

FINAL TECHNICAL REPORT  
September 1, 2005, through August 31, 2006

Project Title: **ON THE FEASIBILITY OF HEAT RECOVERY FROM OXY-COAL COMBUSTION PROCESSES**

ICCI Project Number: 05-1/ER7  
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ABSTRACT

The overall project objective was to investigate the feasibility of an oxy-combustion process for coal fired power plants. While this process is expected to reduce pollution generation and removal costs, it is hampered by the power required to operate an air separation plant. An additional objective was to investigate opportunities to further integrate the two processes (the power plant and the air separation unit), and thus achieve additional energy conservation.

The results of a thermodynamic analysis are as follows. Four cases were considered, in order of increasing retrofit costs: 1) oxy-combustion with  $CO_2$  recycle, 2) oxy-combustion without  $CO_2$  recycle, 3) oxy-combustion with  $CO_2$  recycle and with heat integration, 4) oxy-combustion without  $CO_2$  recycle and with heat integration. From the perspective of overall power plant efficiency we found the following for each case: 1) 5.75% loss, 2) 0.3% gain, 3) 4.85% loss and 4) 1.2% gain. These results suggest that  $CO_2$  recycle suffers greatly with respect to operational costs. They also suggest that the heat recovery can improve efficiency by 0.9% regardless of the recycle option used.

Regarding pollution control, a qualitative analysis indicated that Illinois coal would benefit more than other (low sulfur/nitrogen/ash) coals. However, this benefit was due to the large costs currently incurred by Illinois coal, and thus its greater opportunity for savings.

## EXECUTIVE SUMMARY

The notion of oxy-combustion is to burn fuel in the presence of pure oxygen rather than air. Potential benefits include: reduced formation of  $NO_x$  pollutants, reduce sensible heat losses (the usual  $N_2$  portion of the flue gas need not be heated), lower flow rate to the pollution control devices and a nearly pure stream of  $CO_2$  available for sequestration. On the downside, oxy-combustion will significantly increase flame temperature and likely require a redesign or retrofit of the boiler. In lieu of these retrofit costs many suggest  $CO_2$  recycle as a means to reduce flame temperature. Finally, the power required operate the Air Separation Unit (ASU) will adversely impact the plant efficiency.

The first goal the project was to investigate, from an energy standpoint, the impact of operating a coal fired power plant using oxy-combustion, and contrast the various operating scenarios. The second objective was to search for heat recovery opportunities made possible by operation of the ASU in the vicinity of the power plant.

Toward the first objective, the results are as follows. A 5.75% loss in overall efficiency is expected in the  $CO_2$  recycle case. In the non-recycle case, a minor increase ( $\sim 0.3\%$ ) is expected (the reduction in sensible heat losses about equals the power required to operate the ASU). Concerning heat recovery, a simple insertion of a power cycle (between the flue gas and an available cryogenic stream) resulted in a 0.1% improvement in plant efficiency. As this option seemed unreasonable we searched for alternatives. The result was the development of a novel process that involved the diversion of a liquid  $N_2$  stream to be used as a once through working fluid in a standard supercritical power cycle. In this case, a 0.9% improvement in plant efficiency was projected, assuming minimal perturbation of the ASU. However, removal of this assumption suggested that as much as a 2.2% improvement may be possible.

## OBJECTIVES

The overall objective of the project is to develop a heat recovery system for oxy-coal combustion processes. The tasks scheduled for this year are as follows. Task 1 – Determine the thermodynamic feasibility of the proposed process (Task 1a: 1<sup>st</sup> and 2<sup>nd</sup> law analysis, Task 1b: economic analysis concerning heat exchange surface areas). Task 2 – Design of experiments to address key operational issues of the recovery unit.

## INTRODUCTION AND BACKGROUND

The notion of oxy-combustion is to burn fuel in the presence of pure oxygen rather than air. While there are many potential benefits to this notion, the primary motivation in the scenario of a coal fired power plant is to reduce the formation of  $NO_x$  pollutants. (The general idea is as follows. Removal of  $N_2$  from the flame leaves fuel bound nitrogen as the only route. Additionally, while the increase in flame temperature associated with oxy-combustion would suggest greater  $NO_x$  production, a reduction in O/C ratio, to about stoichiometric conditions without reducing carbon conversion, results in a blocking of the fuel bound path.) A second benefit of oxy-combustion is to reduce sensible heating losses. Specifically, eliminating the need to heat the  $N_2$  portion of the flue gas suggests that less coal is needed to deliver an equal amount of power to the steam turbine. However, this aspect will also result in a significant increase in flame temperature and thus may require a redesign and/or new metallurgy for the boiler components. Toward reducing flame temperature and thus retrofit costs, flue gas recirculation has been advocated, which in the oxy-combustion case is denoted as  $CO_2$  recycle. This option, however, is expected to eliminate most if not all of the reductions in sensible heat loss. The third benefit concerns the flue gas composition at the pollution control devices. Specifically, elimination of  $N_2$  from the flue gas stream will lower its flow rate (which decreases baghouse power costs and increases scrubber residence time) as well as increase the concentration of trace elements and thus improve removal driving forces. At the final stage, the flue gas stream will consist almost entirely of  $CO_2$ , and thus will be in an excellent position for sequestration.

