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# Evaluating the Technical and Economic Feasibility of Adding a Power Recovery System to the Steam Condenser of a Lignite Coal-Fired Power Plant

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## Abstract

Steam is the typical working fluid to drive turbo-generators in coal-fired power plants. It is an effective working fluid, but some of its energy is extracted in an unusable form when condensed. A Power Recovery Cycle (PRC) using a more volatile Secondary Working Fluid (SWF) added to the steam cycle could improve energy efficiency. PRCs have been applied to the flue gas and for combined cycle systems but not to traditional plant steam cycles. This paper details an analysis of adding a steam cycle PRC to a 500 MW lignite coal-fired power plant. A validated model of the plant was developed and PRCs using the three most attractive SWFs, benzene, methanol and hydrazine, were then added to the model. Adding a benzene, methanol, or hydrazine steam cycle PRC will produce an additional 59, 34, and 49 MW, respectively. An AACE Class 4 factored broad capital cost estimate and comparable operating costs and revenue estimates were developed to evaluate PRC feasibility. The benzene, methanol, and hydrazine processes had 2019 Net Present Values (NPVs) @12% of −\$32, −\$59, and +\$35 million ± 40%, respectively. Thus, a PRC may be profitable at current or modest increases to U.S. Upper Midwest electricity prices of around \$0.0667/kWh.

## Keywords

Lignite Coal, Heat Recovery, Power Plant, Organic Rankine Cycle, Bottoming Cycle

## 1. Introduction

Coal is a major source of electrical power, accounting for about 19% of the U.S.

supply in 2020 [1]. Because of the concerns of climate change, it is important for coal-fired power plants to become more efficient in order to maximize the energy generated per unit of fossil carbon consumed. Coal-fired power plants often operate utilizing the concept of the Rankine cycle or a similar power cycle with steam as the working fluid to drive the turbo-generators that produce the electrical power. The average U.S. coal-fired power plant operates at about 33% efficiency [2] calculated as the electrical power produced divided by the heat produced by the combustion process.

The heat extracted in the steam condenser by a utility cooling water stream is one of the largest losses of heat in the form of unusable energy in the Rankine cycle. This is because the latent heat of the working fluid is not easily utilized. One way to utilize the latent heat is to add an additional Power Recovery Cycle (PRC) that employs a Secondary Working Fluid (SWF) having a lower boiling point than the Primary Steam Working Fluid (PWF). The SWF is frequently an organic fluid and the cycle is often referred to as an Organic Rankine Cycle (ORC) [3]. The SWF, which is more volatile than the PWF, would then be vaporized under pressure as it condenses the steam under low-pressure (vacuum) conditions. The vaporized SWF would be routed through a turbo-generator to produce additional electrical power. This has the potential to increase the electrical power produced for the same quantity of original fuel.

A few studies have been published to evaluate the thermodynamic potential of PRCs applied to the steam cycle of a power plant. Ziolokowski and coworkers compared the efficiency of a supercritical coal-fired 900 MW unit in which the PRC replaced the LP (low pressure) turbine [4] [5]. Six SWFs were evaluated—propane, isobutane, pentane, ethanol, RF245fa, and R236ea—with estimated energy efficiency increases of 10% - 18% compared to the baseline. However, economics were not considered in this study.

Other applications of a PRC to increase energy efficiency have also been previously proposed. These include using a PRC to recover additional usable energy from: 1) the boiler exhaust gas [6], 2) the gas turbine exhaust in a gas turbine or combined cycle power generation system [7], and 3) from lower temperature heat sources such as geothermal [8]. Similarly, Chacartegui *et al.* evaluated the use of six SWFs—R113, R245, isobutene, toluene, cyclohexane, and isopentane—to replace steam in the bottoming cycle in a combined cycle gas turbine power plant. Overall optimum performance with toluene resulted in a 4% increase in overall efficiency compared to steam [9].

A similar method to recover heat from the condenser is to use a Kalina cycle which utilizes a water and ammonia mixture. The use of a binary mixture of compounds lowers the boiling point compared to steam, allowing more usable energy to be extracted. The Kalina cycle typically has a higher power recovery efficiency than the addition of an ORC when the SWF vaporizes at temperatures above 200˚C [10] [11]. As a result, it may not be as effective at the lower temperatures required in a bottoming cycle to condense steam. The cost of additional equipment, process complexity because of the necessary separation steps, and the higher pressures used are also drawbacks of the Kalina cycle compared to an ORC [12]. The ammonia and water mixture can be quite corrosive, so more expensive materials would need to be used compared to a less corrosive organic SWF such as benzene [13] [14]. Like the ORC, the Kalina cycle has been largely applied to waste heat recovery of gas turbines or internal combustion engines on scales less than 10 MW [15].

Another method to recover heat from the condenser is the use of a trilateral flash cycle. The main differences between the trilateral flash cycle and the Rankine cycle are that the fluid is only heated to a saturated liquid and a two-phase expander is used in the trilateral flash cycle. This cycle has been shown to extract more power than the use of a simple ORC using the same SWF [16]. However, it has the following drawbacks: the heat exchangers for the process tend to be very large, more cooling water is required, and the two-phase expanders are less efficient than the dry turbines of an ORC [17]. To date, industrial application is still in the early stages in comparison to the Kalina and organic Rankine cycles which have been applied at commercial scales for many years [18].

In the work presented here, a comprehensive study was performed to examine the use of single fluid SWFs, both organic and inorganic, as a supplemental power recovery cycle to increase the electrical power efficiency of a large-scale coal-fired power plant. This study extends previous work by optimizing the design based on economic factors, allowing commercial feasibility to be assessed.

