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# Use of a Nuclear High Temperature Gas Reactor in a Coal-to-Liquids Process

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# SUMMARY

AREVA's High Temperature Gas Reactor (HTGR) can potentially provide nuclear-generated, high-level heat to chemical process applications. The use of nuclear heat to help convert coal to liquid fuels is particularly attractive because of concerns about the future availability of petroleum for vehicle fuels. This report was commissioned to review the technical and economic aspects of how well this integration might actually work. The objectives were to review coal liquefaction processes and propose one or more ways that nuclear process heat could be used to improve the overall process economics and performance.

Shell's SCGP process[a] was selected as the gasifier for the base case system. It operates in the range of 1250 to 1600°C to minimize the formation of tars, oil, and methane while maximizing the conversion of the coal's carbon to gas. Synthesis gas from this system is cooled, cleaned, reacted to produce the proper ratio of hydrogen to carbon monoxide, and fed to a Fischer-Tropsch (FT) reaction and product upgrading system. The design coal-feed rate of 18,800 ton/day produces 26.000 barrels/day of FT products.

Thermal energy at approximately 850ºC from a HTGR does not directly integrate into this gasification process efficiently. However, it can be used to electrolyze water to make hydrogen and oxygen, both of which can be beneficially used in the gasification/FT process. These additions then allow carbon-containing streams of carbon dioxide and FT tail-gas to be recycled in the gasifier, greatly improving the overall carbon recovery and thereby producing more FT fuel for the same coal input. High temperature heat from the HTGR the replaces the heat formerly generated by burning the tail-gas. The final process configuration, scaled to make the same amount of product as the base case, requires only 5,800 ton/day of coal feed. Because it has a carbon utilization of 96.9%, the process produces almost no carbon dioxide byproduct

The nuclear-assisted process requires six AREVA reactors to supply the heat, so the capital cost is high. The conventional plant is projected to cost $1.7 billion while the nuclear version would cost $5.2 billion. The products from each have a projected cost of $1.87/gal and $2.87/gal respectively. While the nuclear version is more expensive, a sensitivity analysis shows that it is minimally affected by future changes in the cost of coal or possible future penalties for the production and emission of carbon dioxide. However, the conventional process can see a five-fold variation in its projected cost of fuel depending on a plausible range of these costs. At the high end of the coal and carbon dioxide cost ranges, the nuclear-assisted coal-to-liquid process produces synthetic fuel at lower cost than the conventional process.

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# ACRONYMS AND ABBREVIATIONS

| | |
|---|---|
| ASPEN | Brand name of process modeling software |
| BTU | British thermal unit |
| ºC | Celsius |
| $CH_4$ | Methane |
| CO | Carbon monoxide |
| $CO_2$ | Carbon dioxide |
| CTL | Coal to liquids |
| CVX | Brand name of a coal gasification process |
| DOE | U.S. Department of Energy |
| E-GAS | Brand name of a coal gasification process |
| FT | Fischer-Tropsch |
| gal | Gallon |
| GSP | Brand name of a coal gasification process |
| $H_2$ | Hydrogen |
| $H_2S$ | Hydrogen sulfide |
| He | Helium |
| hr | Hour |
| HTE | High-temperature electrolysis |
| HTR | High-temperature reactor |
| HTGR | High temperature gas reactor (nuclear) |
| INL | Idaho National Laboratory |
| Kg | Kilogram |
| kWe | Kilowatts electric |
| kWhr | Kilowatt-hour |
| lb | Pound |
| MW | Megawatt |
| $MW_{th}$ | Megawatt thermal |
| $N_2$ | Nitrogen |
| NETL | National Energy Technology Laboratory |
| $O_2$ | Oxygen |
| R&D | Research and development |
| ROI | Return on investment |
| SCGP | Brand name of a coal gasification process |
| Std dev | Standard deviation |

# PRODUCT DISCLAIMER

# Use of a Nuclear High Temperature Gas Reactor in a Coal-To-Liquids Process

## 1. BACKGROUND

### 1.1 Project Objectives

AREVA's High Temperature Gas Reactor (HTGR) can potentially provide nuclear-generated, high-level heat to chemical process applications. The use of nuclear heat to help convert coal to liquid fuels is a particularly attractive concept because of concerns about the future availability of petroleum for vehicle fuels. If it is feasible, the large scale at which such fuel processes would be practiced makes converting coal to liquid fuels an attractive market for HTGRs when compared to other more specialized and therefore smaller or less numerous applications.

This report was commissioned to review the technical and economic aspects of how well this integration might actually work. The objectives were to review coal liquefaction processes and to propose one or more ways that nuclear process heat could be used to improve the overall process economics and performance of coal liquefaction. Though direct economic improvement would be most desirable, increasing U.S. energy security by using domestic coal and uranium to produce vehicle fuels would also be desirable because of the likelihood of future governmental incentives for such processes.

As part of this review, coal conversion technology was reviewed to select the approach that seemed most practicable. ASPEN modeling software was used to generate mass and energy balances to predict overall process performance. With that baseline flowsheet, opportunities to use high-temperature nuclear heat in the process were explored. The best applications would match the temperatures of the HTGR (see Figure 1), about 800-850ºC or about 500ºC and would have a duty matching the reactor's 600 MW

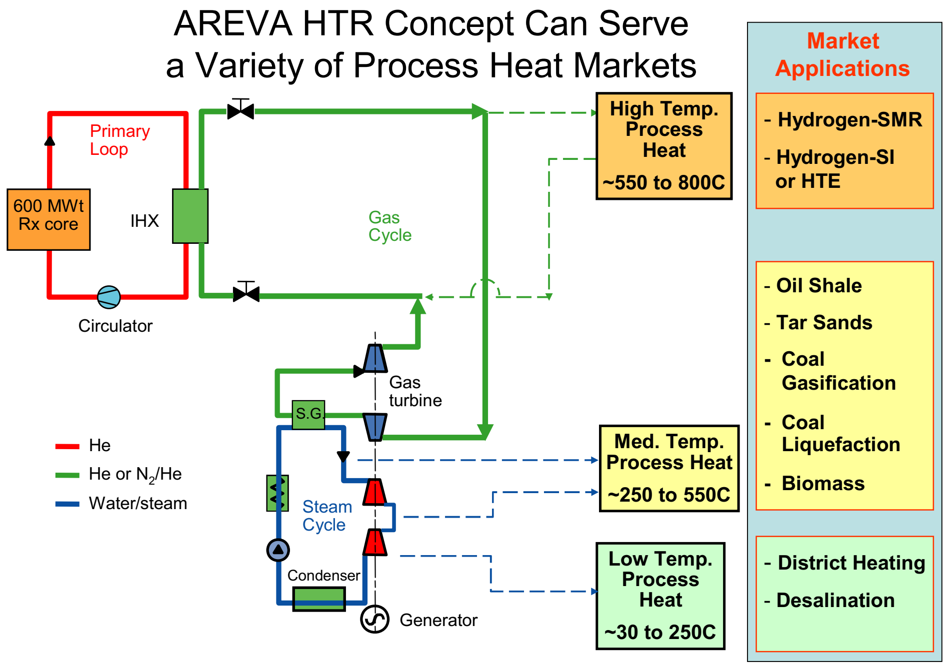

Figure 1. Process heat integration with the AREVA High Temperature Gas Reactor .

thermal output. The use of hydrogen generated with nuclear heat was also considered because raising the hydrogen content of coal is necessary to make liquid fuels.

As various process changes were evaluated, the operating improvements were catalogued and, at the final configuration, the capital cost was estimated. This design, though an improvement over conventional nonnuclear coal liquefaction processes, was not optimized. The projected cost of liquid fuels from this combined nuclear/coal process could then be estimated to project the overall competitiveness of such a process.

## 1.2 Previous Research and Development on Nuclear-Assisted Coal Conversion

In the 1970s and 1980s coal conversion to liquid or gaseous fuels received a great deal of research attention. The preferred targets were synthetic petroleum, which was to be refined for vehicle fuel, or synthetic natural gas to be fed into the existing distribution infrastructure. A broad variety of processes in three general classes were developed, typically to the level of pilot plant or demonstration processes. The first class, gasification processes, relies on heat, steam, and oxygen to produce either methane or synthesis gas, a mixture of hydrogen, carbon monoxide, and other compounds. The other two classes were liquefaction processes. These could be either direct, producing liquids by direct reaction of coal with hydrogen-containing liquid species, or indirect, first gasifying the coal to synthesis gas then purifying and converting the synthesis gas into any of broad variety of fuel liquids such as methanol or Fischer-Tropsch (FT) products [DTI 1999]. Direct processes received the most attention. However, the economic incentives to continue development of synthetic fuel processes disappeared in the 1980s because of low petroleum prices.

During the period of active work, the possibility of using nuclear energy to augment the coal conversion was also explored with particular emphasis on making synthetic natural gas. The objective was improved economics by using nuclear energy to provide some of the energy for the endothermic steam-reforming reaction at the heart of the gasification process. At the time, issues of global warming from carbon dioxide emissions or of security of energy supply were not of primary importance in evaluating processes, though they were recognized [Engelman et al.1984]. Today, these issues provide important additional reasons for considering nuclear-assisted coal conversion processes.

While there was work in the U.S. [Kosky and Flock 1981, Spiewak et al. 1976, Jones et al. 1977] and France [Rastion et al. 1979], by far the greatest effort to use nuclear energy was in the Federal Republic of Germany [Baust et al.1984, Jager et al. 1984]. That work, which progressed to the level of large pilot plant testing, was aimed at making synthetic natural gas by hydrogasifying coal at the relatively low temperature of about 850ºC to encourage methane formation; the concomitant large amount of unconverted carbonaceous char was then steam gasified [Neef and Weisbrodt 1979, Kugeler 1980, Singh et al. 1984, Kirchhoff et al. 1984, Kubiak et al. 1983]. No work was apparently done on nuclear-assisted coal liquefaction, despite previous German experience with the Lurgi process for indirect coal liquefaction. Despite the breadth and progress of the program, it was shut down in the 1980s. The high-temperature reactor technology underpinning this effort was eventually sold to South Africa when German policy began to move against nuclear reactors in their country.

# 2. CONVENTIONAL COAL-TO-LIQUIDS PROCESS

## 2.1 Selection of Gasifier Type

Nearly all previous research and development (R&D) on nuclear-assisted coal conversion has attempted to match the gasification temperature with the highest temperature that can be made available from a nuclear reactor loop. While this is a logical approach, it results in a low gasification temperature, 950°C or lower. At this low temperature, the reasonable gasification options are limited. They are summarized below.

### 2.1.1 Steam Gasification

Steam gasification of coal is possible at temperatures above 800°C. The primary reaction is as follows:

$$C + H_2O \rightarrow CO + H_2$$

This reaction is endothermic, meaning that a source of high-temperature heat must be provided. The rate of this reaction is highly temperature-dependent. To obtain an acceptable rate, it is typically necessary to operate at 900 to 950 °C with a large excess of steam. Side reactions produce appreciable amounts of carbon dioxide and methane, and lesser amounts of tars, oils, and aromatic compounds. Extensive synthesis gas cleanup is required and the tars and oils can lead to problematic fouling of equipment. Even at a temperature of 950°C the gasification rate is slow so either a long residence time in the gasifier is required or the unreacted carbon must be recycled to the gasifier to improve conversion. There are few other uses for the resulting ash/char mixture, and its land disposal can be problematic due to the leachability of hazardous ash components.

### 2.1.2 Hydrogasification

Coal can be gasified in the presence of hydrogen at 900 to 950°C via the following reaction:

$$C + 2\ H_2 \rightarrow CH_4$$

Because this reaction is exothermic, it is possible to maintain the requisite reaction temperature without an external heat source. However, a source of hydrogen is needed. This is often generated by reforming natural gas, which is predominately methane ($CH_4$), with steam at 750 to 800°C:

$$CH_4 + 2\ H_2O \rightarrow CO_2 + 4\ H_2$$

Normally the endothermic heat of this reaction is supplied by burning part of the natural gas supply, but nuclear heat could be used instead.

Hydrogasification has low carbon conversion, typically only 60 to 70% of the carbon in the coal [Rastoin et al., 1979]. Hence, a suitable disposal path for the char-laden ash must be identified. The char can be steam gasified, but inclusion of a second gasifier design significantly complicates the overall process configuration. It should also be noted that the primary product from hydrogasification is methane, which is not readily converted into conventional liquid vehicle fuel although it can be used in engines designed for liquefied natural gas or compressed natural gas.

### 2.1.3 Direct Coal Liquefaction

During the 1980s, several processes were developed and demonstrated to directly convert coal to liquids at low temperatures (~450°C). The resulting liquids resemble a heavy, sour crude oil. Carbon conversions between 90 and 95% were demonstrated under optimal conditions. However, there are significant drawbacks to direct liquefaction for diesel fuel production. The high concentration of aromatic molecules in the product make it toxic and carcinogenic, and significant refining of the resulting crude liquid is required as aromatics significantly reduce the cetane number for a diesel fuel (although they raise gasoline's octane value). Direct liquefaction products are also prone to solidify at ambient temperatures.

### 2.1.4 Entrained Flow Gasification

In recent years, coal gasification R&D has focused on higher operating temperatures to address the issues associated with low- and moderate-temperature gasification. As a result, most new commercial coal gasifiers are entrained flow designs with operating temperatures typically in the range of 1250 to 1600°C. There are several benefits of operating at higher temperatures:

- Ash is heated above its melting point, resulting in the formation of a vitrified slag which meets environmental regulations for land disposal or can be utilized as feedstock for asphalt production

- Production of tars and oils is minimized

- Methane formation is nearly eliminated

- The resulting synthesis gas is comprised primarily of carbon monoxide and hydrogen, the building blocks for synthesis of chemicals and fuels

- The higher operating temperatures allow much shorter gasifier residence times and, therefore, smaller or fewer gasifiers

- Carbon conversion of greater than 99% is easily achievable.

Entrained flow gasifiers vary somewhat in design but can be broadly classified as either dry-fed or slurry-fed gasifiers. Shell's SCGP and Future Energy's GSP are examples of commercially available dry-fed gasifiers, while GE's CVX and Conoco-Phillips E-GAS are examples of commercially available slurry-fed gasifiers. While there are differences between these gasifiers, each is proven and is well-suited to produce a synthesis gas for synthetic fuel production. Shell's SCGP process was selected as the reference gasifier design for this work due to its very low production of methane and carbon dioxide.

## 2.2 Flowsheet Description

To evaluate the potential benefits of integrating a high-temperature nuclear reactor with coal gasification to produce synthetic fuels, a reference coal-to-liquids flowsheet was defined and simulated using ASPEN process modeling software. The reference flowsheet is shown in Figure 2, with a larger drawing and more description in Appendix A.