On the subject of Illinois coal, it is typically characterized as having high carbon content (i.e. more energy per lb), but also containing larger amounts of sulfur, nitrogen and ash, all of which increase the size and cost of operating pollution control devices. Thus, the main driver for not using Illinois coal is the cost of pollution control. Turning to the envisioned oxy-combustion process, the expected reduction in flue gas flow rate (due to the removal of  $N_2$ ) will reduce the burden to pollution control devices for all types of coal. However, since Illinois coal requires more pollution control, a reduction in burden will provide greater benefit. Stated another way, the envision reduction in pollution control burden for all coal types will level the playing field. This along with the higher carbon content could push Illinois coal back into favor with power producers. (Please note that a detailed analysis of pollution control devices is outside the scope of the current project. We, however, will focus on the bulk efficiency aspects of oxy-combustion, which is perceived as the first hurdle toward adoption of the process.)

While the obvious downside of oxy-combustion is the cost/power required to operate the air separation unit, the fact that this unit will operate in the vicinity of the boiler creates an additional opportunity for energy conservation. Specifically, the flue gas stream at the exit of the boiler contains a significant amount of energy, even in the oxy-combustion case. The problem is that this energy is of low quality (around 400-600C) and difficult to convert to useful work, especially if an ambient temperature heat sink is used for the power cycle. However, access to the ultra-low temperatures of a cryogenic air separation unit may provide a more efficient route to the recovery of flue gas heat. Investigating the feasibility of such a recovery device is the primary goal of the project.

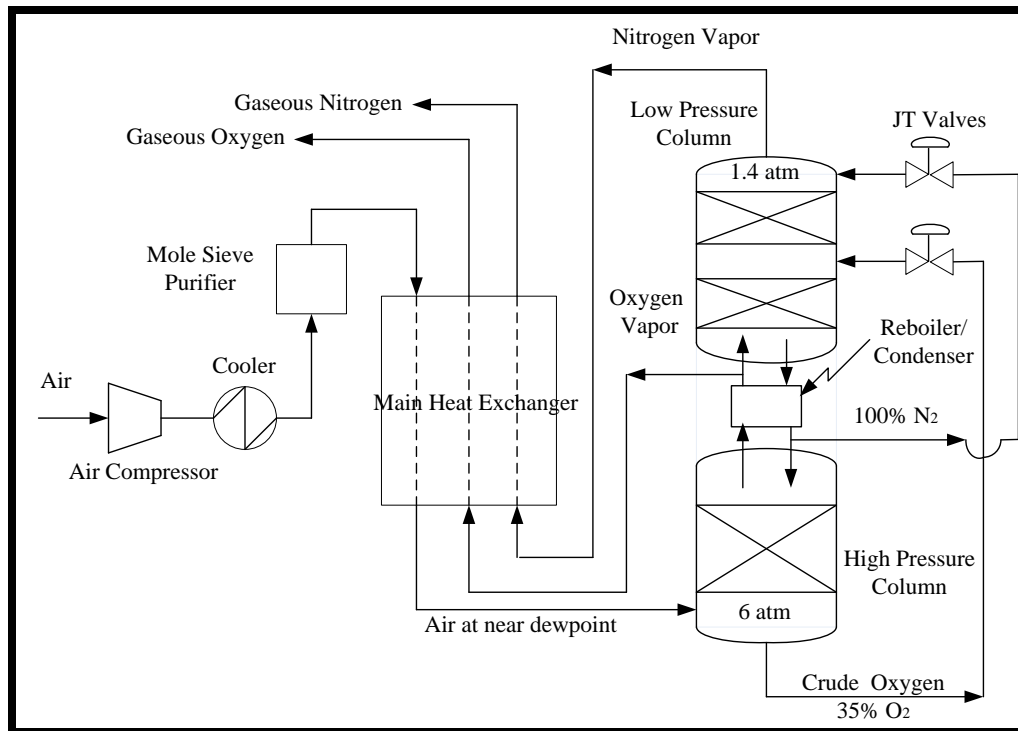


Figure 1. Cryogenic double-column for air separation (adapted from Wilson, 2000).

As the design of the envisioned heat recovery unit will require some modification of the Air Separation Unit (ASU), a description of the existing technology is in order (see figure 1). Ambient air is compressed in a multistage compressor to 6 atmospheres. Due to inter-stage cooling the power cost of this compression is about equal to that of isothermal compression, and represents most of the power required to operate the ASU. (Assuming the compression temperature to be 298K and a compressor efficiency of 90%, results in power requirement of 166.4 kJ / kg of air to be separated.) After purification the pressurized air is cooled in the main heat exchanger to near its dew point ( $\sim 105$  K) and fed to the high pressure column, where it serves as a re-boil stream. The liquid bottom stream from this column (denoted as crude  $O_2$ ) is sent to the low pressure column, where it serves as the feed stream ( $\sim 35\%$   $O_2$ ). Returning to the high pressure column, about 60% of the condensed  $N_2$  vapor at the top of the column is used for reflux, while the balance is sent to the low pressure column, where it also serves as liquid reflux. Creating

these liquid reflux streams of nearly pure  $N_2$  is the key to the entire process, which is achieved by matching the high pressure condenser with the re-boiler of the low pressure column (the saturation temperature of  $N_2$  at 6 atm is  $\sim 95\text{K}$ , while that of  $O_2$  at 1.4 atm is  $\sim 90\text{K}$ ). As suggested above, the product streams of the low pressure column are nearly pure (both over 99.5%), and exit in the vapor phase (given the state of the air feed, this is expected). Finally, the product streams are sent back to the main heat exchanger, and exit at near ambient conditions.