## 2. Materials and Methods

### 2.1. Working Fluids

The selection of the SWF is the principal factor in efficiently increasing the power output of the recovery cycle. Candidate SWFs will have a volatility that allows vaporization while under pressure at temperatures lower than the condensation temperature of low-pressure (vacuum conditions) steam. The ideal SWF will have a relatively high heat capacity in order to reduce the size of the condenser, minimize degradation and corrosion characteristics, and minimize supply cost. The SWFs examined in this study focused on those with acceptable thermodynamic properties. These are listed in **Table 2** in the results section.

### 2.2. Design

A baseline preliminary design through the process flow diagram level of detail was developed for a traditional lignite coal-fired power plant with a 500 MW drive train. The PRC was then created as an "add-on" installation to this power train. The baseline design was modeled after the Coal Creek power station located near Underwood, ND at a capacity of 605 MW per boiler and validated with proprietary facility data. The size was then adjusted down to the typical unit size of 500 MW to provide ease of comparison in other applications. The equipment size was estimated using preliminary sizing methods included in the

ChemCad™ software application supplemented by methods described in Ulrich and Vasudevan [19]. The utility and chemical requirements were estimated by mass and energy balances.

### 2.3. Economic Analysis

A broad capital cost estimate was developed using an American Association of Cost Engineers (AACE) Class 4 factored estimating method [20] at an accuracy of ±40%. Vendor cost estimates were obtained for the coal-fired boiler, the steam turbines, and boiler water feed pumps. The rest of the equipment costs were estimated using cost charts in Ulrich and Vasudevan [19].

Line item module costs were estimated per the following equation [19]:

$$C_{BM,i} = F_{BM,i} * C_{p,i} \tag{1}$$

where $F_{BM,i}$ is the factored estimate taken from the appropriate chart in Ulrich and Vasudevan using any anticipated pressure or materials of construction factors from the preliminary design for equipment item "*i*", and $C_{p,i}$ is the estimated equipment cost of "*i*" at the basis date.

The Total Bare Module Cost (TBMC) was calculated as the sum of all of the individual module costs:

$$\text{TBMC} = \sum C_{BM,i} . \tag{2}$$

Contingency and fees of 18% of the TBMC were added to obtain the total module cost (TMC):

$$\text{TMC} = \text{TBMC} * 1.18 . \tag{3}$$

To account for building costs, site preparation, and offsite storage areas, an auxiliary factor of 10% of the TMC was included to yield the Fixed Capital Investment (FCI):

$$\text{FCI} = \text{TMC} * 1.10 . \tag{4}$$

The working capital was estimated to be 5% of the FCI. The addition of the FCI, working capital, and cost of the initial charge for consumable chemicals ($C_{cc}$) give the Total Capital Investment (TCI):

$$\text{TCI} = 1.05 * \text{FCI} + C_{cc} . \tag{5}$$

### 2.4. Process Design Assumptions

1) Each preliminary design was a grassroots project with an operating lifespan of 20 years (a typical time period used for new project economic analysis).

2) Each process had a 95% operating factor [19] under the expectation that this facility would become more efficient than other power units and thus be used as a base loading unit as opposed to a swing load unit.

3) The overall energy conversion efficiency of the baseline power plant (without a PRC) was 31% (a typical efficiency for a lignite coal three stage facility validated with data from the Coal Creek power station).

4) Air was fed in at 20% excess of the stoichiometric mass ratio [21].

5) 90% of the ash produced was fly ash and 10% was bottom ash (a typical efficiency for a lignite coal three stage facility validated with data from the Coal Creek power station).

6) Oxygen present in coal was consumed in the oxidation reactions.

7) The coal combusted completely.

8) The ash in the coal was equal to the ash out of the process.

9) Coal was received already pulverized.

### 2.5. Equipment Design Assumptions

1) Pumps and compressors operate at an efficiency of 65% [19].

2) Turbines have a polytropic efficiency of 89% [19].

3) The height to diameter ratio was 4 in knockout drums [19].

4) The pressure drop across the heated tube banks in the boiler was 34 kPa (5 psi), and 21 kPa (3 psi) for liquids in heat exchange units.

5) The pressure drop across a heat exchanger where a gas is condensing at low pressures is 3.5 kPa (0.5 psi).

6) The pressure drop through a knockout drum was 21 kPa (3 psi).

7) The knockout drum was sized by reducing the inlet velocity by an order of magnitude.

8) The deaerator was adiabatic.

9) The deaerator vent rate was 0.1% of the feed water flowrate (a typical value validated with data from the Coal Creek power station).

10) The boiler heat transfer efficiency was 80% (a typical value validated with data from the Coal Creek power station).

11) Initial system volumes of working fluids were equal to the amount held in surge plus 15% for estimated piping volumes.

### 2.6. Utility Assumptions

1) Cooling water is 17˚C in the cold season (October-March) and 29˚C in the warm season (April-September), based on data from the Coal Creek power station.

2) Cooling water is pressurized in an auxiliary area and enters the system at 138 kPa (20 psia).

### 2.7. Economic Assumptions

1) The economic estimates were normalized to a basis date of June 2021 with a Chemical Engineering plant cost index (CEPCI) value of 702 [22]. The CEPCI value of 400, January 2004, was used as the index basis for all Ulrich and Vasudevan [19] cost chart values. A CEPCI value of 616, January 2019, was used as the index basis for all vendor-quoted costs.

2) Maintenance was estimated to be 4% of the FCI [19].