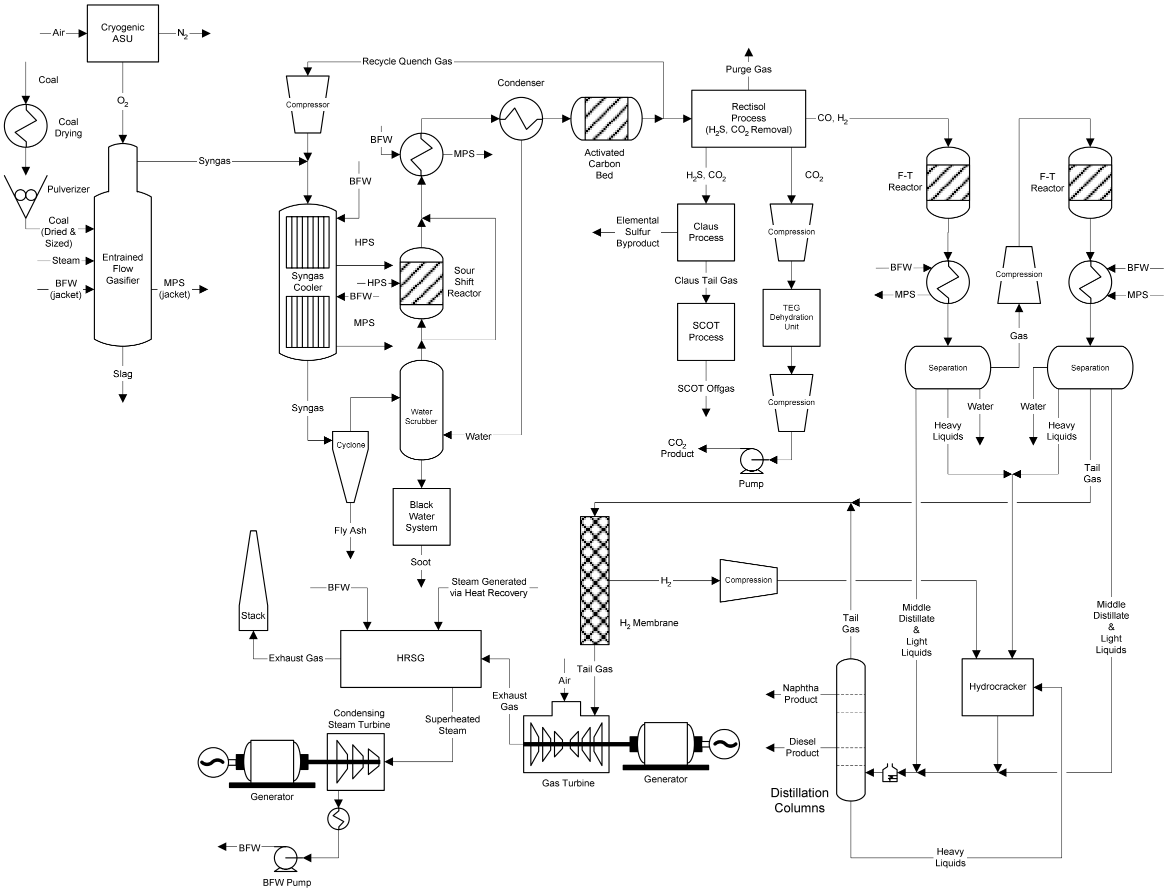

Figure 2. Reference coal-to-liquid process flowsheet.

Shell's SCGP process uses an entrained-flow, dry-fed, slagging, oxygen-blown, up-flow gasifier. The reactor wall is a membrane that is studded and covered with a castable refractory. The gasifier includes diametrically opposed burners (a term based on analogy to boilers) in the side-wall at the bottom of the reactor. Dried coal, ground to below 90 µm particle size to improve flowability, is fed to the burners in an inert gas, typically nitrogen or carbon dioxide. In the burners the coal is mixed with preheated oxygen and steam, and then injected into the gasifier. Slag leaves the gasifier through an opening in the bottom, where it is quenched with water and solidified. Synthesis gas leaves the gasifier through the top and is quenched with cool, recycled synthesis gas to lower the exit temperature to 900°C. Cooling the synthesis gas at the exit of the gasifier minimizes downstream plugging because any entrained slag is cooled below the temperature where it becomes sticky. The synthesis gas is further cooled to about 280°C in a synthesis gas cooler where the heat is recovered to generate high- and medium-pressure steam. After this cooling stage, solids are removed in a cyclone and a candle filter.

In the next step, wet scrubbing removes halide gases (hydrogen chloride and hydrogen fluoride) and ammonia. The composition of the synthesis gas is then adjusted using water-gas shift conversion to achieve the hydrogen/carbon monoxide ratio of approximately 2.15 which is optimal for synthetic fuel production in Fischer-Tropsch reactors using cobalt catalysts. The shift converters produce hydrogen via the water-gas shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2$$

The water-gas shift catalyst must be sulfur tolerant because the synthesis gas at this point in the process still contains as significant amount of hydrogen sulfide. The synthesis gas is cooled (with some heat recovery) and split into two streams. One is compressed and returned to the exit of the gasifier for use as quench gas, while the other stream constitutes the net synthesis gas production and is further treated in beds of granular activated carbon to remove mercury.

The net synthesis gas stream is cooled to approximately 60°C below zero and passed through a series of absorbers to remove sulfur compounds and carbon dioxide. This is the Rectisol process, which uses cold methanol as a physical solvent to reduce the sulfur species to very low levels. The resulting purified synthesis gas contains primarily hydrogen and carbon monoxide in a 2.15:1 ratio, and a few percent of inert gases such as carbon dioxide, nitrogen, and argon.

The conditioned synthesis gas is catalytically converted to paraffins and olefins in the Fischer-Tropsch reactor via an exothermic polymerization-like reaction:

$$CO + 2\ H_2 \rightarrow (\text{-}CH_2\text{-}) + H_2O$$

This reaction produces a distribution of hydrocarbons varying in carbon number from 1 to greater than 300. In addition, small amounts of oxygenates such as alcohols and acids are produced in the reactor. The product distribution from the reactor is sensitive to temperature; hence, careful reactor design and control are required to maximize selectivity to the desired product. In general, multiple reactors (typically two or three) are used in series with removal of liquid products and water between the reaction stages.

The mixture of hydrocarbon products is separated to generate the primary products from the Fischer-Tropsch plant: a heavy hydrocarbon liquid (wax) stream, a middle-distillate liquid stream, a crude naphtha liquid stream, a tail-gas stream containing light hydrocarbon gases and unreacted carbon monoxide and hydrogen, and water. The wax stream is hydrocracked to produce lighter hydrocarbon molecules which are mixed with the middle-distillate and crude naphtha Fischer-Tropsch streams and then distilled to produce diesel fuel and naphtha products.

The tail-gas stream contains unreacted hydrogen which can be recovered using either membrane separation or pressure swing adsorption. Recovered hydrogen is recompressed and used in the hydrocracking step. The remaining tail-gas stream has significant heating value due to the presence of carbon monoxide, hydrogen, methane, and other light hydrocarbons. This tail-gas can be burned in a combined cycle to produce the electricity needed by the processing plant and some excess for sale on the electrical grid.

In this baseline conventional process, a coal input of 18,800 tons/day results in the production of 26,000 barrels/day of Fischer-Tropsch liquids. Of the carbon in the coal fed to the process, only 29.5% appears in the synthetic fuel product. The balance is discharged as carbon dioxide or, in small part, as carbon in the gasifier ash.

# 3. ADAPTATION OF CONVENTIONAL PROCESS

The strategy for using nuclear heat in the CTL process was to identify conceptual approaches to using that energy, then to identify specific points in the process where those approaches could be applied. The baseline ASPEN simulation was then modified to reflect that change, and the consequences for the overall process were determined.

## 3.1 Direct Heat Integration

The obvious place to use high-temperature nuclear heat is in the coal gasifier which is both the hottest part of the process and endothermic, meaning the reactions there absorb heat and convert it to chemical energy in the product molecules. However, the 1200ºC operating temperature of the chosen gasification system is substantially above the nominal 850ºC available from the HTGR. The use of a heat-pump system to boost the temperature of that heat was rejected as being too difficult at these temperatures.

To maximize the overall thermodynamic efficiency of an energy-using process, there should be a close match between the operating temperatures of heat sources, such as the nuclear reactor or power cycle, and heat users, such as chemical reactors. In the CTL process, steps other than the gasifier operate at temperatures much below 800ºC or even 500ºC, the temperature at which the nuclear facility rejects heat from its Brayton cycle. Even if there were users in this temperature range, the CTL process itself would compete with the nuclear source because the 1200ºC gasifier product gas must be cooled before it is further processed. As shown in the baseline flowsheet (Figure 2), this energy is used to generate high-pressure steam for power production because there are no valuable uses for this energy in the chemical processing steps.

As with most chemical processes, at temperatures below about 250ºC there are a number of sources and sinks of heat, with the sources becoming especially numerous as the temperature drops. With an excess of lower grade heat, there is no incentive to thermally couple a nuclear reactor to the cooler parts of the flowsheet. Even if there were, a more conventional light-water reactor would be able to supply this heat instead.

## 3.2 Hydrogen Generation for Coal to Liquids Use

### 3.2.1 Process Improvements with an External Source of Hydrogen

The lack of opportunities for direct heat integration required the development of significantly different strategies for using the nuclear energy. Rather than using the nuclear system to provide thermal energy, it was used to supply or displace chemical energy. The first way this was attempted was based on the recognition that a large fraction of the carbon fed to the process is used to produce hydrogen rather than to go into synthetic fuel. An alternative external source of hydrogen offers several improvements in the CTL process:

- The water-gas shift reactors can be eliminated
- The carbon monoxide that had been converted to carbon dioxide in those reactors is available to make more Fischer-Tropsch feedstock (or alternatively to reduce the amount of coal needed for the same production of synthetic fuel)
- The size of the carbon dioxide removal equipment after the shift reactors is reduced.

If the hydrogen is made by a water-splitting technology such as electrolysis, then pure oxygen is also available and can be used productively. In many gasification processes, some amount of oxygen is fed into the gasifier where exothermic oxidation reactions both heat the feed streams to the reaction temperature and supply heat for the endothermic gasification reaction. Pure oxygen is used rather than air to reduce the amount of feed gas to be preheated and to prevent diluting the gasifier product stream with nitrogen. However, the air separation plant to supply this oxygen is expensive to build, accounting for up to 20% of a CTL plant's capital cost. Therefore, having a supply of oxygen from water-splitting offers a second process simplification.

These several process changes are illustrated in the flowsheets in Appendix B. For the same amount of coal input (18,800 tons/day), the use of hydrogen and oxygen from water-splitting increases the production of Fischer-Tropsch liquids from 26,000 barrels/day in the conventional CTL case to 58,200 barrels/day. The selectivity of carbon in the Fischer-Tropsch fuel product rises from 29.5% for the base case to 65.8% in this scenario.

### 3.2.2 Nuclear-Driven Hydrogen Production Technologies

This improvement requires an external source of hydrogen. The common way of producing large amounts of hydrogen is currently by steam reforming of methane. This use of another, more expensive fossil fuel offers no advantages over using coal in the baseline process, nor does it utilize a nuclear energy source.

The other present alternative for making hydrogen is electrolysis, using electric current to split water into hydrogen and oxygen. Though electrolysis is well established for making other chemicals, this method is too expensive for commodity production of hydrogen because of its low overall efficiency of 25 to 28%, based on about 33% efficiency of making electricity from heat followed by 75-80% efficiency of the electrolysis itself. With electrolysis a nuclear reactor could be used to generate the electricity, though any type of power plant would work. There is no advantage to a high temperature gas reactor for this step because conventional electrolysis operates near ambient temperature. A current-generation, light-water reactor could be used just as well, especially if its cost of electricity was lower.

Hydrogen production processes that take greater advantage of nuclear energy are under development in a major U.S. Department of Energy program. They fall into two broad categories: high-temperature electrolysis (HTE) and thermochemical cycles. Both rely on heat at near 850ºC to help split water, and thus require very high temperature reactors. These processes are far from commercial use, though, with both at scales of only a few hundred liters per hour of hydrogen production.

High temperature electrolysis systems split water using what is essentially a fuel cell operating in reverse. A ceramic membrane at about 800ºC allows oxygen ions, but not hydrogen ions, to be transported through it under the influence of an electric field. Aside from producing this selective ionic conductivity, the high temperature also reduces the amount of energy that must be added to a water molecule to split it into hydrogen and oxygen ions. The overall result is an electrical consumption that is about 70% of conventional electrolysis, the other 30% being replaced by a thermal input. The energy efficiency savings of using some of the heat directly, rather than to make electricity first, makes HTE attractive. However, the technology has development questions associated with building and electrically connecting assemblies of literally millions of small ceramic membranes.

Thermochemical cycles utilize heat at about 850ºC to drive a set of chemical reactions that in total achieve the splitting of water into hydrogen and oxygen. Many different cycles have been proposed. Some cycles use only heat, while others use a combination of heat and power for low-temperature electrolysis of sulfuric acid. The most promising at this point is the sulfur-iodine cycle, named after two of the key elements in its process chemistry. This process has development issues with finding materials of construction that will withstand the extremely corrosive conditions and with demonstrating the complex process flowsheet in an integrated fashion.

## 3.3 Carbon Recycle

The doubling of carbon utilization by the use of nuclear-generated hydrogen led to the second major process insight: that carbon from the coal should be used only to produce carbon monoxide for feed to the Fischer-Tropsch synthesis, with the nuclear plant supplying energy wherever it was needed (rather than burning carbon to generate heat). Two process changes resulted from this.

In the first modification, the tail-gas from Fischer-Tropsch synthesis, a mixture of light alkanes with some byproduct olefins and alcohols, is recovered rather than simply being burned to generate power in a gas turbine and heat recovery steam generator. The process makes more tail-gas than needed to supply its heat and power demand; thus, conventional CTL plants are net exporters of electricity. The objection to this net power production is that a large amount of processing (gasification, shift reaction, acid gas removal, and Fischer-Tropsch synthesis) has to be done to produce this fuel gas when, if electricity was a desired product, the coal could simply be burned directly. The tail-gas is a clean stream but is not directly reusable in Fischer-Tropsch synthesis. To recover the tail-gas's carbon values, it can be returned to the coal gasifier where the conditions already exist to easily convert it to synthesis gas, producing both carbon monoxide and some useful hydrogen.

The second modification is to recognize that the carbon dioxide purged from the acid gas removal section can also be recovered and converted to carbon monoxide. This conversion occurs via the reverse water-gas shift reaction:

$$H_2 + CO_2 \rightarrow H_2O + CO$$

This reaction occurs readily at the conditions in the coal gasifier, so the gas leaving the gasifier is at close to thermodynamic equilibrium. Even though some carbon dioxide will remain in the gasifier effluent, it will be recovered in the acid gas removal section for further recycling to the gasifier. The consequence is a recycle of carbon dioxide large enough that the amount of carbon dioxide converted in one pass is essentially equal to the net formation of carbon dioxide from the coal. This recycle does not mean that no carbon dioxide is purged from the process. There will be other minor species to be removed and some amount of carbon dioxide will be removed with them. The issue requires an optimization of the cost of recovering the last fraction of carbon dioxide compared to the cost penalty of simply discharging it into a vent stream.

With these two changes the process performance improves dramatically. The calculated carbon yield (carbon in synthetic fuel compared to carbon in the coal feed) rises from 29.5% in the base case to 95.7%. The remaining 4.3% appears mostly in a second, smaller carbon dioxide purge stream which could itself be minimized or recycled. At the level of detail of this study, the value of adjusting process conditions to further reduce this number was judged to be minimal. The point is established that essentially all the coal can be converted to synthetic fuel by focusing on carbon monoxide production and by using nuclear energy rather than combustion as the process energy source.

