for Metals

ICP-MS for metals found in the final steam condensate (condensed at zero celcius)					
Huffman Sample Number :			119814-01	119814-02	119814-03
Client Sample Number :			GA Tech Cond #1	GA Tech Cond #2	GA Tech Cond #3
			Blank Run, Feb 15	Coal Run, Feb 17A	Coal Run, Feb 17B
			Sample	Sample	Sample
ATOMIC	ELEMENT	ELEMENT	Conc	Conc	Conc
NUMBER	NAME	SYMBOL	ug/L	ug/L	ug/L
1	Hydrogen	H	NA	NA	NA
2	Helium	He	NA	NA	NA
3	Lithium	Li	1	<1	<1
4	Beryllium	Be	<1	<1	<1
5	Boron	B	8740	4040	8790
6	Carbon	C	NA	NA	NA
7	Nitrogen	N	NA	NA	NA
8	Oxygen	O	NA	NA	NA
9	Fluorine	F	NA	NA	NA
10	Neon	Ne	NA	NA	NA
11	Sodium	Na	967	33	22
12	Magnesium	Mg	83	<2	<2
13	Aluminum	Al	132	<10	<10
14	Silicon	Si	207	1360	731
15	Phosphorus	P	<10	20	<10
16	Sulfur	S	139	800000	114000
17	Chlorine	Cl	NA	NA	NA
18	Argon	Ar	NA	NA	NA
19	Potassium	K	156	52	<50
20	Calcium	Ca	1060	33	61
21	Scandium	Sc	<0.1	<0.1	<0.1
22	Titanium	Ti	<1	<1	<1
23	Vanadium	V	<1	<1	<1
24	Chromium	Cr	9	6	3
25	Manganese	Mn	190	31	12
26	Iron	Fe	117	76	68
27	Cobalt	Co	1	1	<1
28	Nickel	Ni	9	5	<1
29	Copper	Cu	3	1	<1
30	Zinc	Zn	219	17	23

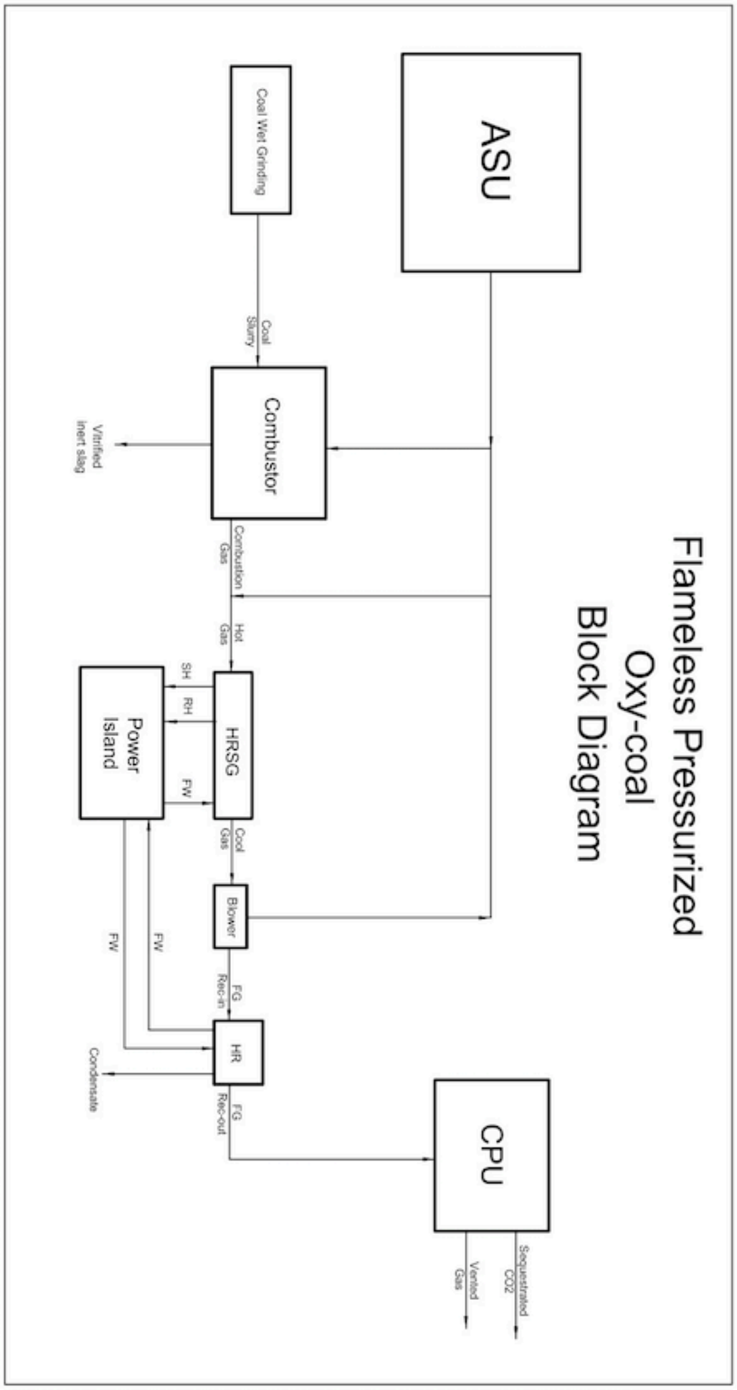
ICP-MS for metals found in the final steam condensate (condensed at zero celcius)					
Huffman Sample Number :			119814-01	119814-02	119814-03
Client Sample Number :			GA Tech Cond #1	GA Tech Cond #2	GA Tech Cond #3
			Blank Run, Feb 15	Coal Run, Feb 17A	Coal Run, Feb 17B
			Sample	Sample	Sample
ATOMIC	ELEMENT	ELEMENT	Conc	Conc	Conc
NUMBER	NAME	SYMBOL	ug/L	ug/L	ug/L
31	Gallium	Ga	<0.1	<0.1	<0.1
32	Germanium	Ge	<0.1	<0.1	<0.1
33	Arsenic	As	<0.1	<0.1	<0.1
34	Selenium	Se	<0.1	<0.1	<0.1
35	Bromine	Br	NA	NA	NA
36	Krypton	Kr	NA	NA	NA
37	Rubidium	Rb	<0.1	<0.1	<0.1
38	Strontium	Sr	<3	<3	<3
39	Yttrium	Y	<0.1	<0.1	<0.1
40	Zirconium	Zr	<0.1	<0.1	<0.1
41	Niobium	Nb	<0.1	<0.1	<0.1
42	Molybdenum	Mo	<0.5	<0.5	<0.5
43	Technetium	Tc	NA	NA	NA
44	Ruthenium	Ru	<0.1	<0.1	<0.1
45	Rhodium	Rh	<0.1	<0.1	<0.1
46	Palladium	Pd	<0.1	<0.1	<0.1
47	Silver	Ag	0.2	<0.1	<0.1
48	Cadmium	Cd	0.7	<0.1	<0.1
49	Indium	In	<0.1	<0.1	<0.1
50	Tin	Sn	11	0.3	9.6
51	Antimony	Sb	<0.2	<0.2	<0.2
52	Tellurium	Te	NA	NA	NA
53	Iodine	I	NA	NA	NA
54	Xenon	Xe	NA	NA	NA
55	Cesium	Cs	<0.1	<0.1	<0.1
56	Barium	Ba	2.7	<0.5	<0.5
57	Lanthanum	La	<0.1	<0.1	<0.1
58	Cerium	Ce	0.2	<0.1	<0.1
59	Praseodymium	Pr	<0.1	<0.1	<0.1
60	Neodymium	Nd	<0.1	<0.1	<0.1