3) The FCI was depreciated using a 17-year Modified Accelerated Cost Recovery System (MACRS) schedule.

4) The U.S. federal tax rate on taxable income was 21% and state tax for North Dakota was 4.31% [23].

5) The hurdle rate for the project was 12%.

6) The loaded cost of a power plant operator in North Dakota was $93,510/yr [24].

7) For project implementation scheduling purposes, the project completion time was estimated based on a procurement time for the boiler of 12 months (vendor estimate). Both design and implementation were estimated to be 75% of the procurement time for the boiler which was assumed to have the longest procurement of all the equipment required.

8) The replacement of the boiler feed water was 5% per year, and the replacement of the SWF was 5% per year (typical values validated with data from the Coal Creek power station).

9) One board operator was required per shift with the number of outside operators based on the number of equipment systems employed. Operating labor supervision was estimated to be 15% of the total outside and board operator labor costs [19].

10) An auxiliary factor used for site preparation and supporting facilities was 10% [19].

11) Working capital was 5% of the FCI [19].

12) The flue gas desulfurization process capital and operating costs were obtained from a previous, unpublished techno-economic analysis using consistent estimating methods.

13) The average price of electricity is the average breakeven price of the baseline plant over a 20-year period. This value was calculated to be $0.0667/kWh at the basis date of June 2021.

14) A salvage value of $0.00 was used.

## 3. Results and Discussion

### 3.1. Process Design—Traditional Coal-Fired Power Plant

The base case has a total drive-train power production of 500 MW of electricity. The coal used in the combustion process was lignite coal from the Center, ND mine, which was analyzed at the University of North Dakota. The proximate and ultimate analyses of coal are shown in **Table 1**. Using a coal higher heating value of 16,140 kJ/kg (6939 Btu/lb) and assuming an overall base case plant efficiency of 31%, the coal required for 500 MW production is a throughput of 3.2 million metric tons/yr (3.5 million tons/yr).

An overview of the baseline process can be seen in **Figure 1**. The overall process was divided into two process areas. Process area 1 contains the boiler and auxiliaries while process area 2 contains the turbo-generators and associated equipment. More detail on the process can be found in the process flow diagrams provided in the Supplementary Materials.

In process area 1, air is preheated in a secondary tube bank in the economizer

**Table 1.** The proximate and ultimate analyses of lignite coal near center, ND (wt%).

| | |
|---|---|
| **Proximate Analysis** | |
| Moisture | 37.1% |
| Volatiles | 29% |
| Fixed Carbon | 28.9% |
| Ash | 5.1% |
| **Ultimate Analysis** | |
| Carbon | 40.9% |
| Hydrogen | 7.0% |
| Nitrogen | 0.5% |
| Sulfur | 0.7% |
| Oxygen | 45.8% |
| Ash | 5.1% |

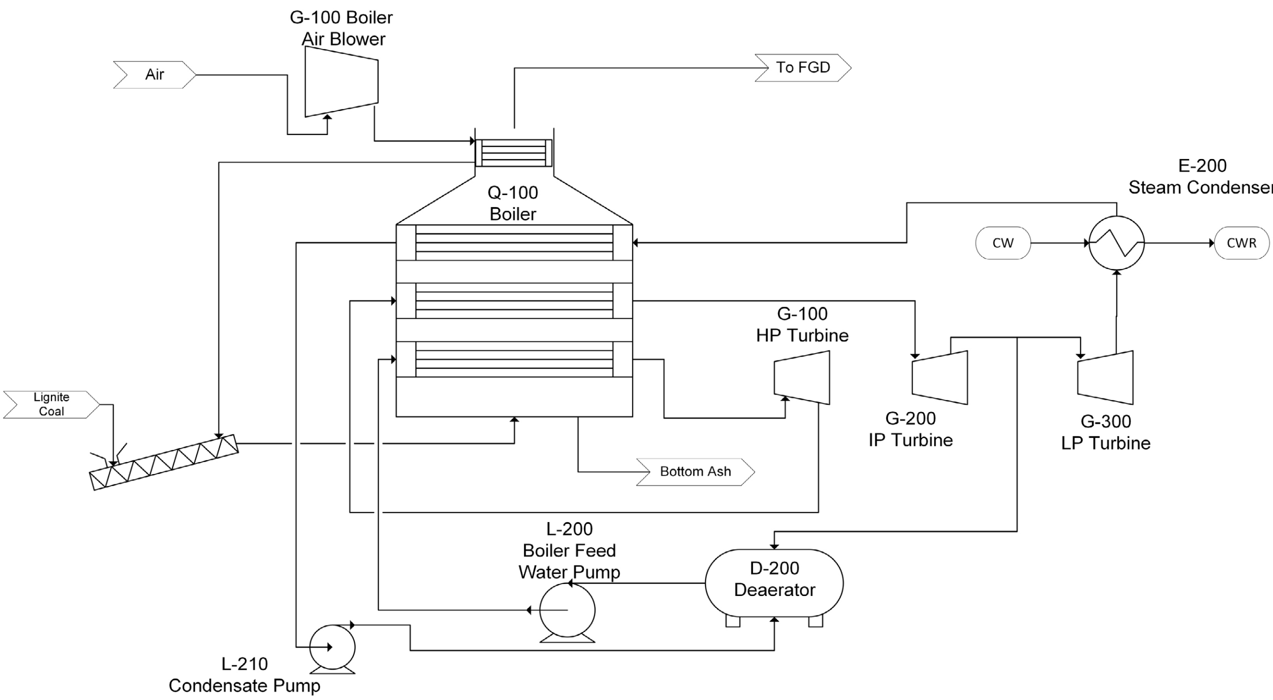


**Figure 1.** Process areas 1 and 2 for the baseline process.

section of the boiler. The pulverized coal feed is mixed with primary air and then combusted in the boiler. The boiler is used to vaporize high pressure boiler feed water to steam and superheat the feed going into the High-Pressure (HP) turbine in the primary tube bank, reheat the steam exhausted from the HP turbine going into the Intermediate-Pressure (IP) turbine in a secondary tube bank, and preheat the condensate going into the deaerator in an additional tube bank.

