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Gasification of Oil Refinery Waste for Power and Hydrogen Production

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Abstract

This paper describes the design of an Integrated Gasification Combined Cycle (IGCC) power plant with hydrogen production capabilities with petroleum coke as a feedstock. The economic analysis of the operating costs, unit costs, payback period and possible plant locations are examined. An IGCC power plant was successfully simulated using AspenPlus Simulation software. The power plant uses 175,000 kg/hr of Petroleum Coke to produce 437 MW of power and 12,850kg/hr of hydrogen at 25.6% efficiency. This is based on a 50% clean syngas swing cut. A preliminary costing assessment indicates that a capital investment of 1.07 billion dollars would be required for a plant of this size. Model and costing data indicates that at 3.45¢/kW·h for electricity production and $1.38/kg of hydrogen produced, IGCC is a viable alternative to other technologies. In current market conditions for hydrogen and electricity costs, it is more beneficial to produce a split of 50:50 hydrogen and power. This operating strategy leads to larger profits when compared to 100% power production, with an expected payback period of 6.9 years. Ideal locations for a petcoke IGCC require a large supply of fresh water and availability of petcoke from multiple refineries. Locations such as Southeastern Louisiana or Ufa, Russia meet these conditions.

Keywords

refinery waste, syngas treatment, shift reaction, economic analysis, ASPEN simulation

1. Introduction

Petroleum coke (petcoke) is primarily a carbonaceous material that is a by-product during the coking process in an oil refinery. Petcoke can be used in a gasification process to create synthesis gas (syngas). Syngas is a very useful mixture of carbon monoxide and hydrogen gas. This gas can be combusted in turbines to create power, or undergo further processing to remaining CO into hydrogen. The low price of petcoke compared to coal indicates that petcoke has the potential to replace/retire existing coal power plants.

Current research efforts are underway to develop a full scale petcoke fed power plant. Pilot plants such as the "Tampa Electric Polk Power Power Station" are operating with an integrated gasification combined cycle (IGCC) power plant design, producing 250MW net power (Tampa Electric Company 2002). The use of an IGCC can result in efficient power generation with inherently far lower pollution than current coal fired plants. Additionally, this process is also ideal for a carbon capture system which can theoretically eliminate carbon emissions.

2. Background

The main goal of this paper was to design and evaluate a gasification process that would generate power and hydrogen from a refinery waste stream. An Aspen model was created for simulations assumed a feed rate of 4,500 tonne/day petcoke, with scenarios of 100% power production and 50/50 power/hydrogen production. The second goal was to perform preliminary costing estimates for capital and operation costs. The third goal was to develop operating strategies to maximize profit for the plant. This involved a comparison of the profit margins of hydrogen and electricity.

Gasification of petcoke is a complex process involving the carbon-based petcoke, water and oxygen. The reactions to produce syngas are carried out in a gasifier unit at high temperatures and pressures. Operating conditions for these units are specific to the feed and gasifier type. The four major types of gasification technology currently used are: Sasol-Lurgi Dry Ash, GE (Teaxco), Shell and ConocoPhillips E-gas (Dow) (Maurstad 2005). Due to popularity of the Texaco entrained flow slagging type, this gasification technology was chosen for the modelling process.

3. Model Development

The Aspen Simulation model was based on a previous simulation for a Master's Thesis (Ordorica-Garcia 2003).

3.1 Gasification Pre-Treatment

Petroleum coke is not defined in the component libraries of Aspen but the user may specify a nonconventional solid component and specifically assign it elemental compositions. Suncor petroleum coke from the Alberta tar sands was used in this study. The Specifications are in Table 1.

Table 1: Properties of Suncor Petroleum Coke (Furimsky 1998)

Proximate Analysis (wt. %)		**Ultimate Analysis (wt. %)**	
Moisture	0	Ash	3.6
Fixed Carbon	84	Carbon	84.2
Volatile Matter	12.4	Hydrogen	3.8
Ash	3.6	Nitrogen	1.5
		Chlorine	0
		Sulphur	5.8
		Oxygen	1.1

The petroleum coke undergoes pre-treatment before it can be gasified. The coke is mixed with water. The amount of water is determined by an inline FORTRAN calculator block and is added in sufficient quantity to yield a mixture of 65% solids. This slurry is then pumped (efficiency: 0.92) to a discharge pressure of 44.8 bars to a train of deconstruction calculators that are ostensibly three reactor blocks. The first block, DECOMP (RYIELD block) breaks the slurry down into its elemental constituents, as well as ash (15° C, 1.01 bar, components = H_2O, C, Ash, H_2, N_2, Cl_2, S, O_2). This is done by another inline FORTRAN calculator block that breaks down the slurry based on an ultimate analysis of the petroleum coke. The SOOT and SLAG blocks (RSTOIC blocks) simulate the production of soot and slag as would happen in an actual Gasifier. These reactions cannot be put into the Gasifier block in the simulation because an RGIBBS block cannot produce nonconventional solids. The reaction present in the SOOT block is (15° C, 0 bar, C fractional conversion = 0.0036):

$$0.916\text{ASH} + 0.0069935\text{C} \rightarrow \text{SOOT}$$

The slurry is then sent to the SLAG block which converts the remaining ash into slag. The reaction present in the SLAG block is (15° C, 0 bar, Ash fractional conversion = 1):

$$0.95\text{ASH} + 0.004162\text{C} \rightarrow \text{SLAG}$$

The above reactions were reported in Ordorica-Garcia (2003) and based on actual Texaco Gasifier performance. The slurry is then mixed with oxygen from the Air Separation Unit. The oxygen flow rate is determined by a calculator block that ensures there is an oxygen-to-coke mass ratio of 0.969 (Figure 1).