ICP-MS for metals found in the final steam condensate (condensed at zero celcius)					
Huffman Sample Number :			119814-01	119814-02	119814-03
Client Sample Number :			GA Tech Cond #1	GA Tech Cond #2	GA Tech Cond #3
			Blank Run, Feb 15	Coal Run, Feb 17A	Coal Run, Feb 17B
			Sample	Sample	Sample
ATOMIC	ELEMENT	ELEMENT	Conc	Conc	Conc
NUMBER	NAME	SYMBOL	ug/L	ug/L	ug/L
61	Promethium	Pm	NA	NA	NA
62	Samarium	Sm	<0.1	<0.1	<0.1
63	Europium	Eu	<0.1	<0.1	<0.1
64	Gadolinium	Gd	<0.1	<0.1	<0.1
65	Terbium	Tb	<0.1	<0.1	<0.1
66	Dysprosium	Dy	<0.1	<0.1	<0.1
67	Holmium	Ho	<0.1	<0.1	<0.1
68	Erbium	Er	<0.1	<0.1	<0.1
69	Thulium	Tm	<0.1	<0.1	<0.1
70	Ytterbium	Yb	<0.1	<0.1	<0.1
71	Lutetium	Lu	<0.1	<0.1	<0.1
72	Hafnium	Hf	<0.1	<0.1	<0.1
73	Tantalum	Ta	<0.1	<0.1	<0.1
74	Tungsten	W	<0.2	<0.2	<0.2
75	Rhenium	Re	<0.1	<0.1	<0.1
76	Osmium	Os	NA	NA	NA
77	Iridium	Ir	NA	NA	NA
78	Platinum	Pt	<0.1	<0.1	<0.1
79	Gold	Au	<0.3	<0.3	<0.3
80	Mercury	Hg	<0.1	<0.1	<0.1
81	Thallium	Tl	<0.2	<0.2	<0.2
82	Lead	Pb	2.6	0.3	<0.1
83	Bismuth	Bi	<0.1	<0.1	<0.1
90	Thorium	Th	<0.1	<0.1	<0.1
92	Uranium	U	<0.1	<0.1	<0.1

NA - element not analyzed or quantified by the ICP-MS scans performed.

Appendix 3

FPO process diagram



Appendix 4

FPO Firing Section 3D prospect