In process area 2, the condensate is pumped from the deaerator to the boiler to be vaporized and superheated before going into the HP turbine. After being exhausted from the HP turbine, the steam is reheated before entering the IP turbine. Some of the exhaust steam from the IP turbine is utilized in the deaerator to assist in the removal of oxygen and other dissolved gases from the condensate. The rest of the steam is used to drive the LP turbine. The steam is exhausted from the LP turbine at 7 kPa (1 psia) and condensed using cooling water in a bank of heat exchangers. The condensate is then fed back to the deaerator.

### 3.2. Process Design—Power Recovery Cycle Addition

The preliminary design was developed for the addition of a power recovery cycle (PRC). The cycle starts by replacing cooling water in the LP turbine steam exhaust condenser with a SWF. The PRC was simulated in the ChemCad version 7 process simulator for an array of candidate organic and inorganic fluids to estimate the energy produced in the recovery cycle turbine per unit mass of the different working fluids. This comparison was made to minimize the size of the equipment and therefore the capital cost of the project. Each candidate fluid was compared to benzene as shown in **Table 3**. The working fluids that had the best energy production in comparison to benzene per unit mass were methanol and hydrazine.

As shown in **Table 2**, hydrazine and methanol were the top two fluids. Ethanol would result in larger equipment than methanol, so it was removed from consideration [25]. Preliminary designs were then developed for three alternative SWFs: benzene, methanol, and hydrazine. Aside from the LP turbine and the use of a cross-exchanger instead of a cooling water condenser, process areas 1 and 2 are the same as the base case. The LP turbine exhaust pressure was increased to allow enough low-grade heat to vaporize the pressurized SWF in the cross exchanger. The power production of the process varies by the season because the exhaust pressure of the Power Recovery (PR) turbine was limited by the

**Table 2.** A comparison of power extracted from different working fluids in the PRC turbine. The power extracted is divided by the comparable power when using benzene to provide a clear comparison.

| Working Fluid | Warm Season | Cold Season | Seasonally Averaged |
|---|---|---|---|
| Benzene | 1.0 | 1.0 | 1.0 |
| Methanol | 2.0 | 2.1 | 2.0 |
| Hydrazine | 2.5 | 2.2 | 2.3 |
| Ethanol | 1.8 | 1.8 | 1.8 |
| Toluene | 1.0 | 0.9 | 0.9 |
| Hexane | 0.8 | 0.9 | 0.8 |
| Cyclohexane | 0.9 | 0.9 | 0.9 |
| Cyclopentadiene | 0.9 | 0.9 | 0.9 |
| Isopropyl Alcohol | 1.5 | 1.4 | 1.4 |
| 2-butanone | 1.0 | 1.0 | 1.0 |

available cooling water temperature during the summer season.

For each of the different working fluids, the exhaust pressure of the LP turbine was simulated at 103, 138, 172, and 207 kPa (15, 20, 25, and 30 psia). The overall power outputs for the LP and PR turbines were summed for the warm and cold seasons and averaged. The highest average power production at the respective outlet pressure for the LP turbine was selected as the operating condition of the design. For benzene and methanol, the highest season-averaged power productions for these two turbines were 318 MW and 289 MW, respectively at a LP turbine outlet pressure of 103 kPa (15 psia). For hydrazine, the highest power production was 307 MW at a LP turbine outlet pressure of 207 kPa (30 psia). A comparison to the baseline power production is shown in Table 3.

The PRC addition was added to the baseline design as a separate process area, Process area 3. A simple schematic of area 3 is shown in Figure 2. The process

Table 3. Overall power production for the evaluated process alternatives.

| Alternative | Overall Power Warm Season (MW) | Overall Power Cold Season (MW) | kWh/kg Coal Fed (Warm Season) | kWh/kg Coal Fed (Cold Season) | Increase in Power Cold Season[b] | Increase in Power Warm Season[b] |
|---|---|---|---|---|---|---|
| Baseline[a] | 500 | 500 | 1.38 | 1.38 | - | - |
| Benzene | 550 | 570 | 1.51 | 1.57 | 10% | 14% |
| Methanol | 530 | 540 | 1.46 | 1.49 | 6% | 8% |
| Hydrazine | 550 | 550 | 1.51 | 1.51 | 10% | 10% |

[a]500 MW lignite coal fired power train with no power recovery cycle, [b]Increase compared to the baseline case.