With this higher carbon selectivity, at the baseline coal feed rate of 18,800 tons/day this process configuration produces about 84,700 barrels/day of synthetic fuel or about 3.2 times as much product. Because the scale of the plant grew so much on the basis of equal feed rates, the process was scaled back to produce an equal amount of product, 26,000 barrels/day. In this case, only 5,800 tons/day of coal are needed. Complete results for this case with equal product rates are presented in Appendix C.

This higher yield of synthetic fuel has a price. By recovering more carbon monoxide from each ton of coal, there is a correspondingly increased demand for more hydrogen in the Fischer-Tropsch feed. This implies that more water hydrolysis and more nuclear reactors to power that system are needed. Even in the constant-product case of Appendix C, the thermal load from the nuclear reactors is projected to be 3223 $MW_{th}$, or six AREVA reactors operating at a nominal 600 $MW_{th}$ each.

This reactor capacity equals 0.56 $MW_{th}$ per ton/day of coal feed or a nuclear energy input of 22,700 BTU/lb of coal. The overall energy conversion efficiency of the process is about 37% when expressed as synthetic fuel energy produced per the total input of coal energy and nuclear energy. The fossil energy conversion ratio, not counting the nuclear energy contribution, is quite good. The total energy content of the Fischer-Tropsch product is about 136% of the coal's energy and is in a much more versatile and valuable form. The increase in energy is from the nuclear-produced hydrogen added to the coal's carbon.

The process changes during recycling of Fischer-Tropsch light tail-gas and carbon dioxide purge streams do not necessarily require a high-temperature nuclear reactor. However, when the Fischer-Tropsch tail-gas is recycled rather than burned, medium pressure saturated steam generated elsewhere in the process cannot be superheated for efficient power production. The availability of high temperatures from a nuclear reactor permits this, although a light-water reactor would suffice with lower energy efficiency. The more important role for the reactor concerns the higher carbon monoxide flows resulting from both of these recycles. This requires more hydrogen, and while a light-water reactor could produce that hydrogen, high-temperature processes promise to be more efficient when they are developed.

## 3.4 Additional Energy Integration

The flowsheets for which detailed ASPEN simulations were done did not include integration of the chemical plant's steam generation with the nuclear plant's power cycle (see Figure 3) because there was no model of the nuclear plant's electricity production cycles. However, this appears to be a practical way to potentially increase the overall efficiency of the facility while also reducing the capital cost. All of the current flowsheet simulations recover heat from hot gasifier product gas and the Fischer-Tropsch process to make steam that drives turbines used to generate electricity; this occurs in the flowsheet sections labeled "Heat Recovery & Steam Generation" and "Steam Turbine". This process configuration is a holdover from the baseline CTL process, which is a substantial net generator of power.

With a nuclear plant nearby, the steam generated in the chemical facility could instead be fed into the Rankine bottoming cycle used in the nuclear plant. This eliminates the capital cost of installing turbines in the synthetic fuel plant and potentially gains a modest equipment efficiency improvement by recovering that steam's energy in larger equipment. The steam-based Rankine cycle (see Figure 3) operates at three unspecified pressure levels, creating good flexibility to energy-integrate steam at several pressures from the synthetic fuel plant. Because this additional steam would use nuclear-generated heat solely for superheating (rather than boiling water too in the basic Rankine cycle), the better temperature matching would make HTGRs particularly efficient in a thermodynamic sense for coupling to this incremental power production.

Although the integration was not modeled, the mass and energy flows between the plants can be projected from the flows in the energy recovery section of the synthetic fuels plant. These flows (see Table 1) are not optimized either for heat recovery or power production, but are indicative of the scale of flows. They are based on 92% isentropic efficiency in condensing turbines, with the work numbers being the shaft work of the turbine before any mechanical, generation, or electrical transmission losses.

Because the amount of work represented by these steam streams is about equal to one HTGR, compared to the fleet of six HTGRs required for the complete facility, this interplant steam transfer should not create excessive design or operational coupling problems between the nuclear and CTL plants. Concerns about potentially leaking radioactivity into the synthetic fuels plant should be minimal because the steam in this bottoming cycle is isolated from the nuclear reactor core by two recirculation loops (of helium in the primary coolant loop and of a helium/nitrogen mixture in the secondary loop). Transferring the chemical process heat to the nuclear power cycle does not obviate one HTGR because credit for the net power production in the CTL plant has already been taken in this analysis. However, this change would eliminate the need for a separate set of turbines in the CTL plant at the cost of a small percentage in the size of the nuclear power cycle's turbines.

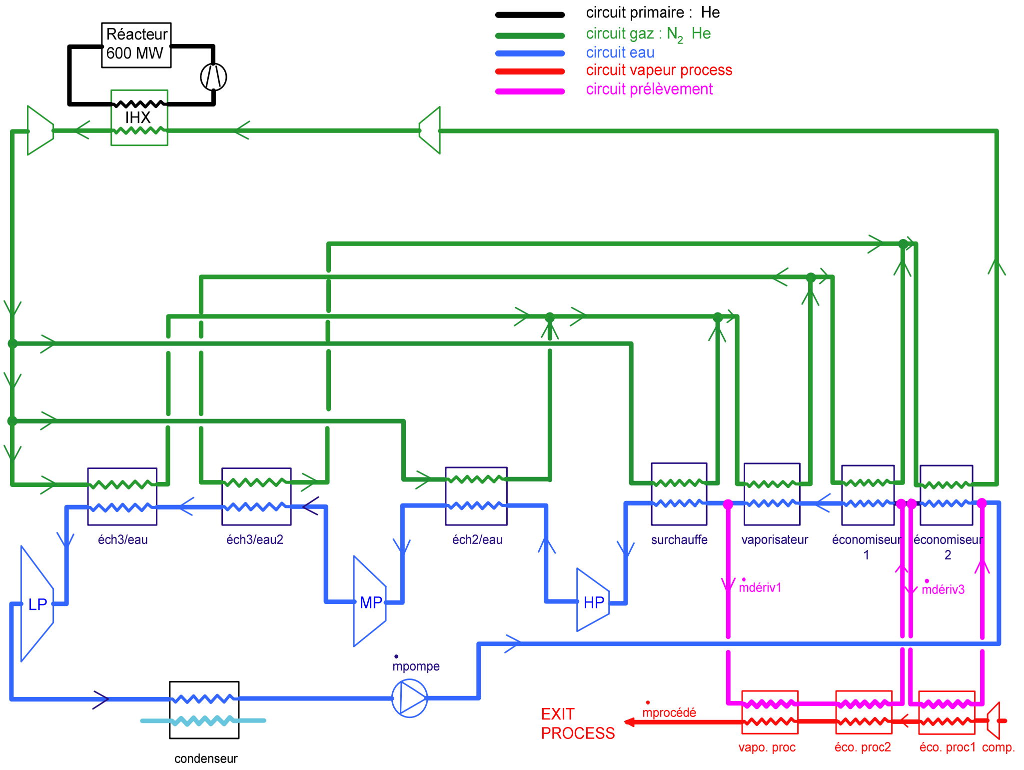

Figure 3. Power generation cycles driven by the AREVA high-temperature reactor (top), with Brayton cycle in middle and Rankine cycle at bottom.

Table 1. Synthetic fuels plant steam streams potentially integrable with the nuclear plant's power cycles.

| | |
|---|---|
| High pressure steam to turbines | 424,000 lb/hr |
| Work from high pressure turbine | 14.4 MW |
| Extraction of medium pressure steam for feed to the gasifier | 25,000 lb/hr |
| Additional saturated medium pressure steam from synthetic fuels plant | 1,560,000 lb/hr |
| Superheating of saturated medium pressure steam to 540ºC with nuclear heat | 147.6 MW |
| Work from medium and low pressure turbines | 233.4 MW |
| Net condensate return to synthetic fuels plant | 1,959,000 lb/hr |

# 4. ECONOMIC ANALYSIS

## 4.1 Capital Cost

A capital cost estimate was developed for the conventional and nuclear-assisted coal-to-liquids flowsheets producing equal amounts of Fischer-Tropsch product. The estimated costs for each configuration are summarized in Table 2. The complete capital cost estimate report is attached as Appendix D.

The following assumptions were used in generating the capital cost estimates:

- \$2,500/kW$_e$ and an overall efficiency of 40% were assumed for the nuclear plant; these values correspond to the \$1000/kW$_{th}$ used in the later sensitivity analysis.
- \$400/kW$_e$ (DOE target value) was assumed for the electrolyzer cost.
- Fischer-Tropsch synthesis and refining costs were scaled from a recent report from Rentech, a developer of Fischer-Tropsch processes [Corkadel and Pedersen (2005)].
- All other costs were scaled from existing National Energy Technology Laboratory (NETL) reports [NETL (2000) and Buchanan et al. (2003)].
- The six-tenths exponent rule was used in scaling major equipment costs [Peters and Timmerhaus (1991)].

Table 2. Summary of conventional and nuclear plant capital costs.

| Plant Area | Conventional Coal-to-Liquids Plant (Million \$) | Nuclear-Assisted Coal-To-Liquids Plant (Million \$) |
|---|---|---|
| Nuclear Plants | 0 | 3,223 |
| High-Temperature Electrolysis Plant | 0 | 545 |
| Air Separation Unit | 167 | 0 |
| Coal Preparation | 64 | 32 |
| Gasification & Heat Recovery | 289 | 144 |
| Synthesis Gas Cleanup | 209 | 72 |
| FT Synthesis & Refining | 160 | 160 |
| Power & Steam Production | 182 | 73 |
| Ancillary Systems, Engineering, & Permitting[†] | 364 | 164 |
| Subtotal | 1,435 | 4,413 |
| Contingency | 263 | 748 |
| Total | 1,698 | 5,161 |

[†] Does not include the nuclear and electrolysis plants; costs for these units are included in the amounts listed above.

The following observations can be drawn from the capital cost estimates:

- 69% of the total cost for the nuclear-assisted CTL plant is for the HTGRs and their contingency
- 16% of the total cost for the nuclear-assisted CTL plant is for the electrolyzers
- Gasification and cleanup costs for the nuclear-assisted CTL plant are substantially lower than for the conventional CTL plant due to smaller flows of coal and synthesis gas.

## 4.2 Operating Costs and Revenues

Because this report is an early-stage scoping analysis, the operating costs were not developed in great detail. The factors that were included in the scoping analysis are listed and explained in the following paragraphs. Some apply only to the conventional CTL process, others only to the nuclear-assisted process. The range and baseline values of each variable used to determine the product fuel cost are listed in Table 3.

*Stream factor:* Also known as availability, this is the percentage of a year that the plant actually makes product. Operation is assumed to always be at full capacity when actually running.

*Coal cost:* This is the delivered cost of Powder River Basin coal under a long-term supply contract. Low costs represent a mine-mouth plant location that does not require shipping, while higher costs could result from either transportation or higher market costs for coal.

*$CO_2$ penalty:* This is an assumed future penalty for release of carbon dioxide to the atmosphere or, equivalently, the regulatorily mandated cost to the CTL plant of sequestering the carbon dioxide, most likely by delivering it to a pipeline for geological disposal.

*Synthetic fuel:* This is the price at which the Fischer-Tropsch product is sold from the production facility, equivalent to a manufacturing cost (not a retail cost).

*Electricity:* This is the value at which electrical production above that needed by the plant can be sold. The conventional CTL plant is by design a net power exporter. Because there are six reactors in the nuclear alternative, it is assumed that electrical production will always be available from at least a few of them. Half of the maximum electrical production is assumed available for sale during the fraction of the year that the gasification and Fischer-Tropsch plants are not operating.

*Oxygen:* The nuclear-assisted alternative uses in the gasifier about 60% of the oxygen produced by electrolysis of water. The remainder can be sold on the merchant market.

*Nuclear reactor capital cost:* Rather than pricing each of the several reactors as package systems, a cost per unit of thermal output was applied to the projected total thermal demand. Normally the capital cost per unit of electrical capacity is cited. The capital cost calculation used in this capacity differs from that by the assumed efficiency of converting heat to electricity.

*Operations/maintenance cost for a conventional and nuclear plant:* This is expressed as a percentage of the capital cost expended per year for labor and supplies to operate and maintain the plants. Different values are used for the chemical and nuclear alternatives. Values for the nuclear power industry were applied to the entire nuclear-assisted process, because the cost of the nuclear reactors is the majority of that plant cost.

*Required return on investment:* This is the percentage of capital cost that must be generated each year to justify making the initial investment.

Costs or revenues related to the overall plant financial performance, such as depreciation, taxes, and governmental incentives, were not included. The one exception to this was to consider the effects of a emissions penalty associated with production of carbon dioxide.

## 4.3 Projected Fuel Costs

A spreadsheet calculation combined the cost assumptions (see Table 3) with the process performance from the ASPEN simulations (see Appendix C) to estimate the overall plant economic performance and the required selling price of the Fischer-Tropsch product. The results are presented in Table 4, where the key process performance values have been entered at the top of the table for use by the spreadsheet. The value for net electrical production for the nuclear plant is the year-round average of assuming that during the synthetic fuels plant's downtime half of the nuclear reactors would be able to produce electric power at full rates. However, the actual production would be in intermittent periods of much higher megawattage.

Table 3. Cost variables used for economic projections.

| Cost Factor | Baseline value | Variable distribution | Upper value or mean* | Lower value or std dev* |
|---|---|---|---|---|
| Stream factor | 90% | uniform | 85% | 95% |
| | | | | |
| Coal cost, \$/ton | 20 | normal | 25 | 5 |
| $CO_2$ penalty, \$/ton | 50 | uniform | 0 | 100 |
| FT liquids value, \$/gal | 3.00 | not used | | |
| Electricity value, \$/kWhr | 0.05 | normal | 0.055 | 0.0075 |
| $O_2$ value, \$/ton | 20 | uniform | 0 | 40 |
| | | | | |
| Nuclear reactor capital cost, \$/kW$_{th}$ | 1000 | uniform | 700 | 1300 |
| Operations, maintenance, insurance – coal process, %capital/yr | 7.0% | normal | 7% | 1% |
| Operations, maintenance, insurance – nuclear process, %capital/yr | 5.0% | normal | 5% | 0.5% |
| | | | | |
| Required return on investment | 15% | uniform | 12% | 18% |
| *Upper/lower values apply to uniform distributions, mean and standard deviation to normal distributions | | | | |

Table 4. Fuel cost calculations for conventional and nuclear-assisted CTL.

| Stream Flow or Cost | Conventional | Nuclear |
|---|---|---|
| Coal feed, ton/day | 18,840 | 5,793 |
| $CO_2$ sequestered, lb/hr | 985,531 | |
| $CO_2$ release, lb/hr | 572,436 | 32,736 |
| FT liquids, bbl/day | 26,037 | 26,036 |
| Net oxygen production, lb/hr | | 239,600 |
| Average net power generation, MW | 334 | 64 |
| Nuclear plant capacity, $MW_{th}$ | | 3,223 |
| | | |
| REVENUES | | |
| Synthetic fuel, $M/yr | 1,078 | 1,078 |
| Electricity production, $M/yr | 132 | 28 |
| $O_2$ sales, $M/yr | | 19 |
| | | |
| EXPENSES | | |
| Coal cost, $M/yr | 124 | 38 |
| $CO_2$ penalty/cost, $M/yr | 307 | 6 |
| Operations, maintenance, insurance, $M/yr | 119 | 258 |
| Net operating income, $M/yr | 660 | 822 |
| | | |
| CAPITAL | | |
| Engineer, procure, construct, $M | 1,435 | 4,413 |
| Contingency, $M | 263 | 748 |
| Total cost, $M | 1,698 | 5,161 |
| | | |
| Pretax simple return on investment (ROI) ( at known fuel value) | 38.9% | 15.9% |
| Required synthetic fuel revenue at specified ROI ($M/yr) | 673 | 1,030 |
| Fuel cost for specified ROI ($/gal) | 1.87 | 2.87 |

In Table 4 the net operating income is the net of revenues and expenses. The capital cost was calculated using fixed values for the conventional CTL plant and for the chemical part of a nuclear-assisted process; only the cost of the nuclear reactors and their contingency varied in response to the assumed cost per unit of thermal capacity.