A quick analysis of the main heat exchanger indicates the following. Cooling 1 kg of pressurized air from 298 K to 105 K requires 209 kJ. On the heating side, 44 kJ is required to bring 0.23 kg of 1 atm  $O_2$  from 90 K to 298 K, while 181 kJ is required to bring 0.77 kg of 1 atm  $N_2$  from 77 K to 298 K. This indicates that 16 kJ / kg of air may be available for the heat recovery system. Clearly, we could not use all of this heat sink, as it would require a infinite surface area in the main heat exchanger, however it does provide an upper bound on what is possible. It should also be noted that this number can be increased by increasing the air feed pressure (and combining with a low temperature expander). This, however, will significantly increase compression costs and was quickly abandoned as an option. It is finally noted that argon recovery is not included in the analysis. However, the addition of argon recovery equipment is expected to result in only minor changes to the process conditions of the double column arrangement described above.

## EXPERIMENTAL PROCEDURES

No experimental activities were performed during the project.

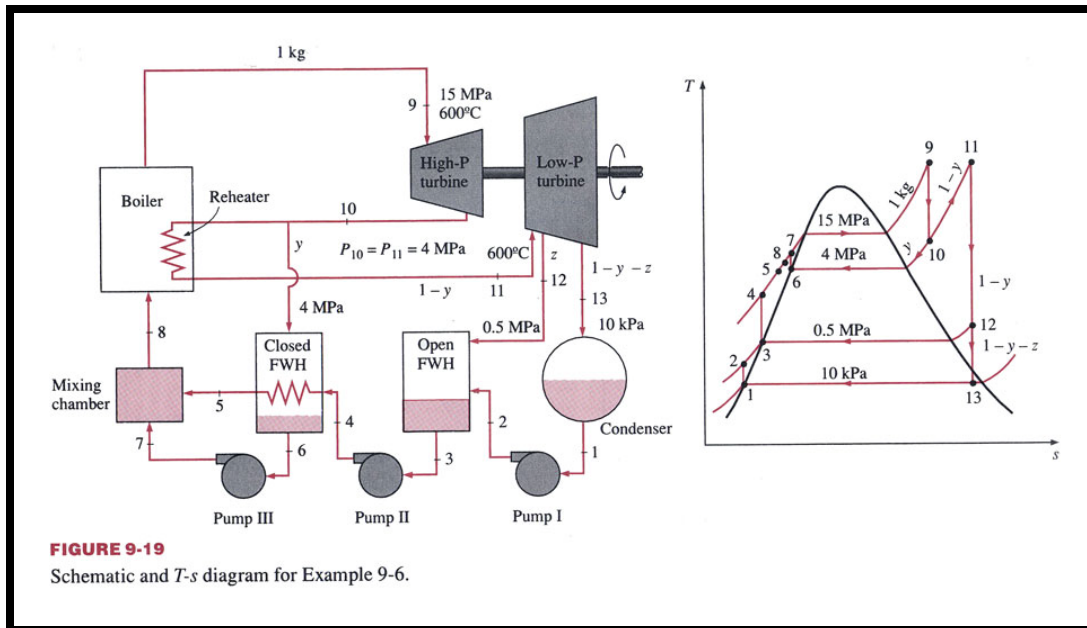


Figure 2. Steam side of the power plant (from example 9-6 of Cengel & Boles, 1998).

The approach / assumptions used in the thermodynamic analysis of the coal fired power plant are as follows. In general, the plant contains two process sides; the steam side and the flue gas side. Concerning the steam side, fairly typical steam cycle conditions were assumed (depicted in figure 2). Using the detailed results presented in Cengel & Boles (1998), we identified the thermal load on the boiler with respect to intensive conditions (i.e.,  $T$ ,  $P$  and enthalpy on a /kg of steam basis). Using these figures it was determined that 409 kg/s of steam would be required to generate 500MW of electric power (turbine and electric generator efficiencies assumed to be 90% and 95%, respectively). From this flow of steam, the heat transfer requirements on the water walls, superheaters, reheater and economizer were determined.

	Illinois #6 Coal	WPRB (low-S)
Carbon	0.575	0.48
Hydrogen	0.037	0.034
Oxygen	0.08	0.11
Sulfur	0.04	0.0048
Nitrogen	0.01	0.0062
Ash	0.16	0.064
Moisture	0.12	0.30
HHV	10100 BTU / lb	8020 BTU / lb

Table 1. Ultimate analysis of coals considered in the study (mass fraction).