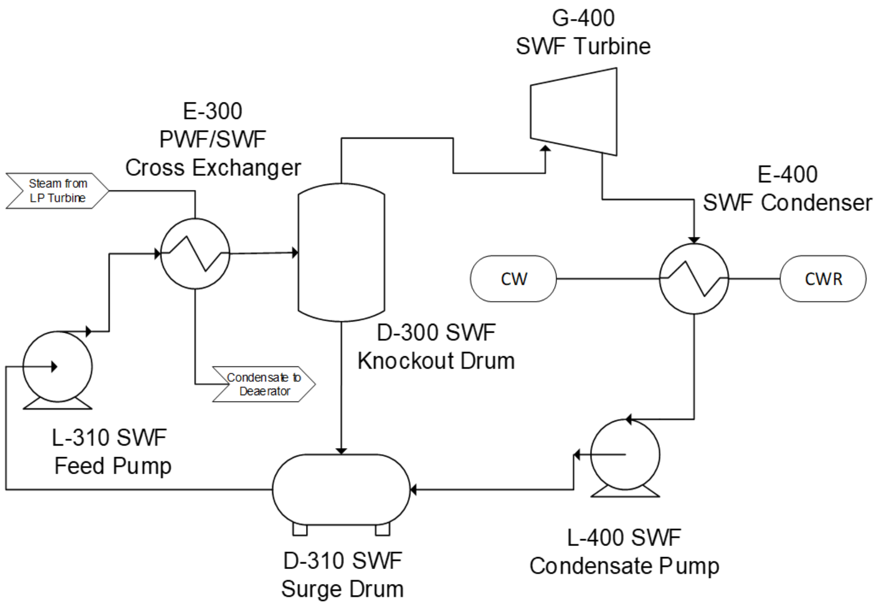


Figure 2. Process area 3 for the SWF process alternatives.

and type of equipment in area 3 was the same for the three alternative SWFs, but sizes and number of parallel equipment pieces varied. The SWF enters the cross-exchanger at 4˚C (7˚F) higher than the available cooling water used in the PRC condenser. The SWF exits the PWF/SWF cross-exchanger 4˚C (7˚F) lower than the inlet LP steam temperature. The vaporized SWF from the cross-exchanger is fed into a knockout drum to remove any remaining liquid and then routed into the PR turbine. The SWF vapor leaves the turbine at the minimum pressure that will allow the use of cooling water to condense the SWF in the downstream condenser. The outlet pressures for benzene, methanol, or hydrazine in the warm season were 22, 34, and 7 kPa (3.2, 5, and 1 psia), respectively. The outlet pressures in the cold season were 14, 18, and 7 kPa (2, 2.6, and 1 psia), in the same order. The SWF condensate is pumped to a surge drum and then pumped back into the cross-exchanger, completing the PRC.

### 3.3. Capital Cost Estimates

A broad estimate of capital costs was developed for all equipment in the process having a preliminary size equivalent to a pump or larger. A condensed summary of the estimated capital costs for the baseline case is provided in **Table 4** and **Table 5**, corresponding to the two process areas. These tables were condensed based on unit classification. The broadest of these classifications was "pressure vessels" which included surge drums, knockout drums, and the deaerator. The tables also include the additional investments required as a part of the total capital investment using a basis date of June 2021. **Table 6** shows the incremental additional capital costs required for each of the three SWFs (process area 3 plus adjustments in the LP turbine and heat exchanger costs). Detailed capital cost sheets can be found in the Supplementary Materials.

Vendor cost estimates were obtained for the boiler, boiler feed pumps, and the turbo-generators. The costs for the rest of the equipment were estimated using costing charts published by Ulrich and Vasudevan [19]. The basis date of

**Table 4.** The capital cost for the baseline process Area 1.

| Equipment Type | # of Units | Total Module Cost ($000) |
|---|---|---|
| Boiler | 1 | 17,000 |
| Compressor | 1 | - |
| **Total Bare Module Cost** | | **17,000** |
| Contingency and Fees | 0.18*TBMC | 3100 |
| Total Module Cost | | 20,000 |
| Auxiliary Facilities | 0.1*TMC | 2000 |
| Fixed Capital Investment | | 22,000 |
| Working Capital | 0.05*FCI | 1100 |
| **Total Capital Investment** | | **23,000** |

**Table 5.** The capital cost for the baseline process Area 2.

| Equipment Type | # of Units | Total Module Cost ($000) |
|---|---|---|
| Process Vessels | 3 | 790 |
| Heat Exchangers | 2 | 10,000 |
| Pumps | 20 | 12,000 |
| Turbines | 3 | 360,000 |
| Generators | 3 | - |
| **Total Bare Module Cost** | | **380,000** |
| Contingency and Fees | 0.18*TBMC | 68,000 |
| **Total Module Cost** | | **450,000** |
| Auxiliary Facilities | 0.1*TMC | 45,000 |
| Fixed Capital Investment | | 500,000 |
| Working Capital | 0.05*FCI | 25,000 |
| **Total Capital Investment** | | **530,000** |

equipment was updated from the basis dates in cost charts and from the time of the vendor quotes using the CECPI overall index (see assumptions for index values) to a basis date of June 2021.

The equipment required for the PRC is similar, except the size of the equipment and number knockout drums and PWF/SWF cross exchangers varied by SWF. The difference in the size and number of the cross exchangers was due to the differences in overall heat transfer coefficients of the different fluids when condensing steam. The overall heat transfer coefficients were estimated using Ulrich and Vasudevan [19]. Benzene had the lowest estimated overall heat transfer coefficient at 820 W/m$^2$-K while the overall heat transfer coefficients for hydrazine and methanol were higher at about 1300 W/m$^2$-K. The FCI of the processes were the sum of the TMC of equipment and the auxiliary factors. The FCI for the baseline, benzene, methanol, and hydrazine processes are $680, $890, $790, and $790 million ± 40%, respectively.