The financial results are calculated in two ways. Under the assumption of a known selling price for the product fuel ($3.00/gal in the base case), the pretax simple return on investment is calculated. Alternatively, for a specified target return on investment (15% in the base case) the required sales price of fuel is determined. At the reference conditions, conventional CTL fuel must sell for $1.87/gal, while the nuclear-assisted product requires $2.87/gal. The difference is largely due to the high capital cost of the nuclear reactors, which affects both the capital recovery charge and the annual combined cost for operations, maintenance, and insurance, which are a percentage of the capital cost. On the other hand, the nuclear system is much less sensitive to changes in the price of coal or prospective carbon dioxide emissions penalties. As illustrated in Figure 4, the three lines for the conventional process show that its product cost varies by a factor of five over the coal and carbon dioxide cost ranges shown, while the nuclear process has only about a 25% variation in product fuel price over the same range with almost no effect of varying carbon dioxide costs.

A more thorough analysis of cost sensitivity was done by performing a Monte Carlo analysis on all the input variables affecting fuel cost. The distribution for each of the variables is listed in Table 3. A total of 20,000 trials were run to forecast the fuel price distribution shown in Figure 5. The median of this distribution is $2.87/gal, while the mean and standard deviations are $2.85 and $0.41 respectively. This makes the 95% probability range $2.05 - $3.65/gal under the price assumptions used.

This cost, especially as an early stage estimate with much uncertainty, does not appear unreasonable. Current retail prices for motor fuels are about $3/gal, but those include costs for blending, distribution, and retail profit margin that are not in the Fischer-Tropsch estimates. A fuel retail price per gallon of $4 or more (as $3 production cost implies) does not seem implausible for the foreseeable future. The Fischer-Tropsch product will be particularly desirable to refineries formulating and distributing fuels because, with a desirable high cetane number resulting from its molecular composition, it is an especially good component for blending into petroleum-based diesel fuels. With the improving performance of diesel engines for passenger vehicles, especially when contrasted with the reduced energy content of ethanol-blended gasoline, the market outlook for this product seems positive.

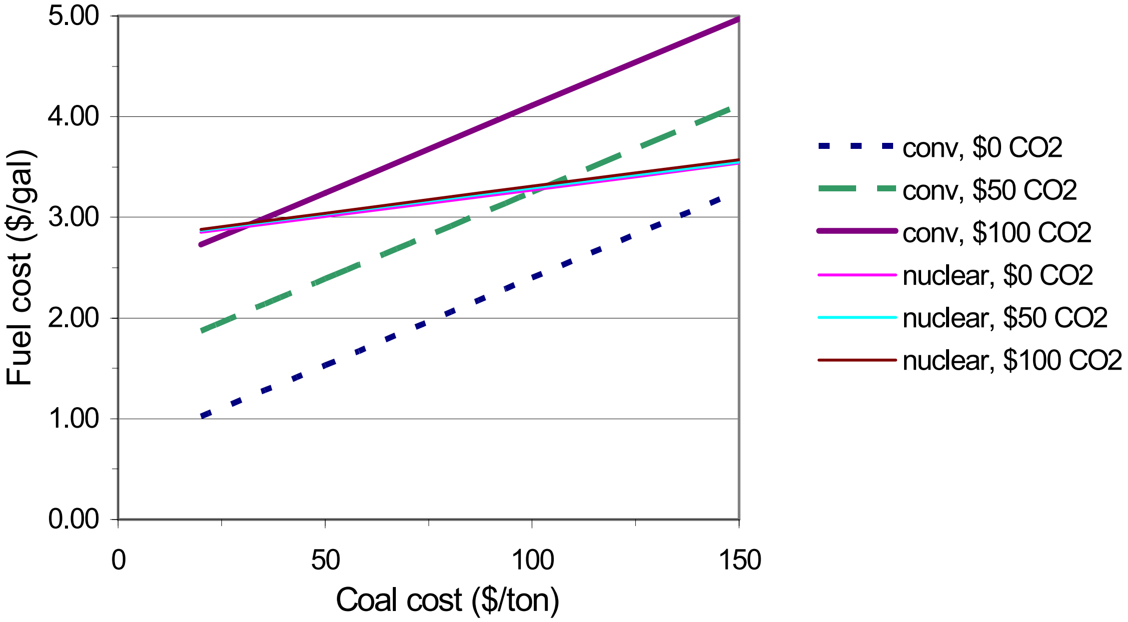

Figure 4. Product cost from conventional and nuclear-assisted CTL processes at various coal costs and carbon dioxide production penalties ($/ton).

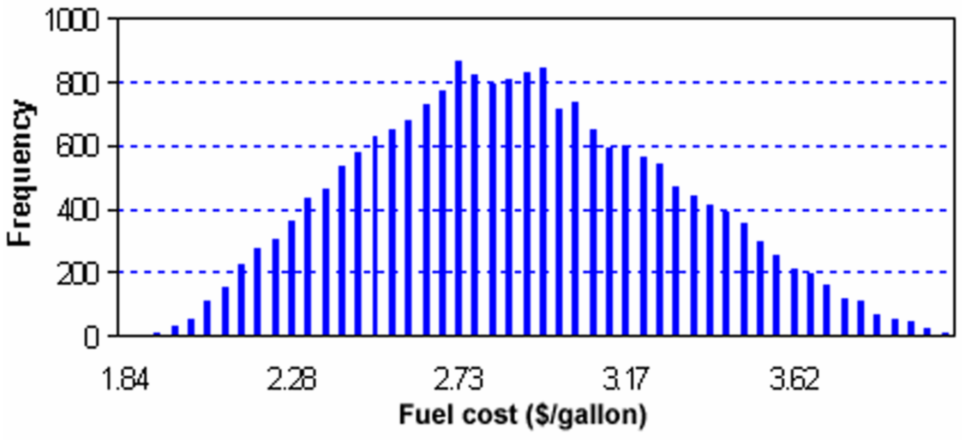

Figure 5. Potential distribution of nuclear-assisted synthetic fuel manufacturing cost.

The cost of the Fischer-Tropsch fuel can be seen from a different perspective. Because the overall effect of a CTL process is to raise the hydrogen to carbon mole ratio from the coal's value of about 1 to liquid fuel's value near 2, making synthetic fuel requires a very large scale production of hydrogen. Using the midpoint of the DOE's projected large scale hydrogen cost range of $2-3/kg, the hydrogen needed for synthesizing the Fischer-Tropsch feed mixture contributes $2.11/gal to the final cost, about three fourths of the total fuel cost.

In an alternative view of the cost of this synthetic fuel, the cost per unit of energy of the Fischer-Tropsch product is equivalent to electricity at about $0.08/kWh, a value that is high but not unseen in the U.S. Given the volatility of fuel prices and the likelihood of future carbon emissions penalties, the projected performance of a nuclear CTL process appears to be worth further evaluation.

# 5. FEASIBILITY OF NUCLEAR INTEGRATION WITH CTL PROCESSES

## 5.1 Assessment of Results

The use of high-temperature nuclear reactors to augment conventional CTL processing works well and offers significant potential advantages. The integration is one of electricity and process feeds rather than heat flows, with those process flows carrying the nuclear energy as chemical energy (see Figure 6). This allows the three major process technologies in this complete system (nuclear power plant, water electrolysis, and indirect coal liquefaction via gasification and Fischer-Tropsch synthesis) to operate in much the same manner as they traditionally have. The modifications proposed within the fuel synthesis process to recycle carbon and increase the overall yields are not technically difficult and should not present operational problems. Handling of the other major emissions from a coal gasification plant, hydrogen sulfide and mercury, is done in the conventional ways at the conventional point in the process. Integration with a nuclear energy source does not change that.

The sizing of the integrated process described in this report was somewhat arbitrary. An existing ASPEN model of a coal gasification plant was used for the base case, and this established the coal feed rate of 18,800 tons/day and the Fischer-Tropsch product rate of 5800 barrels/day. That coal feed rate is comparable to a large coal-fired conventional power plant. As the simulation work progressed, the heat from the nuclear plant was treated as though it came from a single source. In a more thorough engineering analysis, there would have to be significant attention paid to the number of reactors required, as they each have a fixed maximum thermal output. Here the final demand was for 3223 $MW_{th}$, or 5.4 reactors of 600 MW thermal output. For the best economic performance, the design capacity of the CTL plant must be adjusted to use an integral number of nuclear reactors.

The number of reactors needed also depends on how much downtime they will have and how well that downtime can be integrated with the CTL plant's operating schedule. With six nuclear reactors, it is reasonable to assume that several of them will be available at all times. The nominal performance expected of these reactors is 90% availability on an annual basis, with the downtime being about 4% unplanned and 6% planned. (These performance figures are similar to what a coal-fired power plant or a

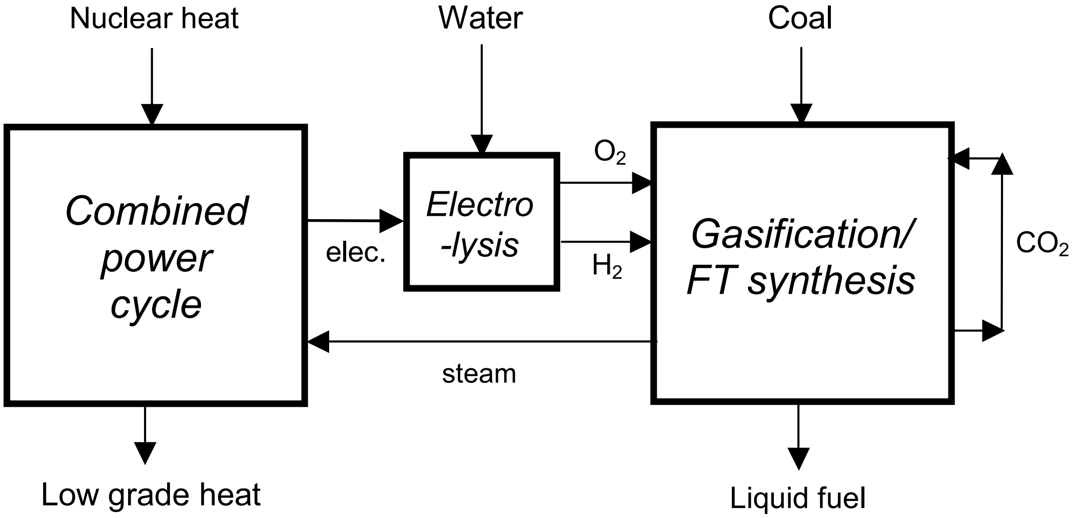

Figure 6. Arrangement of major process sections.

standard chemical plant might achieve.) The reactors require about three weeks of down time every eighteen months for fuel replacement. The scheduling of fuel replacement—whether one reactor every three months, two at a time every six months, or some other plan—would be coordinated with the maintenance needs of the CTL plant and, perhaps, with the ability of the CTL to operate efficiently at reduced rate when not all nuclear plants were in service.

The ultimate success of nuclear-assisted coal-to-liquids technology will probably not be based on the economics of its cost of production. A conventional CTL plant can produce the same product for less cost than the nuclear hybrid in an energy policy environment similar to today's. But over time petroleum will become increasingly scarce and therefore more expensive, while the total global demand for automobiles, or more specifically for miles of driving, will increase. Hydrocarbon liquid fuels are and will remain highly suited for vehicle use because of their energy density and ease of use. The demand for alternative liquid vehicle fuels, whether ethanol, biodiesel, methanol from natural gas, or coal-derived Fischer-Tropsch products, will remain strong. The discriminator between these fuels will, in part, be the taxes and penalties imposed on them to address the release of carbon dioxide associated with their production and use. Biomass derived fuels will have a major advantage in this situation, but they cannot meet the expected demand for fuels. Coal and natural gas derived fuels will be doubly penalized for their net emissions when they are burned and for the copious carbon dioxide emitted during their production. The nuclear-assisted CTL process can avoid the production-related penalties because, with 97% carbon conversion efficiency, it emits almost no carbon dioxide. CTL synthetic fuel will be a net producer of carbon dioxide because the coal's carbon does eventually end up in the atmosphere. Nevertheless, it will be a useful intermediate term option as petroleum declines as a resource and before hydrogen as fuel becomes common, assuming its storage problems can be addressed.

The much higher carbon conversion efficiency of this nuclear-assisted synthetic fuels process has strategic value for domestic U.S. energy policy. The large reserves of coal in the U.S. have led to much of the interest in CTL processes. Those reserves correspond to about 200 years of use at current rates. However, if that coal is called upon to supply the current demand for energy from coal plus that from oil, there is only about 85 years of supply. That number makes no allowance for future growth in energy consumption, nor does it account for less than 100% conversion efficiencies in making synthetic fuel from coal. If coal must also eventually replace the demand for natural gas as well, the lifetime of U.S. reserves drops to 65 years. Allowing for demand growth, it is conceivable that the major fossil energy reserves might not outlast the plants designed to convert them. Such issues, though they will be debated and refined at length, point to a great future value on high processing efficiency as delivered by the nuclear-assisted CTL process.

## 5.2 Recommended Research and Development Topics

This study found that high-temperature nuclear reactors can complement a conventional CTL process to both extend the life of coal reserves and to greatly reduce the amount of carbon dioxide produced during the manufacture of synthetic liquid fuels. It has not addressed or even identified many questions to be answered before this actually happens. A variety of engineering studies can be done immediately to better understand CTL-nuclear integration. Some worthwhile candidates are outlined here, presented in order of decreasing potential impact on the overall technology integration.

*Evaluate Whether a Lower Nuclear System Temperature Can Be Used*

There are advantages in the design of the AREVA high-temperature reactor if the heat delivery temperature can be dropped from the nominal 850ºC to perhaps 750ºC. Because there is very high-temperature process heat available from heat recovery after the coal gasification process, it is likely the CTL process can be reconfigured to use heat produced by the gasifier for the highest temperature demands, with the nuclear heat being used elsewhere. This task could become rather involved if the total heat integration of the gasification/Fischer-Tropsch synthesis process must be reconsidered. A concern would be that if the current relatively separate operation of the main process units is sacrificed for tighter heat integration, it may possibly increase the propagation of upsets between the separate processes. If this risk can be limited, the technical benefits could be substantial.