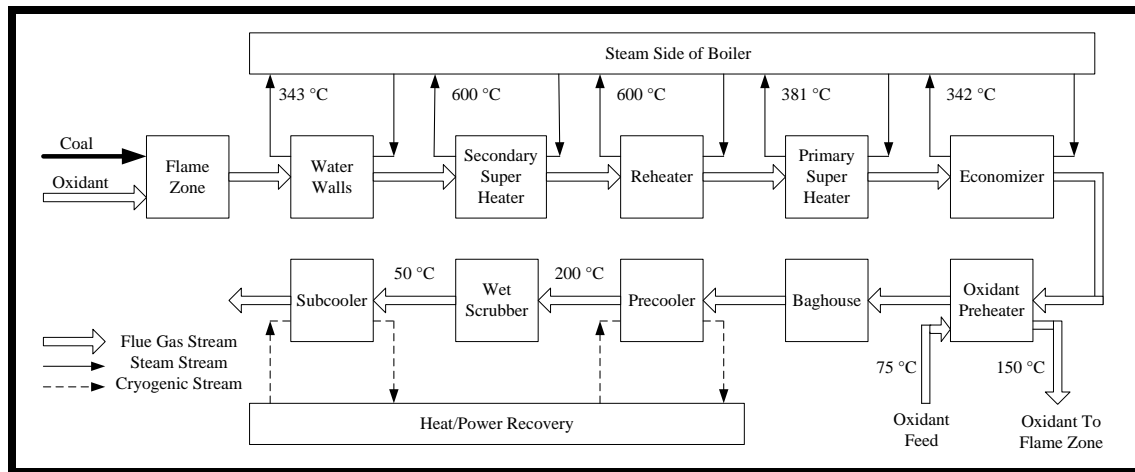


Figure 3. Flue gas side of the power plant (adapted from Stultz & Kitto, 1992).

On the flue gas side (depicted in figure 3), we considered two types of coal, each using 3 types of oxidant feed (see tables 1 and 2). Within the flame zone all reactions were assumed to take place. Further assumptions include; 95% conversion of carbon to  $CO_2$ , 0% conversion of ash and moisture, and 100% conversion of all others (to  $H_2O$ ,  $SO_2$  and  $NO_2$ ). It was also, assumed that 25% of the unconverted carbon and 75% of the ash left

the combustion zone (water walls) via the hopper rather than up the flue. In cases 1a and 2a, the feed rate of air was assumed to be 1.1 times theoretic air. In cases 1b and 2b, the feed rate of oxygen was the same as in the a cases, namely the same air with  $N_2$  removed. (In the actual operation of an oxy-combustion plant the excess oxygen will likely be reduced to 1. However, toward simplicity of the analysis we left this value at 1.1 for all cases.) Concerning  $CO_2$  recycle (the c cases), we simply assumed the conditions of the a cases, but then charged the scenario with the air separation power costs.

Case	Description
1a	Illinois #6 Coal - Air Feed
1b	Illinois #6 Coal - $O_2$ Feed
1c	Illinois #6 Coal - $O_2$ Feed and $CO_2$ Recycle
2a	WPRB Coal - Air Feed
2b	WPRB Coal - $O_2$ Feed
2c	WPRB Coal - $O_2$ Feed and $CO_2$ Recycle

Table 2. Cases analyzed.

Case	Adiabatic Flame Temp.	Water Walls	Secondary Super-heater	Re-heater	Primary Super-heater	Economizer	Oxidant Pre-heater
1a, 1c	1926	1519	1159	964	897	<b>650</b>	591
1b	4345	3308	2297	1705	1490	<b>647</b>	608
2a, 2c	1768	1408	1093	923	864	<b>650</b>	594
2b	3740	2850	2003	1511	1334	<b>647</b>	613
Steam Side Temperature	-	342	600	600	381	<b>342</b>	150

Table 3. Temperature profiles of flue gas in boiler (all in  $^{\circ}C$  and at unit outlet).

To determine the baseline efficiencies, we will need to select an appropriate coal feed rate for each case. While a decrease in coal feed will result in higher efficiency, sufficient heat transfer driving forces should be maintained in each unit. Thus, the coal flow for each case was selected to be such that a  $300^{\circ}C$  temperature difference occurred in the economizer. The resulting temperature profiles are given in table 3, whereas the coal and oxidant flow rates used to generate these profiles are given in table 5. Concerning oxidant pre-heat, it is assumed that low pressure steam will heat to  $75^{\circ}C$  and the flue gas will heat to  $150^{\circ}C$  (as indicated in figure 2). (While this is a fairly low air pre-heat duty, which will bring all baseline efficiencies down, we expect that the existing requirements of low  $NO_x$  operation to warrant such a duty (De Nevers, 1999). In the oxy-combustion cases, where low  $NO_x$  operation is achieved by other means, this low duty requirement could be lifted.

However, again toward simplicity of the analysis, we assumed uniformity with respect to the cases.)

In the baghouse, we assume all remaining ash and unconverted coal is removed adiabatically. Then based on the wet scrubber process described in De Nevers (1999), we assume a pre-cooler is required to achieve a scrubber inlet temperature of 200°C. Given this and the figures of table 4, the pre-cooling duty of each case was calculated. Typically this duty is provided by a pre-stack reheat of the flue gas from the scrubber (exit temperature 50°C) with the balance being provided by cooling water. However, in the context of our recovery system, this duty represents the heat available for conversion to power. In the scrubber, we assume all  $SO_2$  to be removed along with any remaining  $O_2$ . Additionally, we assume all water vapor to be condensed, leaving only  $CO_2$  and  $N_2$  at the exit (see the last column of table 4). In case c, a majority of this flow will be sent back to the boiler as recycle, leaving the figure of case b as the actual exit flow from the plant. This flow represents the  $CO_2$  available for sequestration, which will likely involve an adiabatic compression and cooling to a liquid state for transportation and/or storage.