The other contributions to the capital costs were the working capital and the initial charge of chemicals. The working capital was estimated to be 5% of the FCI. The initial charge of boiler feed water was valued at $2.60/1000kg ($1.18/1000lbs) [26]. This was the only chemical required for the baseline process. The costs of benzene, methanol, and hydrazine used were $0.68/kg ($0.31/lb) [27], $0.48/kg ($0.22/lb) [28], and $9.09/kg ($4.13/lb) [29], respectively. The prices for benzene and methanol were the average trend prices projected from market data. The price of hydrazine was a spot price. The sum of the FCI, working capital, and initial chemical charges totals up to the TCI. The TCI for the baseline, benzene, methanol, and hydrazine processes are $720, $940, $830, and $850 ± 40%, respectively.

**Table 6.** The incremental capital cost for process area 3 of the PRC alternatives.

| Equipment Type | # of Units (Benzene/ Methanol/ Hydrazine) | Total Module Cost Benzene ($000) | Total Module Cost Methanol ($000) | Total Module Cost Hydrazine ($000) |
|---|---|---|---|---|
| Heat Exchangers | 12/8/8 | 110,000 | 58,000 | 58,000 |
| Pumps | 4/4/4 | 1600 | 1800 | 1500 |
| Process Vessels | 7/6/6 | 3400 | 2500 | 2500 |
| Turbines | 1/1/1 | 110,000 | 93,000 | 120,000 |
| Generators | 1/1/1 | - | - | - |
| LP Turbine Adjustment | - | (65,000) | (65,000) | (93,000) |
| Heat Exchanger Adjustment | - | (10,000) | (10,000) | (10,000) |
| Total Bare Module Cost | | 160,000 | 81,000 | 81,000 |
| Contingency and Fees | 0.18*TBMC | 29,000 | 15,000 | 15,000 |
| **Total Module Cost** | | **190,000** | **96,000** | **96,000** |
| Auxiliary Facilities | 0.1*TMC | 19,000 | 9600 | 9600 |
| Fixed Capital Investment | | 210,000 | 110,000 | 110,000 |
| Working Capital | 0.05*FCI | 10,000 | 5300 | 5300 |
| Initial Chemical Charge | | 4400 | 1100 | 18,000 |
| **Total Capital Investment** | | **220,000** | **110,000** | **130,000** |

### 3.4. Operating Cost Estimates

The annual operating costs of the process were found from the sums of the raw materials consumed, consumable chemicals replaced, operating labor, maintenance, utilities, and the annual operating cost of the flue gas treatment. The total estimated operating costs for the baseline, benzene, methanol, and hydrazine processes are $140, $141, $155, and $138 million ± 40%/yr, respectively. A summary breakdown of the annual operating costs for each option can be seen in Table 7 with a more detailed breakdown provided in the Supplementary Materials.

The only raw material consumed in this process is lignite coal. The price used

Table 7. The summarized annual operating costs for each process alternative ($000, June 2021).

| Process Alternative | Raw Materials | Consumable Chemicals | Operating Labor | Maintenance | Utilities | Flue Gas Treatment | Total |
|---|---|---|---|---|---|---|---|
| Baseline | 73,000 | 19 | 5800 | 27,000 | 13,000 | 21,000 | 140,000 |
| Benzene | 73,000 | 220 | 7100 | 35,000 | 3700 | 21,000 | 140,000 |
| Methanol | 73,000 | 57 | 6500 | 32,000 | 17,000 | 21,000 | 150,000 |
| Hydrazine | 73,000 | 910 | 6500 | 32,000 | 4700 | 21,000 | 140,000 |

in this study for lignite coal was $23.90/MT. Because each process has the same throughput of coal, the cost of coal annually is $73 million/yr for all the processes.

The consumable chemicals were boiler feed water and the working fluids for the SWF alternatives. With the assumed loss of 5% of the boiler feed water per year in all the processes, the cost of the makeup for all the processes is $54/yr. The loss of benzene, methanol, and hydrazine were assumed to be 5% per year. The respective costs for the chemicals are $220,000/yr, $57,000/yr, and $910,000/yr.

The operating labor is the sum of loaded operator salaries, supervision, overhead, and administration costs. The loaded salary for a North Dakota power plant operator of $93,510/yr was used [12]. With 14 operators operating the power plant, the annual operator cost is $1.3 million/yr. The cost of supervision was assumed to be 15% of the operator labor. The annual cost of supervision is $200,000/yr. The overhead was estimated to be 60% of the sum of operator labor, supervision, and maintenance per year. The estimated overhead is $17, $22, $20, and $20 million/yr for the baseline, benzene, methanol, and hydrazine processes, respectively. The administration costs annually are estimated to be 25% of the overhead costs. The administrative costs are estimated to be $4.3, $5.5, $5, and $5 million/yr for the baseline, benzene, methanol, and hydrazine processes, respectively. The total cost of operating labor was calculated by summing the operator cost, supervision costs, and administrative costs. The total labor costs for the baseline, benzene, methanol, and hydrazine processes are $5.8, $7.1, $6.5, and $6.5 million/yr, respectively.

The maintenance of the process was assumed to be 4% of the FCI for each of the processes. The annual cost for maintenance of the baseline, benzene, methanol, and hydrazine processes are $27, $35, $32, and $32 million, respectively.

The utilities are the costs for cooling water and electricity. The cost of the cooling water was $14.80/1000kg. The estimated cost for cooling water for the traditional, benzene, methanol, and hydrazine processes each year are $13, $3.7, $17, and $4.7 million, respectively. For the electricity requirements, the traditional process requires 19 MW, and the PRC processes require 20 MW. The electricity is assumed to come from the process, so the values are just subtracted from the revenue.