### *Evaluate Integration of Low-Temperature Steam Production with the Nuclear Power Cycles*

As described in section 3.4 of this report, the energy recovery in the synthetic fuels process can be integrated with the Rankine bottoming cycle of the nuclear plants. The way in which this integration is done, its benefits, and its potential operating problems can be reviewed at several levels of detail. This study would be a natural complement to the previous one on lower temperatures from the nuclear reactor's primary coolant loop.

### *Address Water Usage Issues*

A CTL plant is a large user of water for three different purposes: process water for feed to the gasifier and to electrolysis; boiler feed water for steam generation as part of energy recovery; and cooling water to remove low-level heat from the plant. All three of these uses are net consumers of water. In the final version of the flowsheet, the high-temperature electrolysis alone consumed 1400 gals/minute to make hydrogen and oxygen. Boiler feed-water and cooling-water systems, even though they are nominally closed loop processes, both purge some water to prevent the buildup of contaminants in the system and they consequently have a demand for makeup water. Cooling water systems have a large additional demand to compensate for evaporative losses. The consumption of water by energy processes is a topic of growing concern because water availability can be problematic in the western U.S. where many coal fields are located. There is a need for a complete inventory of the expected water use in a nuclear-assisted CTL process, as well as investigation of ways to recover water from the gasifier effluent and the Fischer-Tropsch process. More speculatively, a plant located at the coal mine would have access to large amounts of low quality water removed from the mine; ways to upgrade that mine water using heat or energy available in the CTL process could be explored.

### *Refine the Carbon Recycling Strategy*

In this work the Fischer-Tropsch tail-gas and the carbon dioxide purge streams were simply returned to the gasifier to be converted to carbon monoxide. While this step would work, it does not take advantage of two key facts: these are both clean gas streams, and they can be reacted with each other. A preliminary simulation of catalyzing the water-gas shift reaction on a mixture of these streams and some added steam indicates that a synthesis gas mixture can be made with the proper hydrogen to carbon monoxide ratio to feed directly to the Fischer-Tropsch synthesis step. This avoids enlarging the gasifiers to handle the throughput of recycled gas and allows use of a catalyst optimized for this clean gas mixture. A more detailed heat and material balance around this operation would be valuable.

### *Explore Producing Hydrogen and Electricity While the Synthetic Fuels Plant Is Down*

With six nuclear reactors needed for this facility, several of them will be operable even if some of the reactors and the synthetic fuels plant are down for maintenance. This task would explore the process, operations, and economic issues with running the available reactors to make electricity or hydrogen for sale on this intermittent basis. Part of this task would be to project the expected operating and maintenance schedule for this small fleet of reactors, including unplanned outages.

### *Evaluate Other Fuel Types as Alternative Products*

This report has used Fischer-Tropsch liquids as the target product. However, depending on market demands, other fuels could be made from synthesis gas, including methanol, ethanol, dimethyl ether, or synthetic natural gas. The oxygenated fuel molecules have a potential process advantage because their synthesis would require somewhat less hydrogen production by electrolysis, and, therefore, less nuclear-generated electricity and capital cost. This is because the oxygen atom in the feed synthesis gas mixture is not removed as part of a water molecule as in Fischer-Tropsch synthesis.

### *Identify Markets or Integration Opportunities for the Unused Oxygen Production*

Somewhat over half of the oxygen produced in the water electrolysis step is used in the coal gasifier. The remainder, about 240,000 lbs/hr, is available for sale. Identification of users of this large amount of oxygen would be valuable, particularly if those processes can be integrated in other ways with a nuclear-assisted CTL plant.

### *Optimize the Acid-Gas Removal Strategy*

With the process improvements described in this report, much less $H_2S$ and carbon dioxide must be rejected from the process. The concentrations of these gases in the process stream from which they are removed will also be different. Both of these changes make it possible that the current acid-gas removal process, Rectisol, may no longer be the best choice in terms of its performance and its operating and capital costs. This process operates at temperatures of about -60 ºC and is therefore expensive to build and operate, but it does achieve very low residual sulfur levels as needed by current Fischer-Tropsch catalysts. Improvements in those catalysts or in other acid-gas removal processes may mean that the Rectisol process can be replaced with something simpler. This is not a critical issue since this is not the major cost factor in producing Fischer-Tropsch fuel, but it is worthy of eventual reevaluation.

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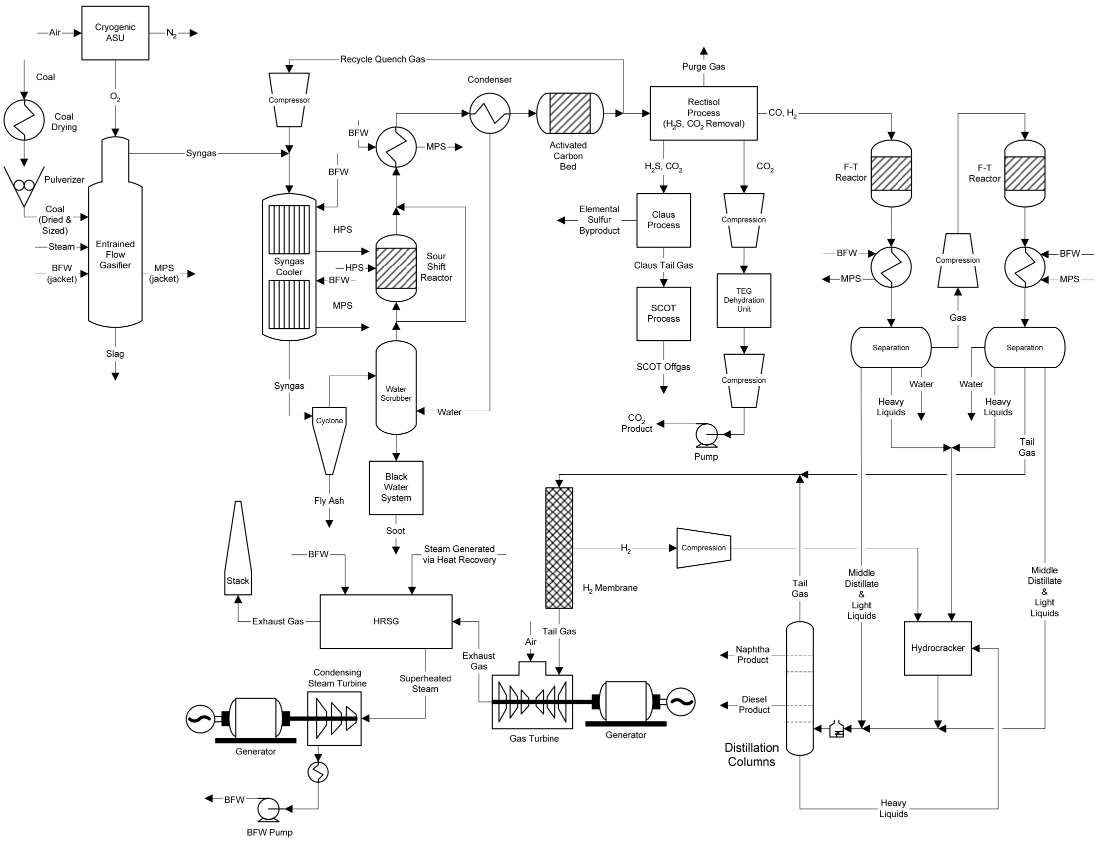
# Appendix A

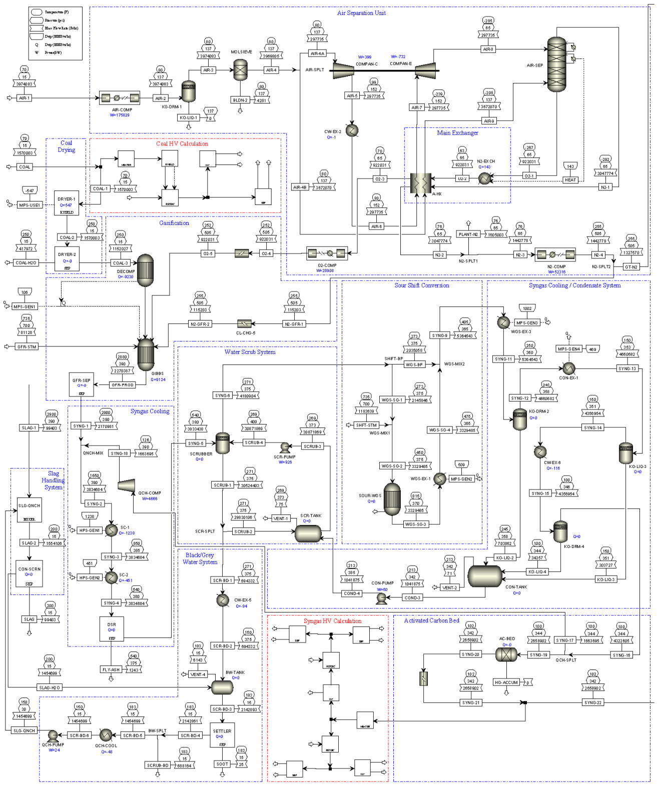
# Reference CTL Flowsheet


Air
Cryogenic ASU
$N_2$
Coal
$O_2$
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
MPS (jacket)
Slag
Syngas
Recycle Quench Gas
Compressor
BFW
BFW
MPS
Condenser
Activated Carbon Bed
HPS
Syngas Cooler
HPS
BFW
Sour Shift Reactor
MPS
Syngas
Cyclone
Water Scrubber
Water
Fly Ash
Black Water System
Soot
Purge Gas
Rectisol Process ($H_2S$, $CO_2$ Removal)
CO, $H_2$
$H_2S$, $CO_2$
$CO_2$
Elemental Sulfur Byproduct
Claus Process
Compression
Claus Tail Gas
TEG Dehydration Unit
SCOT Process
SCOT Offgas
Compression
$CO_2$ Product
Pump
F-T Reactor
F-T Reactor
BFW
MPS
Compression
BFW
MPS
Gas
Separation
Separation
Water
Water
Heavy Liquids
Heavy Liquids
Tail Gas
BFW
Steam Generated via Heat Recovery
Stack
Exhaust Gas
HRSG
$H_2$
Compression
$H_2$ Membrane
Tail Gas
Middle Distillate & Light Liquids
Middle Distillate & Light Liquids
Tail Gas
Air
Exhaust Gas
Superheated Steam
Condensing Steam Turbine
Generator
Gas Turbine
Generator
BFW
BFW Pump
Naphtha Product
Diesel Product
Hydrocracker
Distillation Columns
Heavy Liquids

```
CARBON BALANCE SUMMARY:

    PRB COAL FEED RATE =            18,840 TON/DAY

    % TOTAL CARBON TO LIQUID FUEL =   29.5 %
    % TOTAL CARBON TO SLAG & FLYASH =  0.2 %
    % TOTAL CARBON TO CO2   =         70.3 %

FUEL PROPERTIES:
                             #2 DIESEL       NAPHTHA
    PRODUCTION RATE, BBL/DAY   12327.        13710.
    MW                           201.8          92.8
    API GRAVITY                   52.8          78.3
    DENSITY, LB/GAL                6.29          5.61
    CETANE NO.                    96.3          36.0
    HEAT CONTENT, BTU/LB       20349.        20641.
    D86T DIST. CURVE, DEG. C:
        0%                       190.          -79.
        10%                      222.           28.
        20%                      229.           58.
        50%                      257.          103.
        90%                      313.          169.
        100%                     341.          195.

POWER PRODUCTION CALCULATIONS:

    GAS TURBINE POWER OUTPUT =           310.8 MW
    STEAM TURBINE POWER OUTPUT =         377.5 MW

    COAL GRINDING POWER CONSUMPTION =     24.3 MW
    ASU POWER CONSUMPTION =              256.7 MW
    PROCESS POWER CONSUMPTION =            5.7 MW
    SCOT PROCESS POWER CONSUMPTION =       0.2 MW
    RECTISOL POWER CONSUMPTION =          10.8 MW
    REFRIGERATION POWER CONSUMPTION =     25.1 MW
    CO2 COMPRESSION POWER CONSUMPTION =   23.3 MW
    FISCHER TROPSCH POWER CONSUMPTION =    3.9 MW
    PRODUCT UPGRADE POWER CONSUMPTION =    3.4 MW
    COOLING TOWER POWER CONSUMPTION =      7.6 MW
    POWER BLOCK POWER CONSUMPTION =        4.7 MW
    TRANSFORMER POWER LOSSES =             3.4 MW

    NET POWER GENERATED =                319.2 MW

EFFICIENCY CALCULATIONS:

    HEAT IN (HHV BASED):
        COAL HEAT CONTENT =             13270.0 MMBTU/HR

    HEAT OUT (HHV BASED):
        NET POWER OUTPUT =               1089.1 MMBTU/HR
        DIESEL + NAPHTHA HEAT CONTENT =  5540.5 MMBTU/HR
        TOTAL HEAT OUT =                 6629.6 MMBTU/HR

         PLANT EFFICIENCY (HHV BASED) =         50.0 %
```

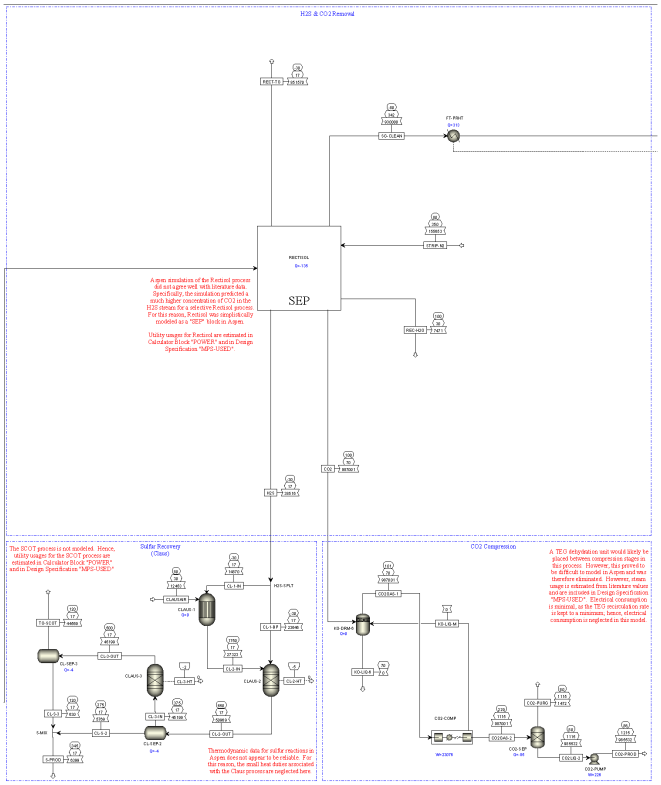
## Synfuel Production Baseline - Powder River Basin Coal

Aspen Plus 12.1 Run:CTL Baseline 8/15/2006 4:41:51 PM



## Synfuel Production Baseline - Powder River Basin Coal

Aspen Plus 12.1 Run: CTL Baseline 8/15/2006 4:41:52 PM



Aspen simulation of the Rectisol process did not agree well with literature data. Specifically, the simulation predicted a much higher concentration of CO2 in the H2S stream for a selective Rectisol process. For this reason, Rectisol was simplistically modeled as a "SEP" block in Aspen.