Case	Baghouse Exit Temp. (°C)	Baghouse Exit Flow (kg/s)	Mass Fraction at Baghouse Exit $CO_2/H_2O/N_2/SO_2$	Pre-cooler Duty (MW)	Scrubber Exit Temp (°C)	Scrubber Exit Flow (kg/s)
1a	591	704	0.22/0.05/0.71/0.009	314	50	650
1b	608	170	0.74/0.17/0.00/0.029	86	50	125
1c	591	704	0.93/0.05/0.00/0.009	314	50	650
2a	594	790	0.22/0.08/0.69/0.001	364	50	716
2b	613	198	0.70/0.25/0.00/0.004	110	50	138
2c	594	790	0.91/0.08/0.00/0.001	364	50	716

Table 4. Post-boiler flue gas conditions.

## RESULTS AND DISCUSSION

Task 1a: The first task of the project was to perform a thermodynamic analysis of the oxy-combustion process without heat recovery. Given the methods of the previous section, the efficiency results are given in table 5. In general, oxy-combustion provides a 6% boost due to reduced sensible heat losses (28 to 33.9 for Illinois coal and 26 to 32.3 for WPRB), but then loses about the same due to ASU power requirements (down to 28.1 and 26.5 for Illinois and WPRB, respectively). In the  $CO_2$  recycle cases (the c cases), ASU power loss is not offset by a reduction of sensible heat losses, and results in a significant loss of efficiency (down to 22.3 and 20.2 for Illinois and WPRB, respectively). Based on these results, there is clearly little operational incentive to incur the retrofit costs associated with an oxy-combustion process (at best 0.1% and 0.5% efficiency gains are possible in the Illinois and WPRB cases, respectively).



Case	Coal Flow (kg/s)	Oxidant Flow (kg/s)	Generated Power (MW)	Efficiency wo/ ASU Power (%)	Air Flow to ASU (kg/s)	Power to ASU (MW)	Generated Power (MW)	Efficiency w/ ASU Power (%)
1a	76	649	500	28.0	0	0	500	<b>28.0</b>
1b	62.8	125	500	33.9	536	89	411	<b>28.1</b>
1c	76	649	500	28.0	536	89	411	<b>22.3</b>
2a	103	710	500	26.0	0	0	500	<b>26.0</b>
2b	83	139	500	32.3	572	95	405	<b>26.5</b>
2c	103	572	500	26.0	572	95	405	<b>20.2</b>

Table 5. Efficiency analysis **without** heat recovery.

Turning to the question of heat recovery, let us begin with the following simple analysis. From table 4 it is clear that a significant amount of heat is still available from the oxy-combustion flue gas (86 and 110MW for the Illinois and WPRB cases, respectively). Returning to the main heat exchanger discussion at the end of the introduction section, at most 16 kJ / kg of air is available to the recovery unit. Scaling up based on the air flow rates of table 4, indicates that ~9 MW is available. Next consider the standard definition of power cycle efficiency:  $\eta = P_{out} / \dot{Q}_{in} = 1 - \dot{Q}_{out} / \dot{Q}_{in}$ . If we then assume a heat sink limited process, it is easily concluded that the maximum recoverable power is  $P_{out} = \dot{Q}_{out} \eta / (1 - \eta)$ . If we then optimistically assume 50% efficiency, the recovered power will be 9MW, which is 10% of the power required to drive the ASU and would result in 0.6% improvement in overall efficiency of the plant. In the more reasonable case of 25% efficiency and a 5MW use of the heat sink, only 1.7MW would be recovered, 1.8% of the ASU power and a 0.1% gain in overall plant efficiency. Thus, it is fairly clear that simple insertion of a power cycle between the flue gas and the cryogenic sink will not provide the recovery goals desired. Based on the above discussion, it is clear that a modification of cryogenic distillation process will be required to achieve a reasonable level of power recovery. In the literature one can find numerous variations of the double column arrangement depicted in figure 1. Unfortunately, all of the processes we found had the problem of requiring more energy to operate (per kg of  $O_2$  produced) or significantly sacrificed oxygen product purity. As such the remainder of the report will focus on the process of figure 1.

As mentioned in the ASU review of the introduction section, the key aspect of the whole process is creation of a high purity liquid  $N_2$  stream to be used for reflux. Furthermore, it is easily concluded that this liquid  $N_2$  is the most precious substance of the process, since creating it by any other means (say liquefaction) is rather costly (recall that saturated  $N_2$  occurs at the lowest temperature of the process, 95 K at 6 atm and 81 K at 1.4 atm). Now, given the importance of liquid  $N_2$ , an inspection of the reflux stream of the low pressure column reveals a something a bit curious (see figure 4). Specifically, after expansion from 6 to 1.4 atmospheres 20% of the stream will be in the vapor phase, suggesting that it goes directly to the vapor product and only 80% of the stream is used for reflux. Now, what if we were to divert this underutilized 20% to the heat recovery process?