### 3.5. Revenue Estimate

The source of revenue from the power plant is from the electricity produced. The price of electricity was back calculated to force the traditional power plant to break even at the end of a 20-year period. The calculated price was \$0.0667/kWh. The season-average saleable capacities of the traditional, benzene, methanol, and hydrazine plants are 481 MW, 540 MW, 515 MW, and 530 MW, respectively. At those capacities, the plants revenues were estimated at \$270, \$300, \$290, and \$290 million/yr, respectively.

### 3.6. Profitability

The lifetime operation of the four processes was assumed to be a 20-year period, although power plants can have a lifetime of over 40 years. The hurdle rate was 12%. The baseline process was normalized to break-even in the 20-year period, so it had a net present value (NPV)@12% of \$0 at a June 2021 basis date. The benzene process had an NPV@12% of −\$32 million ± 40%. The methanol process had an NPV@12% of −\$59 million ± 40%. The hydrazine process had an NPV@12% of +\$35 million ± 40%. Cash flowsheets used to calculate each of these values are provided in the Supplementary Materials.

### 3.7. Sensitivity Analysis

The factors that most affect the profitability of adding a PRC are the revenue price of electricity, the fixed capital investment, SWF cost, and the cost of the coal feedstock. Each of these factors was examined in more detail.

An important factor that varies the economics of the processes is the price at which electricity is sold. The basis electricity price used was \$0.0667/kWh which was the breakeven price for the baseline process. As a regulated utility in the U.S., electricity prices have very little historical price variability and are not set by market factors, but by agreement between the regulatory authority and the power company. To examine the impact of electricity prices on project profitability, the price was varied by ±10%. **Figure 3** shows the NPV of each process vs. the price of electricity. The breakeven prices for the benzene, methanol, and hydrazine processes all fall within this variability range at \$0.068/kWh, \$0.0692/kWh, and \$0.0653/kWh, respectively. The differences between these prices and the baseline value are \$0.0013/kWh, 1.9% for benzene, \$0.0025/kWh, 3.6% for methanol, and −\$0.0014/kWh, −2.1% for hydrazine. A project using hydrazine as the SWF is viable even with electric price decreases of 2.1%. For the SWF that has the lowest potential health and environmental impact, methanol, the project is viable at electric price increases of 3.6% with the intermediate SWF choice benzene viable at increases of 1.9%.

A second influential factor that varies the economics between the three SWF PRC processes is the FCI for process area 3. The basis FCIs for process area 3 for the benzene, methanol, and hydrazine processes are \$210, \$110, and \$110 million, respectively. **Figure 4** shows the NPV@12% vs. the FCI of process area 3 for

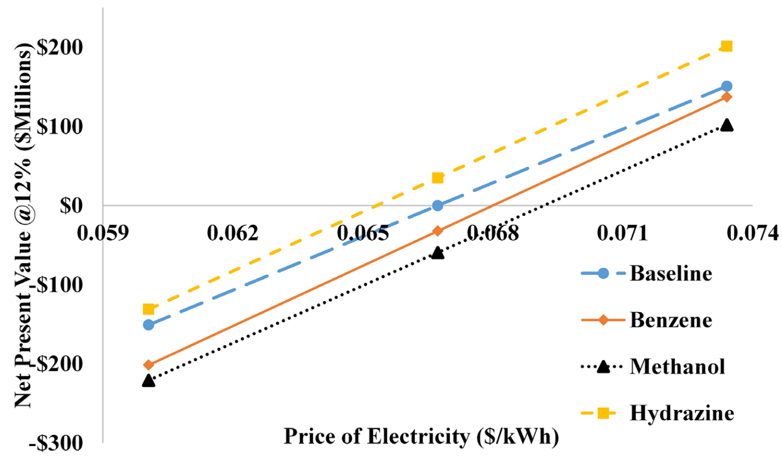


**Figure 3.** Sensitivity of NPV@12% to the price of electricity for the baseline, benzene, methanol, and hydrazine processes. The center points are the basis values. The upper and lower values of electricity were varied by ±10%.

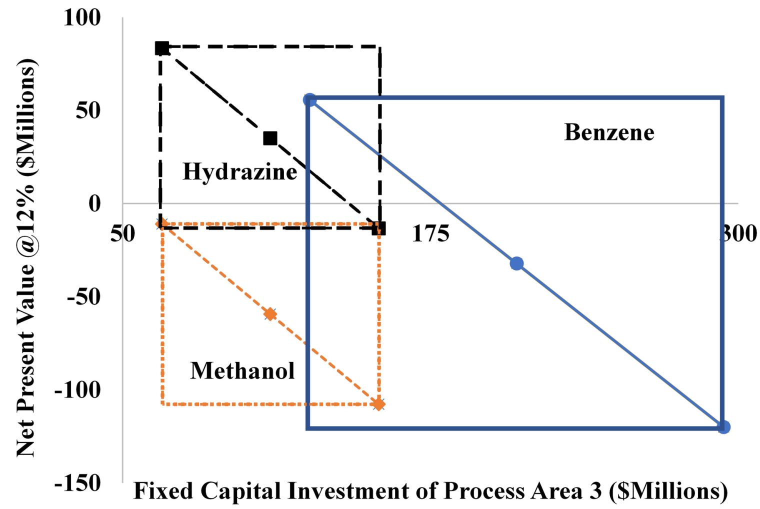


**Figure 4.** Sensitivity of NPV@12% to the FCI for process area 3 for the benzene, methanol, and hydrazine processes. The center points are the basis values. The boxes are the regions of most probable uncertainty within ±40% of the basis FCI.