Utility usages for Rectisol are estimated in Calculator Block "POWER" and in Design Specification "MPS-USED".

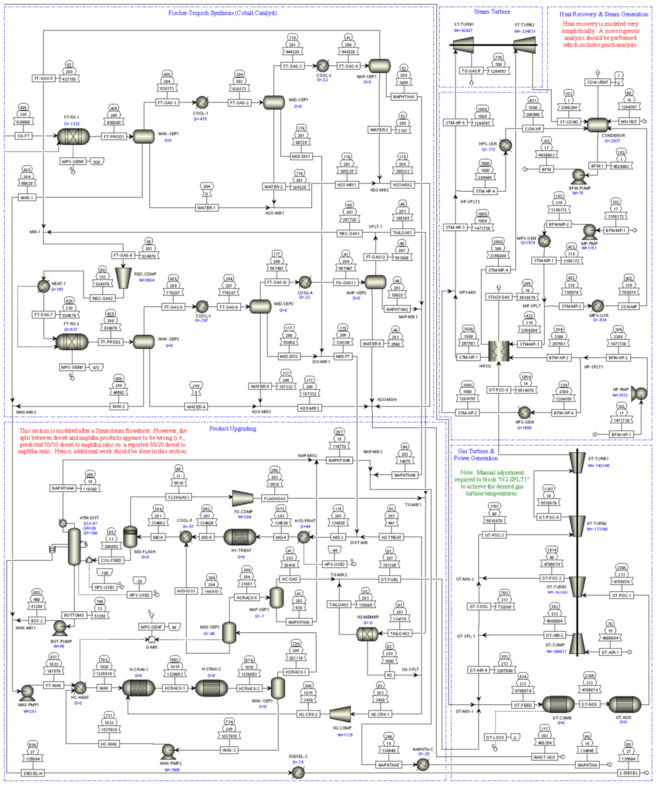
The SCOT process is not modeled. Hence, utility usages for the SCOT process are estimated in Calculator Block "POWER" and in Design Specification "MPS-USED"

Thermodynamic data for sulfur reactions in Aspen does not appear to be reliable. For this reason, the small heat duties associated with the Claus process are neglected here.

A TEG dehydration unit would likely be placed between compression stages in this process. However, this proved to be difficult to model in Aspen and was therefore eliminated. However, steam usage is estimated from literature values and are included in Design Specification "MPS-USED". Electrical consumption is minimal, as the TEG recirculation rate is kept to a minimum; hence, electrical consumption is neglected in this model.

# Synfuel Production Baseline - Powder River Basin Coal

Aspen Plus 12.1 Run:CTL Baseline 8/15/2006 4:41:52 PM



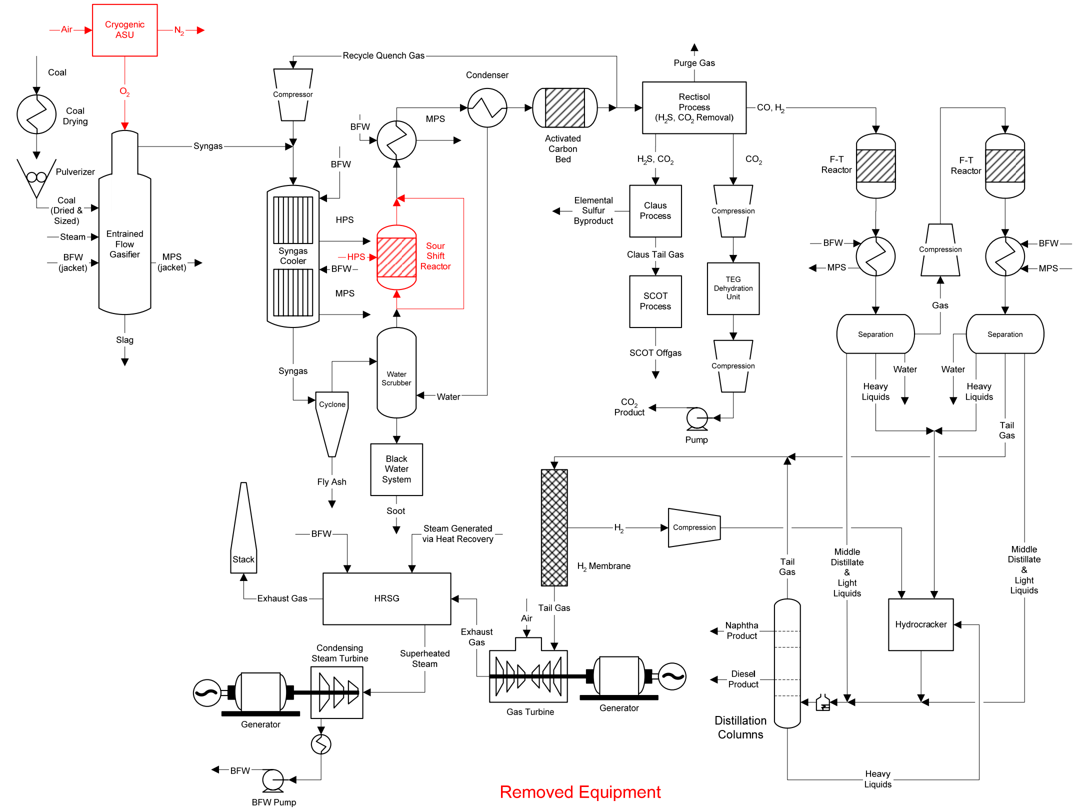
Heat recovery is modeled very simplistically. A more rigorous analysis should be performed which includes pinch analysis.

This section is modeled after a Syntroleum flowsheet. However, the split between diesel and naphtha products appears to be wrong (i.e., predicted 50/50 diesel to naphtha ratio vs. a reported 80/20 diesel to naphtha ratio. Hence, additional work should be done in this section.

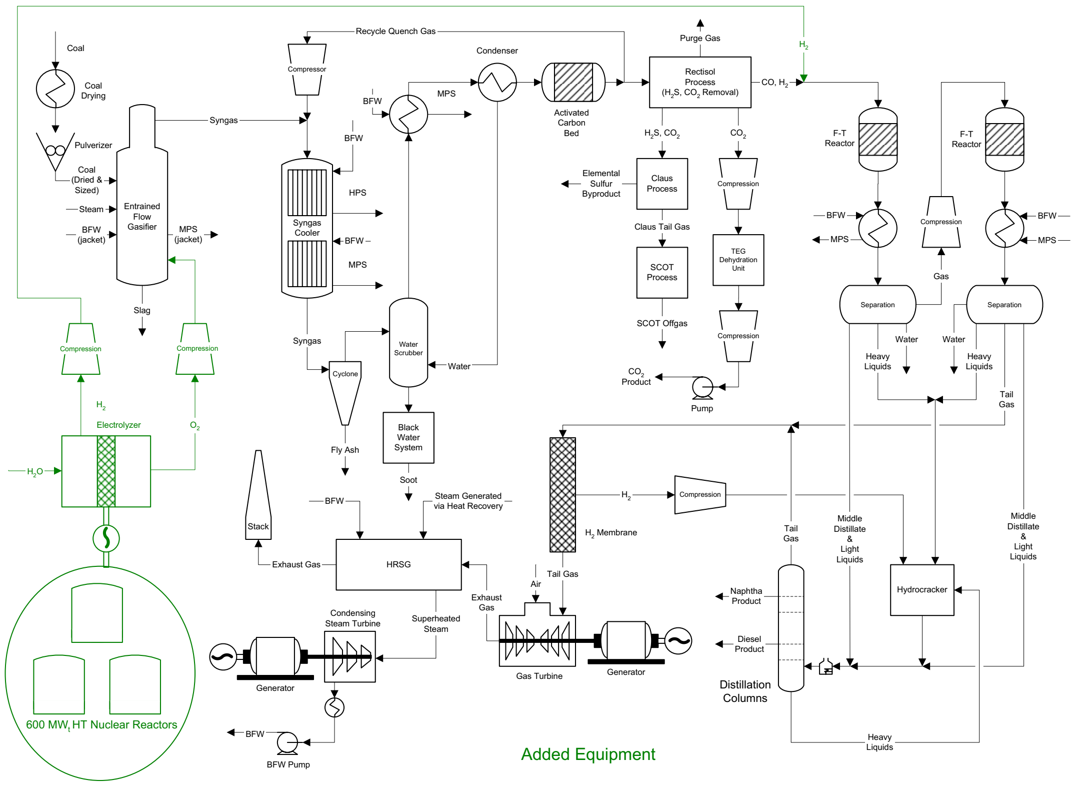
Note: Manual adjustment required to block "N2-SPLT1" to achieve the desired gas turbine temperatures.

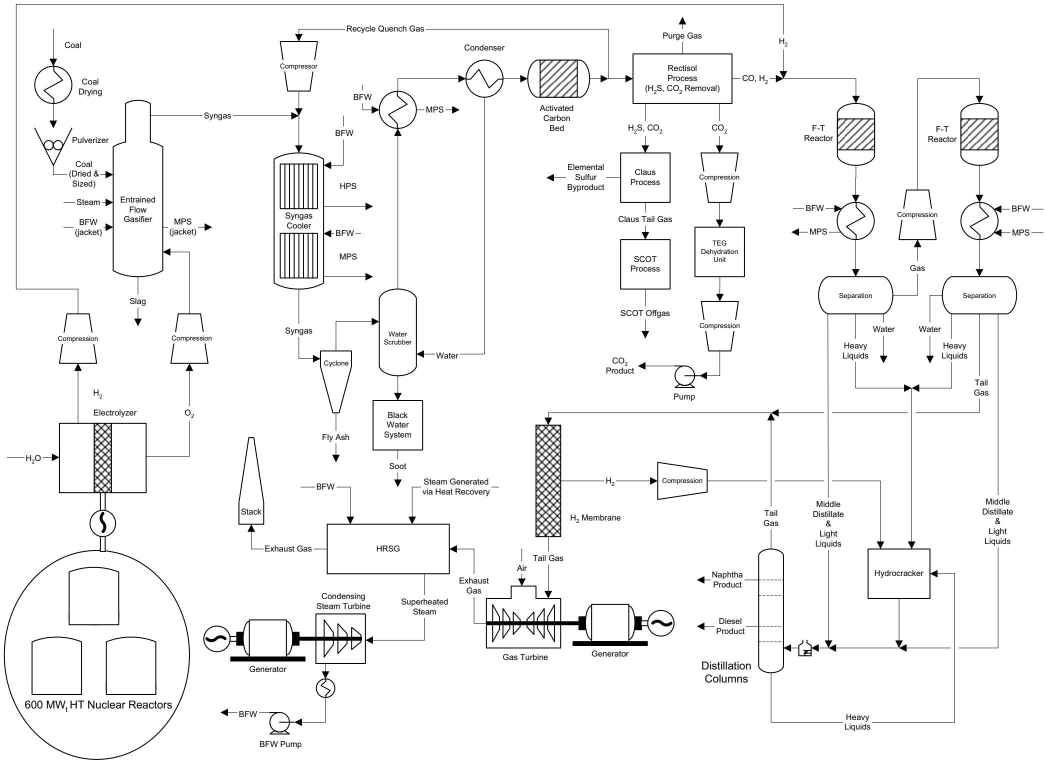
# Appendix B

# CTL Flowsheet with Nuclear Hydrogen Generation



Removed Equipment

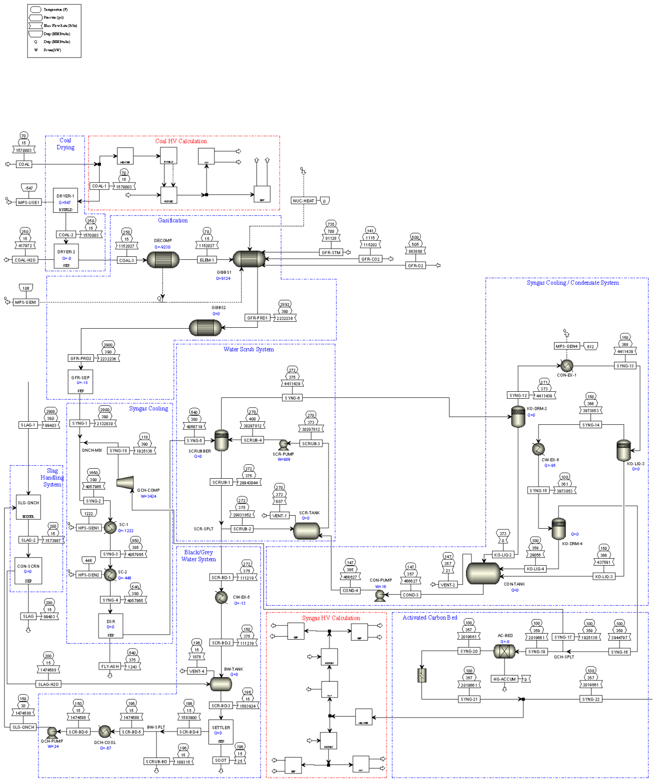

Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
MPS (jacket)
Slag
Syngas
Recycle Quench Gas
Compressor
BFW
Syngas Cooler
HPS
BFW
MPS
Syngas
Cyclone
Fly Ash
BFW
MPS
Condenser
Water Scrubber
Water
Black Water System
Soot
Activated Carbon Bed
Purge Gas
Rectisol Process ($H_2S$, $CO_2$ Removal)
CO, $H_2$
$H_2$
$H_2S$, $CO_2$
$CO_2$
Elemental Sulfur Byproduct
Claus Process
Claus Tail Gas
SCOT Process
SCOT Offgas
Compression
TEG Dehydration Unit
Compression
$CO_2$ Product
Pump
F-T Reactor
F-T Reactor
BFW
MPS
Compression
Gas
BFW
MPS
Separation
Separation
Water
Water
Heavy Liquids
Heavy Liquids
Tail Gas
Stack
BFW
Steam Generated via Heat Recovery
Exhaust Gas
HRSG
Exhaust Gas
Superheated Steam
Condensing Steam Turbine
Generator
BFW
BFW Pump
$H_2$
Compression
$H_2$ Membrane
Air
Tail Gas
Gas Turbine
Generator
Tail Gas
Middle Distillate & Light Liquids
Middle Distillate & Light Liquids
Hydrocracker
Naphtha Product
Diesel Product
Distillation Columns
Heavy Liquids
$H_2$
Compression
Compression
$H_2$
$O_2$
Electrolyzer
$H_2O$
600 MW$_t$ HT Nuclear Reactors
Added Equipment


Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
MPS (jacket)
Slag
Syngas
Recycle Quench Gas
Compressor
BFW
HPS
Syngas Cooler
BFW
MPS
Syngas
Cyclone
Fly Ash
BFW
MPS
Condenser
Water Scrubber
Water
Black Water System
Soot
Activated Carbon Bed
Purge Gas
Rectisol Process ($H_2S$, $CO_2$ Removal)
CO, $H_2$
$H_2$
$H_2S$, $CO_2$
$CO_2$
Elemental Sulfur Byproduct
Claus Process
Claus Tail Gas
SCOT Process
SCOT Offgas
Compression
TEG Dehydration Unit
Compression
$CO_2$ Product
Pump
F-T Reactor
F-T Reactor
BFW
MPS
Compression
Gas
BFW
MPS
Separation
Separation
Water
Heavy Liquids
Water
Heavy Liquids
Tail Gas
Compression
Compression
$H_2$
Electrolyzer
$O_2$
$H_2O$
600 $MW_t$ HT Nuclear Reactors
Stack
Exhaust Gas
BFW
Steam Generated via Heat Recovery
HRSG
Superheated Steam
Exhaust Gas
Condensing Steam Turbine
Generator
BFW
BFW Pump
$H_2$
Compression
$H_2$ Membrane
Tail Gas
Air
Gas Turbine
Generator
Tail Gas
Middle Distillate & Light Liquids
Middle Distillate & Light Liquids
Hydrocracker
Naphtha Product
Diesel Product
Distillation Columns
Heavy Liquids

```
CARBON BALANCE SUMMARY:
    PRB COAL FEED RATE =            18,840 TON/DAY

    % TOTAL CARBON TO LIQUID FUEL =   65.8 %
    % TOTAL CARBON TO SLAG & FLYASH =  0.2 %
    % TOTAL CARBON TO CO2   =         34.0 %

FUEL PROPERTIES:
                             #2 DIESEL       NAPHTHA
    PRODUCTION RATE, BBL/DAY   27198.        30984.
    MW                           201.6          92.3
    API GRAVITY                   52.8          78.6
    DENSITY, LB/GAL                6.29          5.61
    CETANE NO.                    96.1          36.1
    HEAT CONTENT, BTU/LB       20349.        20647.
    D86T DIST. CURVE, DEG. C:
        0%                       190.          -80.
        10%                      222.           26.
        20%                      229.           56.
        50%                      257.          103.
        90%                      313.          169.
        100%                     345.          195.
POWER PRODUCTION CALCULATIONS:

    GAS TURBINE POWER OUTPUT =           502.9 MW
    STEAM TURBINE POWER OUTPUT =         600.1 MW

    COAL GRINDING POWER CONSUMPTION =     24.3 MW
    ASU POWER CONSUMPTION =                0.0 MW
    PROCESS POWER CONSUMPTION =            4.4 MW
    SCOT PROCESS POWER CONSUMPTION =       0.2 MW
    RECTISOL POWER CONSUMPTION =           7.8 MW
    REFRIGERATION POWER CONSUMPTION =     32.2 MW
    CO2 COMPRESSION POWER CONSUMPTION =    4.3 MW
    FISCHER TROPSCH POWER CONSUMPTION =    7.9 MW
    PRODUCT UPGRADE POWER CONSUMPTION =    7.5 MW
    COOLING TOWER POWER CONSUMPTION =     11.4 MW
    POWER BLOCK POWER CONSUMPTION =        6.4 MW
    TRANSFORMER POWER LOSSES =             5.5 MW

    NET POWER GENERATED =                991.0 MW

EFFICIENCY CALCULATIONS:

    HEAT IN (HHV BASED):
        COAL HEAT CONTENT =             13270.0 MMBTU/HR
        H2 HEAT CONTENT =               13046.1 MMBTU/HR
        TOTAL HEAT IN =                 26316.1 MMBTU/HR

    HEAT OUT (HHV BASED):
        NET POWER OUTPUT =               3381.5 MMBTU/HR
        DIESEL + NAPHTHA HEAT CONTENT = 12370.4 MMBTU/HR
        TOTAL HEAT OUT =                15751.9 MMBTU/HR

         PLANT EFFICIENCY (HHV BASED) =         59.9 %
```

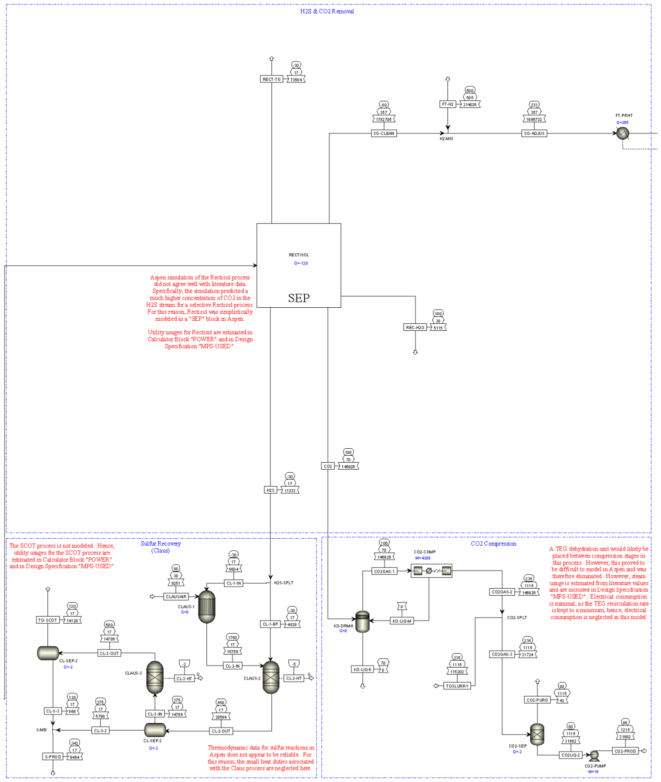
# Synfuel Production with Nuclear Integration - Powder River Basin Coal - No Light Product Recycle

Aspen Plus 12.1 Run:CTL With Nuclear Integration - No Light Recycle 8/15/2006 5:27:17 PM



# Synfuel Production with Nuclear Integration - Powder River Basin Coal - No Light Product Recycle

Aspen Plus 12.1 Run:CTL With Nuclear Integration - No Light Recycle 8/15/2006 5:27:17 PM



Aspen simulation of the Rectisol process did not agree well with literature data. Specifically, the simulation predicted a much higher concentration of CO2 in the H2S stream for a selective Rectisol process. For this reason, Rectisol was simplistically modeled as a "SEP" block in Aspen.

Utility usages for Rectisol are estimated in Calculator Block "POWER" and in Design Specification "MPS-USED".

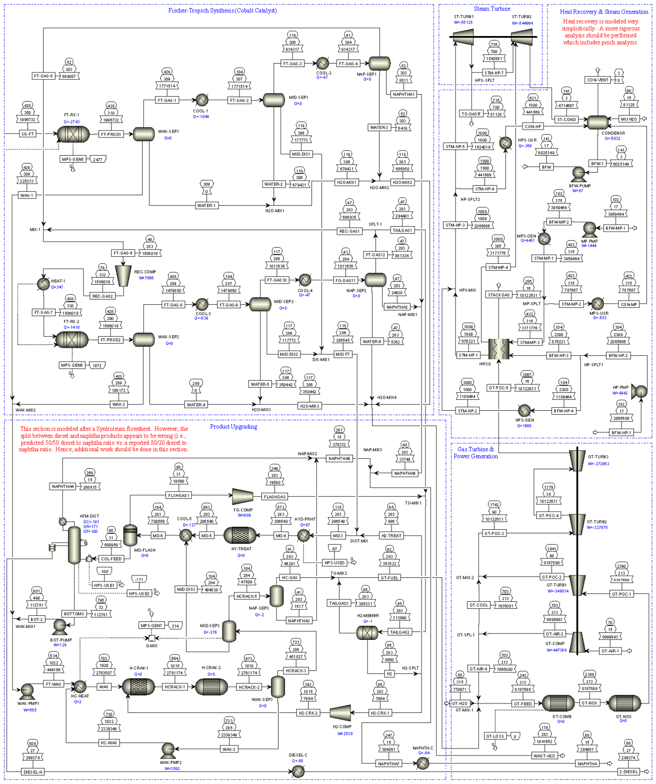
The SCOT process is not modeled. Hence, utility usages for the SCOT process are estimated in Calculator Block "POWER" and in Design Specification "MPS-USED"

Thermodynamic data for sulfur reactions in Aspen does not appear to be reliable. For this reason, the small heat duties associated with the Claus process are neglected here.

A TEG dehydration unit would likely be placed between compression stages in this process. However, this proved to be difficult to model in Aspen and was therefore eliminated. However, steam usage is estimated from literature values and are included in Design Specification "MPS-USED". Electrical consumption is minimal, as the TEG recirculation rate is kept to a minimum; hence, electrical consumption is neglected in this model.

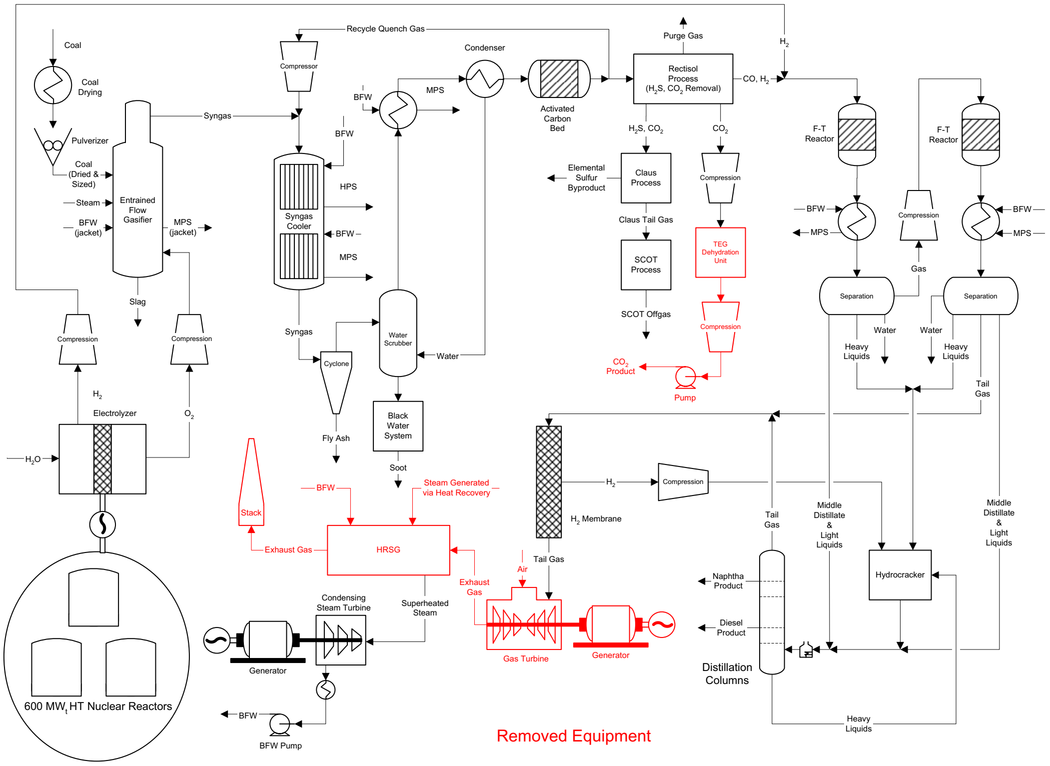
## Synfuel Production with Nuclear Integration - Powder River Basin Coal - No Light Product Recycle

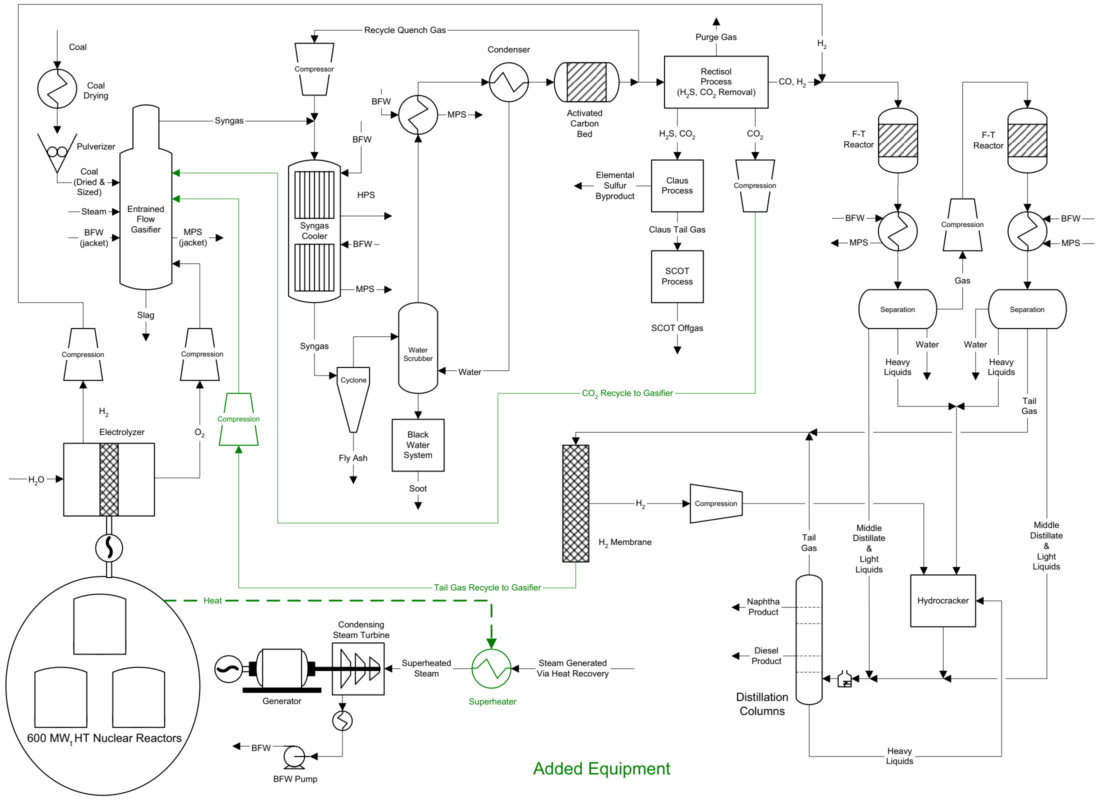
Aspen Plus 12.1 Run:CTL With Nuclear Integration - No Light Recycle 8/15/2006 5:27:17 PM