Specifically, prior to expansion we would like to divert 20% of this stream to the recovery unit, and then sub-cool the remaining liquid to 81 K so that after expansion it would all remain as liquid to be used for reflux (see figure 5). This, however, hinges on our ability to cool the stream using the 1 atmosphere vapor stream coming from the low pressure column. Using the data from figure 6, the vapor temperature exiting the sub-cooler is easily determined ( $\sim 92$  K). While a heat exchange temperature difference of only 3K seems a bit small (suggesting a large surface area, recall that the combined re-boiler/condenser (of the two column arrangement of figure 1) operates on a 5 K temperature difference while servicing much higher throughputs. (Additional information concerning the above calculation is as follows. The crude  $O_2$  stream is assumed to have an  $O_2$  mole fraction of 0.35, indicating that the liquid  $N_2$  stream from the low pressure column is 40% of the fed air. In the original configuration 79% of the air exits the low pressure column as  $N_2$  vapor, but in the diverted case only 71% of the air exits the low pressure column as vapor. Similarly, only 32% of the fed air needs to be sub-cooled, while the remaining 8% is diverted to the heat recovery system.)

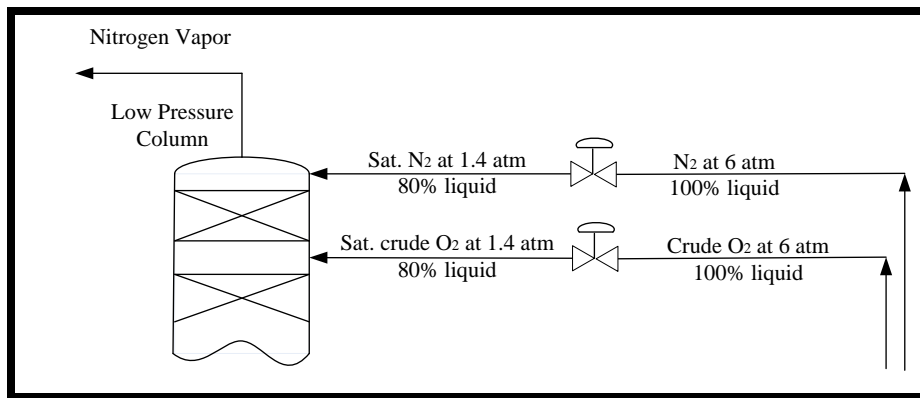


Figure 4. Typical inlet conditions of the low pressure column.

It should also be noted that we selected a 20% diversion as an attempt to minimize perturbations of the original process. That is, from the perspective of the low pressure column no change has occurred. However, it will soon be seen that an increase in this diverted flow will increase the amount of heat we can recover. Thus, it will be useful to determine the maximum amount of liquid  $N_2$  we can divert, or equivalently the minimum reflux needed by the low pressure column. Using Underwood's method (Doherty & Malone, 2001) for minimum reflux (assuming a constant relative volatility of 4), the minimum reflux ratio is found to be 0.56, indicating that the liquid in should be 36% of the vapor out. Then consulting again the figures of the above paragraph, we find that 22% of the air fed must be used as liquid reflux to the low pressure column, and thus 45% of the 6 atm liquid  $N_2$  may be diverted. If we additionally assume sub-cooling on the crude  $O_2$  stream, so that it is fed as saturated liquid, rather than 80% liquid, we find that 49% may be diverted. Clearly, both of these figures represent unrealistic scenarios, as they would require an infinite number of trays in the low pressure column. They do however provide an upper bound that will be useful in determining the maximum amount of heat one may recover from the proposed process.

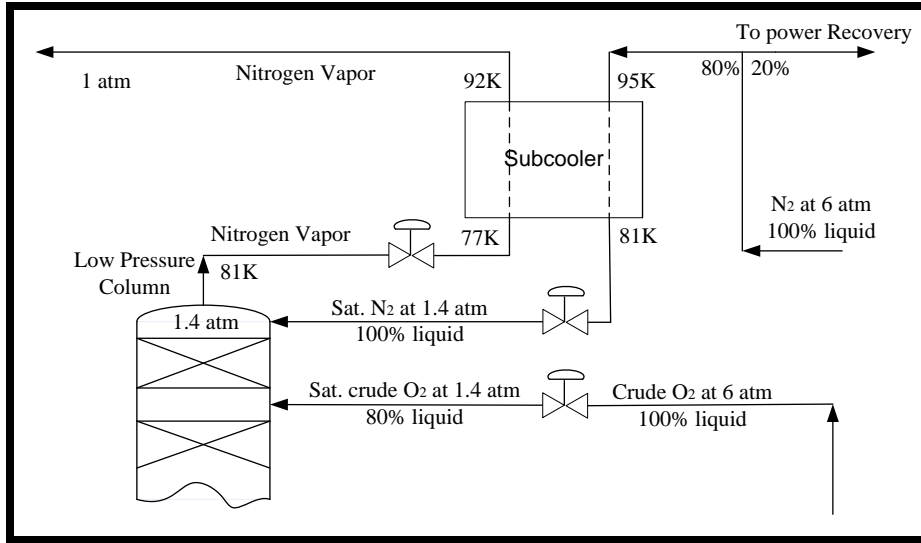


Figure 5. Proposed inlet conditions for low pressure column.