the PRC alternatives. The breakeven FCI's are: $179, $56, and $142 million for benzene, methanol, and hydrazine, respectively. The benzene process is profitable at a process area 3 FCI of less than $179 million which represents 32% of the region of most probable uncertainty. The methanol process is not profitable within the region of most probable uncertainty. The hydrazine process is profitable when the FCI of process area 3 is less than $142 million which represents 86% of the region of most probable uncertainty. While a hydrazine-based project is estimated to be worthwhile at the basis estimate value, benzene- and methanol-based projects require a decrease in the FCI of 15% and 49%, respectively to be worthwhile. These results highlight the need for further, more detailed study of the actual costs for adding a PRC to an existing coal-fired power plant in

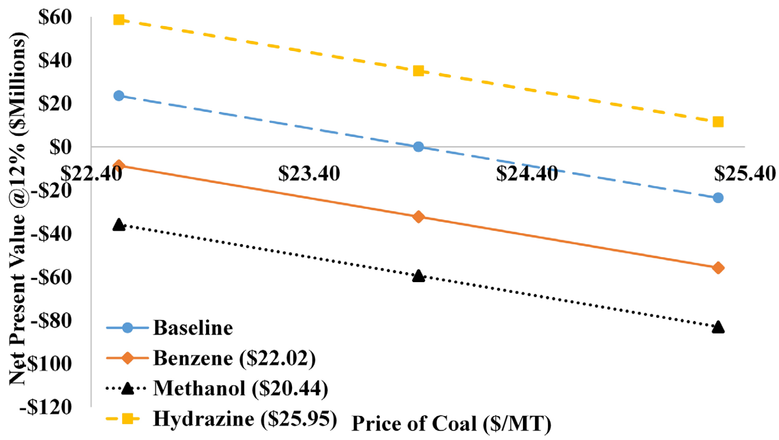


**Figure 5.** Sensitivity of NPV@12% to the price of coal for the baseline, benzene PRC, methanol PRC, and hydrazine PRC processes. The center points are the basis values and the breakeven values are shown in parentheses in the legend.

order to reduce the uncertainty in these results.

The effects of SWF price and lignite coal price on economics were also examined for the processes with the additional PRC. The prices of benzene and methanol were varied at their highest and lowest prices based on their trend prices. The price of hydrazine was varied by ±40% of the spot price used as its basis price. The price of the SWFs does not appear to have a significant effect on the profitability of any of the processes.

The price of lignite coal was varied using its four-year historical prices from 2017 to 2020 [30]. For the benzene, methanol, and hydrazine processes to break even, the price of coal must be $22.02, $20.44, and $25.94/MT, respectively. **Figure 5** shows the NPV vs. coal price for all the processes. The benzene and methanol processes are not profitable in their regions of most probable uncertainty, so they are very likely to not be profitable at recent coal prices in comparison to the baseline process without an increase in electricity price. The hydrazine process region of most probable uncertainty is all profitable, so it is likely that it will be more profitable than the baseline process based on recent coal prices.

## 4. Conclusions

The objectives of this study were to determine the commercial feasibility of adding a power recovery cycle to recover additional energy from the steam cycle and thus improve the efficiency of a lignite coal-fired power plant and identify the most viable secondary working fluid candidates for such a system. The base case used was a lignite coal-fired power plant with a 500 MW capacity located in central North Dakota, USA. A preliminary screening study identified benzene, methanol, and hydrazine as the most promising candidate secondary working fluids for the PRC. At the same lignite coal feed rate of 3.2 million metric tons/yr, the addition of a PRC with benzene, methanol, or hydrazine SWFs increased the season-average power generation by 60, 35, and 50 MW, respectively.

When electricity was sold at the breakeven rate for the traditional power plant, which is close to the current wholesale price for power in the U.S. Upper Midwest, the addition of a PRC is projected to be profitable over a 20-year period with a hydrazine SWF but slightly unprofitable for the benzene and methanol process alternatives. The NPV@12% for the benzene, methanol, and hydrazine alternatives were −$32, −$59, and +$35 million ± 40%, respectively. Breakeven electrical prices for the three cases were $0.068/kWh, $0.0692/kWh, and $0.0653/kWh, for the three SWFs, respectively which are 1.9%, 3.6%, and −2.1% different than the baseline price. Therefore, a hydrazine PRC is likely to be economically justified at current prices while only a slight increase in electricity prices would provide the economic incentive for benzene or methanol PRCs.

This study also highlighted that the additional capital costs of the equipment required for a PRC project is an important factor affecting profitability. Companies interested in evaluating this technology may need to generate a more accurate design and cost estimate in order to determine if the PRC is worthwhile and whether hydrazine or benzene should be chosen as the SWF. A hydrazine-based PRC appears to be the more profitable, but also represents a greater risk of health and/or environmental impact. A methanol-based system has the lowest health/environmental risks and is the easiest SWF to implement but has the least attractive economics. A benzene-based system may be the best compromise of impact on electricity price, efficiency gain, and health/environmental safety risks.

It can be concluded from this study that the addition of a PRC to existing coal-fired power plants is worthy of consideration and can likely be adopted with only a modest impact on electricity prices. While this general conclusion should be widely applicable, it must be noted that the presented work is based on the conditions and circumstances present in the central North Dakota region of the U.S. A comparable specific feasibility study should be conducted for any location considering the addition of a PRC to increase the overall energy efficiency of their facility. If feasible, a more detailed level of design and economic analysis is recommended to verify that the addition of a PRC is worthwhile. Further, decisions on whether to utilize hydrazine or benzene as the SWF should also be made on a case-by-case basis that considers economic, corrosivity, and health/safety factors.

## Conflicts of Interest

The authors declare no conflicts of interest regarding the publication of this paper.

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