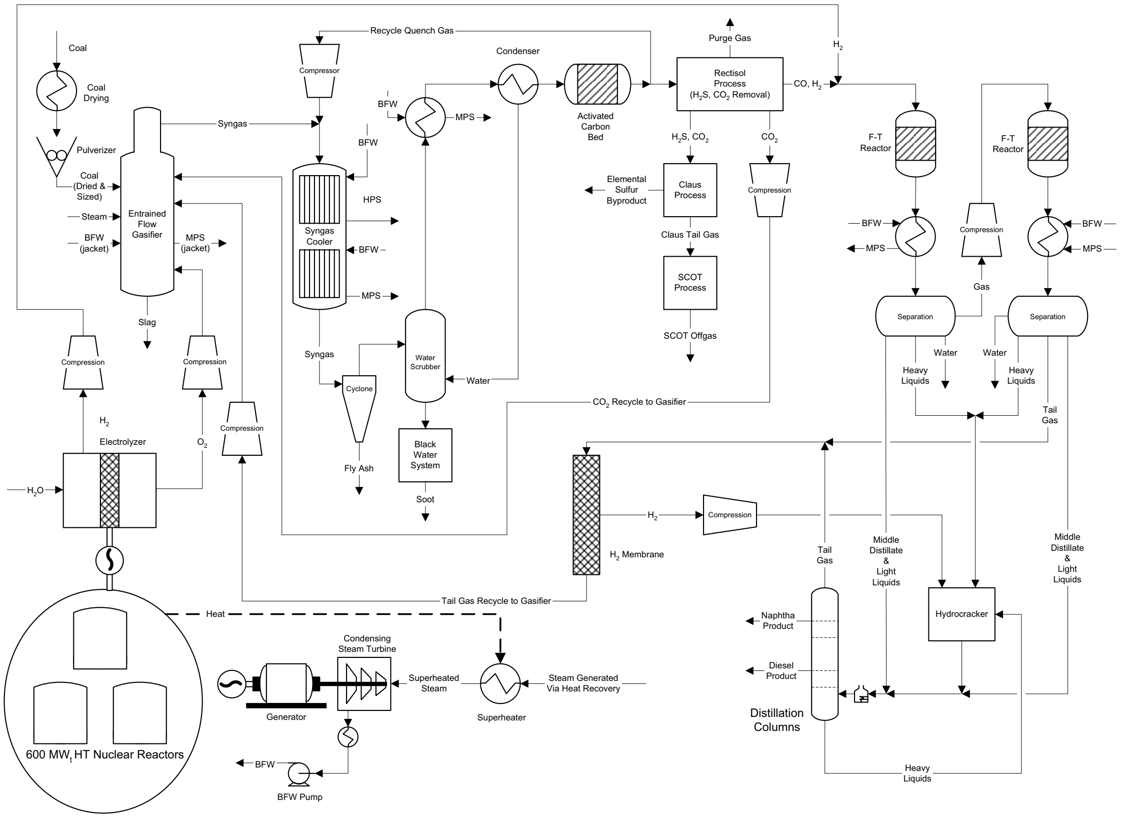


# Appendix C

# CTL Flowsheet with Nuclear Hydrogen Generation and $CO_2$ Recycle


Recycle Quench Gas
Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
MPS (jacket)
Slag
Syngas
Compressor
Condenser
BFW
MPS
BFW
Syngas Cooler
HPS
BFW
MPS
Activated Carbon Bed
Purge Gas
$H_2$
Rectisol Process ($H_2S$, $CO_2$ Removal)
CO, $H_2$
$H_2S$, $CO_2$
$CO_2$
Elemental Sulfur Byproduct
Claus Process
Claus Tail Gas
SCOT Process
SCOT Offgas
Compression
TEG Dehydration Unit
Compression
$CO_2$ Product
Pump
F-T Reactor
F-T Reactor
BFW
MPS
Compression
BFW
MPS
Gas
Separation
Separation
Water
Water
Heavy Liquids
Heavy Liquids
Tail Gas
Compression
Compression
Syngas
Cyclone
Water Scrubber
Water
$H_2$
Electrolyzer
$O_2$
$H_2O$
Fly Ash
Black Water System
Soot
Stack
BFW
Steam Generated via Heat Recovery
Exhaust Gas
HRSG
Exhaust Gas
Air
$H_2$
Compression
$H_2$ Membrane
Tail Gas
Tail Gas
Middle Distillate & Light Liquids
Middle Distillate & Light Liquids
Hydrocracker
Naphtha Product
Diesel Product
Distillation Columns
Heavy Liquids
Condensing Steam Turbine
Superheated Steam
Generator
Gas Turbine
Generator
600 $MW_t$ HT Nuclear Reactors
BFW
BFW Pump
Removed Equipment

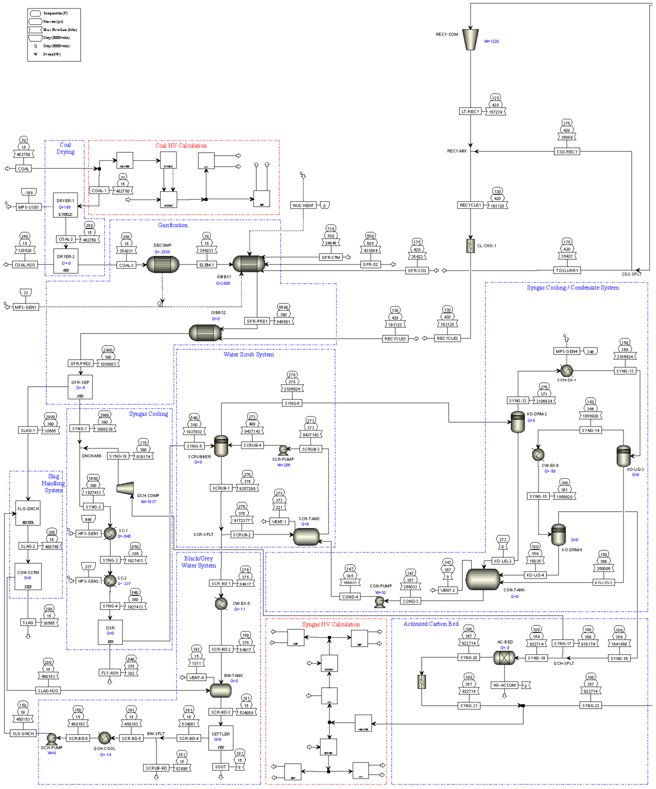

Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
MPS (jacket)
Slag
Syngas
Compressor
Recycle Quench Gas
Syngas Cooler
BFW
HPS
BFW
MPS
Syngas
Cyclone
Fly Ash
BFW
MPS
Condenser
Water Scrubber
Water
Black Water System
Soot
Activated Carbon Bed
Purge Gas
Rectisol Process ($H_2S$, $CO_2$ Removal)
$CO$, $H_2$
$H_2$
$H_2S$, $CO_2$
$CO_2$
Elemental Sulfur Byproduct
Claus Process
Claus Tail Gas
SCOT Process
SCOT Offgas
Compression
$CO_2$ Recycle to Gasifier
F-T Reactor
F-T Reactor
BFW
MPS
BFW
MPS
Compression
Gas
Separation
Separation
Water
Water
Heavy Liquids
Heavy Liquids
Tail Gas
$H_2$ Membrane
$H_2$
Compression
Tail Gas
Middle Distillate & Light Liquids
Middle Distillate & Light Liquids
Hydrocracker
Naphtha Product
Diesel Product
Distillation Columns
Heavy Liquids
Compression
Compression
$H_2$
$O_2$
Electrolyzer
$H_2O$
Compression
Tail Gas Recycle to Gasifier
Heat
600 $MW_t$ HT Nuclear Reactors
Condensing Steam Turbine
Generator
Superheated Steam
Superheater
Steam Generated Via Heat Recovery
BFW
BFW Pump
Added Equipment


Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
MPS (jacket)
Slag
Syngas
Compressor
Recycle Quench Gas
Syngas Cooler
HPS
BFW
MPS
BFW
Syngas
Cyclone
Fly Ash
BFW
MPS
Condenser
Water Scrubber
Water
Black Water System
Soot
Activated Carbon Bed
Purge Gas
Rectisol Process ($H_2S$, $CO_2$ Removal)
$H_2S$, $CO_2$
Claus Process
Elemental Sulfur Byproduct
Claus Tail Gas
SCOT Process
SCOT Offgas
$CO_2$
Compression
$CO_2$ Recycle to Gasifier
$H_2$
CO, $H_2$
F-T Reactor
F-T Reactor
BFW
MPS
BFW
MPS
Compression
Gas
Separation
Separation
Water
Water
Heavy Liquids
Heavy Liquids
Tail Gas
Compression
Compression
$H_2$
Electrolyzer
$O_2$
$H_2O$
Compression
$H_2$ Membrane
$H_2$
Compression
Tail Gas Recycle to Gasifier
Tail Gas
Middle Distillate & Light Liquids
Middle Distillate & Light Liquids
Hydrocracker
Naphtha Product
Diesel Product
Distillation Columns
Heavy Liquids
Heat
600 $MW_t$ HT Nuclear Reactors
Generator
Condensing Steam Turbine
Superheated Steam
Superheater
Steam Generated Via Heat Recovery
BFW
BFW Pump

```
CARBON BALANCE SUMMARY:

    PRB COAL FEED RATE =            5,793 TON/DAY

    % TOTAL CARBON TO LIQUID FUEL =   95.7 %
    % TOTAL CARBON TO SLAG & FLYASH =  0.2 %
    % TOTAL CARBON TO CO2   =          4.1 %

FUEL PROPERTIES:
                             #2 DIESEL       NAPHTHA
    PRODUCTION RATE, BBL/DAY   12159.        13877.
    MW                           201.6          92.4
    API GRAVITY                   52.8          78.5
    DENSITY, LB/GAL                6.29          5.61
    CETANE NO.                    96.1          36.1
    HEAT CONTENT, BTU/LB       20349.        20647.
    D86T DIST. CURVE, DEG. C:
        0%                      190.          -80.
        10%                     222.           26.
        20%                     229.           56.
        50%                     257.          103.
        90%                     313.          169.
        100%                    343.          195.

POWER PRODUCTION CALCULATIONS:
    STEAM TURBINE POWER OUTPUT =         247.8 MW

    COAL GRINDING POWER CONSUMPTION =      7.5 MW
    ASU POWER CONSUMPTION =                0.0 MW
    PROCESS POWER CONSUMPTION =            2.1 MW
    SCOT PROCESS POWER CONSUMPTION =       0.1 MW
    RECTISOL POWER CONSUMPTION =           4.0 MW
    REFRIGERATION POWER CONSUMPTION =     15.1 MW
    CO2 COMPRESSION POWER CONSUMPTION =    1.3 MW
    FISCHER TROPSCH POWER CONSUMPTION =    3.5 MW
    PRODUCT UPGRADE POWER CONSUMPTION =    3.4 MW
    COOLING TOWER POWER CONSUMPTION =      4.9 MW
    POWER BLOCK POWER CONSUMPTION =        2.3 MW
    TRANSFORMER POWER LOSSES =             1.2 MW

    NET POWER GENERATED =                202.4 MW

EFFICIENCY CALCULATIONS:

    HEAT IN (HHV BASED):
        COAL HEAT CONTENT =              4080.3 MMBTU/HR
        H2 HEAT CONTENT =                5110.2 MMBTU/HR
        NUCLEAR HEAT INPUT =              503.9 MMBTU/HR
        TOTAL HEAT IN =                  9694.4 MMBTU/HR

    HEAT OUT (HHV BASED):
        NET POWER OUTPUT =                690.5 MMBTU/HR
        DIESEL + NAPHTHA HEAT CONTENT =  5535.7 MMBTU/HR
        TOTAL HEAT OUT =                 6226.2 MMBTU/HR
         PLANT EFFICIENCY (HHV BASED) =         64.2 %
```

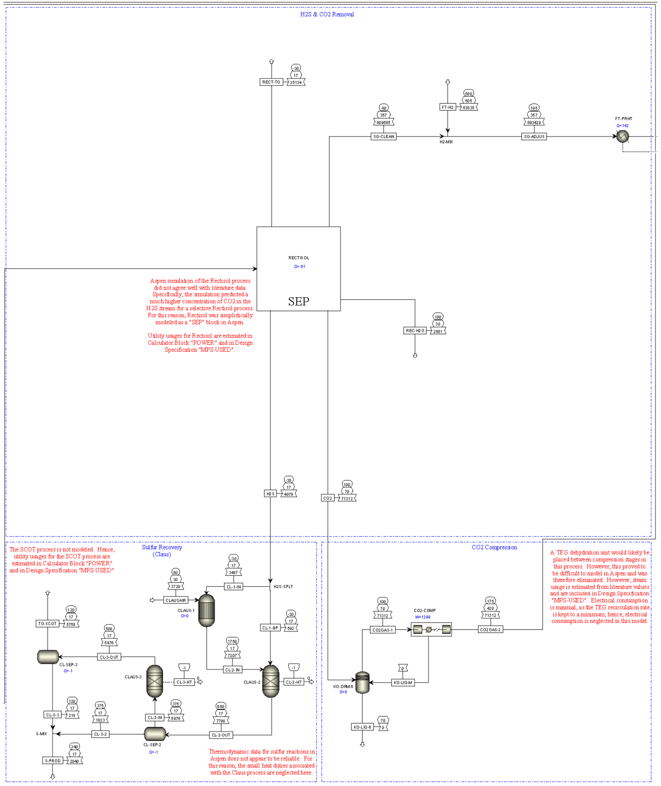
## Synfuel Production with Nuclear Integration - Powder River Basin Coal

Aspen Plus 12.1 Run:CTL With Nuclear Integration 8/15/2006 5:53:10 PM



## Synfuel Production with Nuclear Integration - Powder River Basin Coal

Aspen Plus 12.1 Run:CTL With Nuclear Integration 8/15/2006 5:53:10 PM



Aspen simulation of the Rectisol process did not agree well with literature data. Specifically, the simulation predicted a much higher concentration of CO2 in the H2S stream for a selective Rectisol process. For this reason, Rectisol was simplistically modeled as a "SEP" block in Aspen.

Utility usages for Rectisol are estimated in Calculator Block "POWER" and in Design Specification "MPS-USED".

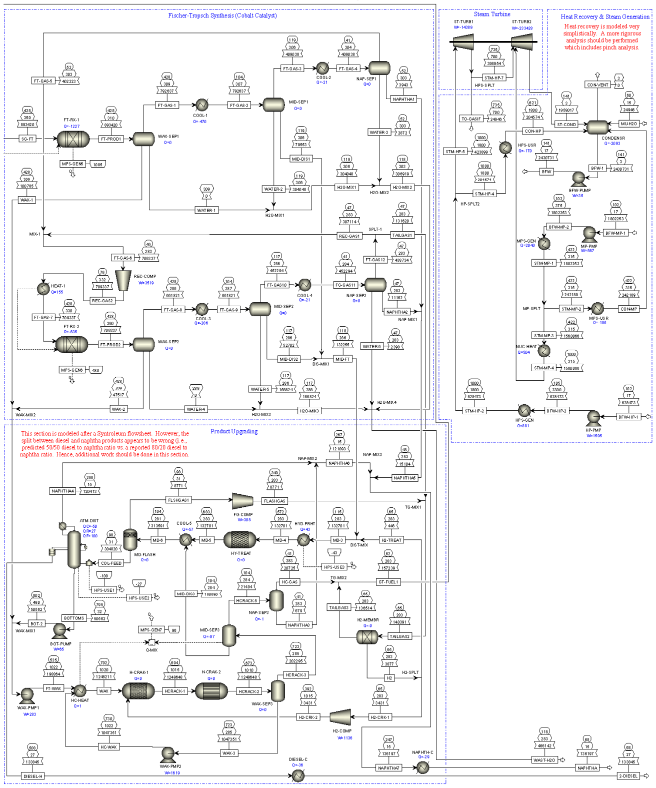
The SCOT process is not modeled. Hence, utility usages for the SCOT process are estimated in Calculator Block "POWER" and in Design Specification "MPS-USED"

Thermodynamic data for sulfur reactions in Aspen does not appear to be reliable. For this reason, the small heat duties associated with the Claus process are neglected here.

A TEG dehydration unit would likely be placed between compression stages in this process. However, this proved to be difficult to model in Aspen and was therefore eliminated. However, steam usage is estimated from literature values and are included in Design Specification "MPS-USED". Electrical consumption is minimal, as the TEG recirculation rate is kept to a minimum; hence, electrical consumption is neglected in this model.

# Synfuel Production with Nuclear Integration - Powder River Basin Coal

Aspen Plus 12.1 Run: CTL With Nuclear Integration 8/15/2006 5:53:11 PM



# Appendix D

# Capital Cost Estimate