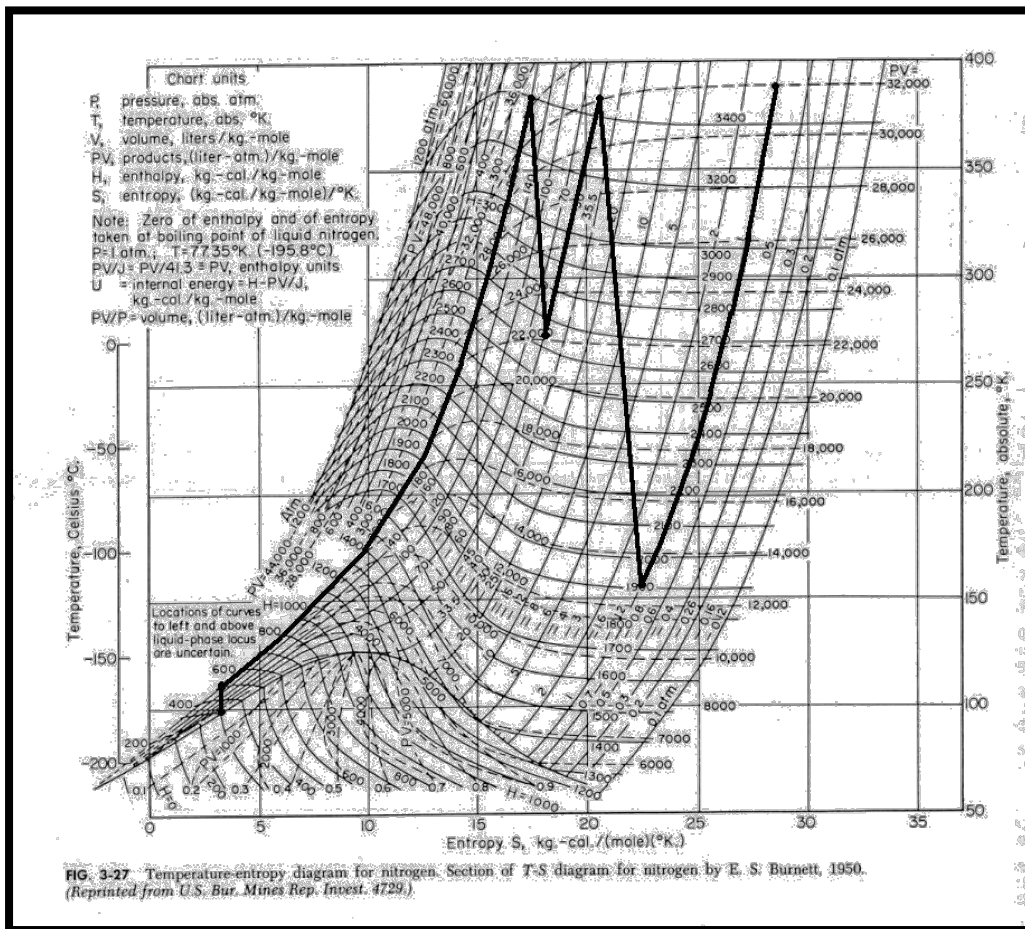


Figure 6. T-s diagram for  $N_2$  in proposed heat recovery unit (from Perry & Green, 1984).

The T-s diagram of figure 6 indicates the proposed use of the diverted liquid  $N_2$  stream. Specifically, it will begin by being pressurized in a liquid pump. It will then be heated using the waste heat of the flue gas. From there any number of turbine expansions and reheats can be applied until it is brought back to atmospheric pressure. Figure 7 highlights the specifics of the curve shown in figure 6 (90% efficiency assumed for the turbines). Summarizing, it is found that 319 kJ of work can be generated per kg of  $N_2$ , while 747.6 kJ of heat is consumed (per kg  $N_2$ ). It should be noted that one may need to pull a portion of this heat from the main heat exchanger, since the flow rate of cool  $N_2$  will be decreased. However, in the case of a 20% diversion (with the sub-cooler), we found that the un-diverted  $N_2$  vapor along with the  $O_2$  vapor was more than sufficient to cool the air feed. (In the original case there was an excess of 16 kJ / kg air, while in the 20% diversion case 12 kJ remained.)

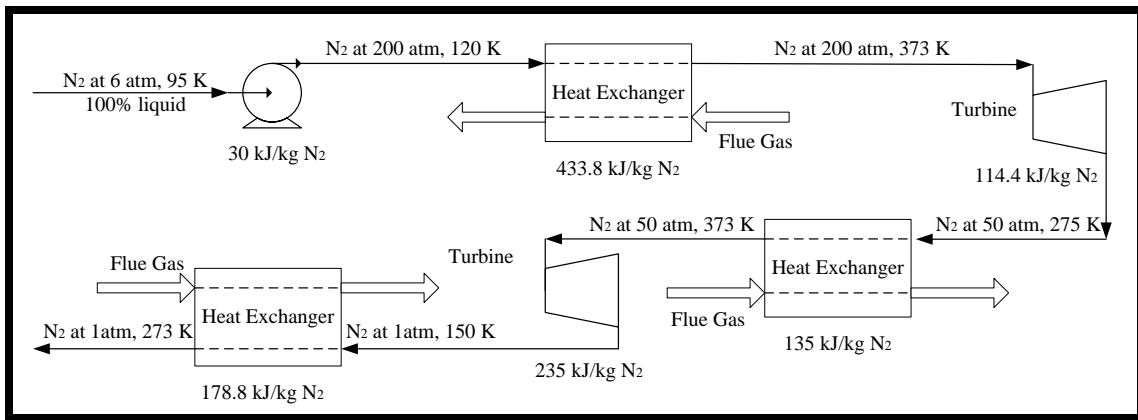


Figure 7. Process flow diagram for the proposed heat recovery unit.

Table 6 presents the scaled up recovery figures based on the air flow rates of the various cases. In the 20% diversion case, we find that  $\sim 14.5$  MW of power is produced (about 16% of the power required to run the ASU), while  $\sim 33$  MW of heat is taken from the flue gas (about 35% of that available). For the 50% diversion case,  $\sim 35.5$  MW is generated (38% of the ASU power), and  $\sim 83$  MW of heat is taken from the flue gas ( $\sim 85\%$  of available). Concerning overall efficiency, we find across the board a 0.9% boost for the 20% diversion case, and a 2.2% boost for the 50% diversion case.

**Task 1b:** Concerning an economic analysis, we did not size any of the proposed heat exchangers. We did, however, keep a close watch on the approach temperatures. In general, large temperature differences are expected in the heat exchangers of the recovery unit (typically in excess of 300K), which will result in fairly small surface area requirements. The exception is with regard to the new sub-cooler of figure 5, which as noted will have a temperature difference of  $\sim 3$ K. It was also noted that such operation is not atypical for the air separation industry, and is likely achievable. It should also be noted that the approach temperature in the main heat exchanger of the ASU has been reduced, and thus will require an increase in surface area.

Case	Air Flow to ASU (kg/s)	Power to ASU (MW)	Efficiency w/ ASU w/o Power Recovery (%)	Power Recovered: 20% Diversion (MW)	Efficiency w/ ASU w/ Power Recovery (%)	Power Recovered: 50% Diversion (MW)	Efficiency w/ ASU w/ Power Recovery (%)
1a	0	0	28.0	0	<b>28.0</b>	0	<b>28.0</b>
1b	536	89	28.1	14	<b>29.0</b>	34	<b>30.4</b>
1c	536	89	22.3	14	<b>23.2</b>	34	<b>24.5</b>
2a	0	0	26.0	0	<b>26.0</b>	0	<b>26.0</b>
2b	572	95	26.5	15	<b>27.4</b>	37	<b>28.7</b>
2c	572	95	20.2	15	<b>21.1</b>	37	<b>22.4</b>

Table 6. Efficiency analysis **with** heat recovery.

Task 2: No design of experiment activities were performed. In addition to time constraints, the envision recovery process has changed significantly from the one outline in the original proposal. As such, many of the operational issues originally identified are no longer of concern and thus eliminated the need for much of this activity.

On the subject of scrubber operation, consider the following analysis. Table 4 gives a qualitative indication of possible improvements. Specifically, the mass flow rate of flue gas to the scrubber will be reduced by about 75%, indicating a four fold increase in the residence time within the scrubber. Additionally, the mass fraction of  $SO_2$  in the oxy-combustion case will increase three to four times the original, while the mass flow of  $SO_2$  actually decreases (due to less coal fed). All of these aspects can be used to reduce the size, power and limestone requirements of the wet scrubber.

## CONCLUSIONS AND RECOMMENDATIONS

Concerning oxy-combustion without heat recovery, it was found that a 5.75% loss in overall efficiency must be charged due to the power requirements of the air separation unit (i.e., the  $CO_2$  recycle case). However, it was also found that all of this (and a bit more) could be recovered by eliminating  $N_2$  sensible heat losses (i.e., the high flame temperature case). Specifically, the net gain for Illinois coal was 0.1% and 0.5% for WPRB (the higher gain for WPRB was due to the fact that WPRB has more sensible heat losses to recover).

Concerning heat recovery through integration with the ultra-low temperatures of the cryogenic air separation unit, two scenarios were considered. The first option was to insert a power cycle between the flue gas (heat source) and an unused cryogenic stream (heat sink). In this case, the gain in overall efficiency was projected to be a mere 0.1%, and maxed out at 0.6%. The second option involved a diversion of liquid  $N_2$  to be used as a once through working fluid in a standard supercritical power cycle. In this case, the gain in overall efficiency was projected to be 0.9%, and maxed out at 2.2%.

The overall impact to power producers is follows. If one is willing to retrofit the boiler to accommodate much higher flame temperatures, then a 1% overall improvement in efficiency is possible. However, in the case of  $CO_2$  recycle (requiring minimal retrofit in the boiler), a 3.5% loss of efficiency is expected.

Concerning the difference between Illinois coal and WPRB, the only difference we found was with respect to pollution control. (While the baseline efficiency numbers were different, the gains and losses for each case were about the same.) Regarding  $SO_2$  removal, no change is expected in the operation of the scrubber for the  $CO_2$  recycle case. However, in the high flame temperature case, flue gas flow through the scrubber will be reduced by 75%, which could significantly reduce capital and operational costs. Since these costs are much higher for Illinois coal, this feedstock is expected to benefit significantly from the technology. A similar argument could be made concerning baghouse operation, owing to the high ash content of Illinois coal. On the subject of  $NO_x$  mitigation, all cases considered should benefit (even the  $CO_2$  recycle case), but again the higher levels of fuel bound nitrogen in Illinois coal suggests it has the most to gain. Finally, all options see an equal benefit concerning the cost of  $CO_2$  sequestration. Namely, a nearly pure stream of  $CO_2$  will be available in all cases.

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