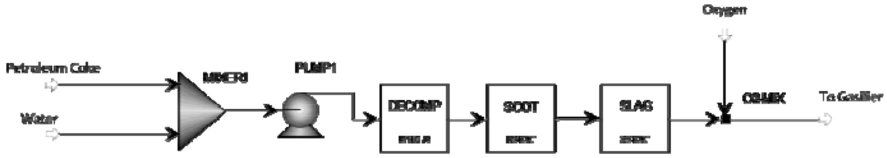

Figure 1: Petroleum Coke Pre-Treatment

3.2 Gasification

The prepared slurry is then sent to the Gasifier (1371° C, 44.8 bar). The gasifier is represented by a RGIBBS reactor block. The chemical equilibrium is restricted to the following reactions in Table 2:

Table 2: Reactions Occurring in the Gasifier

$C + 2H_2 \rightarrow CH_4$

$C + H_2O \rightarrow CO + H_2$

$CO + H_2O \rightarrow CO_2 + H_2$

$2CO + O_2 \rightarrow 2CO_2$

$S + H_2 \rightarrow H_2S$

$N_2 + 3\ H_2 \rightarrow 2NH_3$

$CO + H_2S \rightarrow COS + H_2$

$H_2 + Cl_2 \rightarrow 2HCl$

These reactions were based on those reported in Ordorica-Garcia (2003). A multiphase equilibrium restricted by temperature was used. The raw syngas then moves through a radiant cooler then through a SEP block that simulates the removal of the molten slag (split fraction NC components in slag = 1). In reality, this would leave the gasifier through the bottom chamber. The radiant cooler and convective coolers are supposed to simulate the extraction of heat used in the Heat Recovery Steam Generator to create high pressure steam for the steam cycle (cooler 1: 815° C, 41.2 bar; cooler 2: 343° C, 41.6 bar). In practice, this would involve more complex heat exchanging operations. The model assumes all of the exothermic heat produced from the gasification reaction is transferred to the steam. The slag-free syngas is passed through a solid scrubber (39.44 bar, heat duty = 0) where the remaining solids and ash are scrubbed out by water (Figure 2).

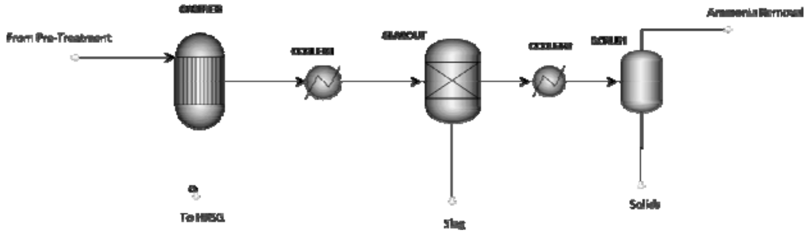

Figure 2: Petroleum Coke Gasification.

3.3 Syngas Treatment and Shift Reactions

The particulate free syngas then moves to an ammonia separation unit, essentially a wet scrubber (split fraction of NH_3 in stream $NH_3 = 1$). In reality, this is a complex process that would require several unit operations. Due to this complexity, the separation has been modelled using a SEP block that effectively removes all of the ammonia, hydrogen chloride, and chlorine gas. The main purpose of this block is to remove the ammonia so that it reduces the amount of fuel-bound nitrogen in the fuel gas, thus reducing the amount of NO_x that would form in the combustors. The syngas then passes through an Aspen Plus *ClChng*. The *ClChng* is an Aspen operation that changes the class of the stream, in this case from a nonconventional class to a conventional class. This change will simplify subsequent unit operation specifications. The syngas then passes through a heater (300°C, 39.4 bar) that preheats the syngas before being mixed with steam and sent to the Shift Reactors. The Shift reactors are implemented to promote the production of hydrogen gas from CO and H_2O through the Water-Gas Shift reaction. In the first reactor, the following reactions take place (350°C, 28.6 bar, CO fractional conversion = 0.9, COS fractional conversion = 0.95):

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

$$COS + H_2O \rightarrow CO_2 + H_2S$$

The majority of the carbonyl sulphide is converted in this reaction as well as most of the carbon monoxide. Before entering the second reactor the gas is cooled (200°C, 28.6 bar). This reactor converts the remaining carbon monoxide into carbon dioxide (Figure 3) and hydrogen gas (220°C, 28.6 bar, CO fractional conversion = 0.9).

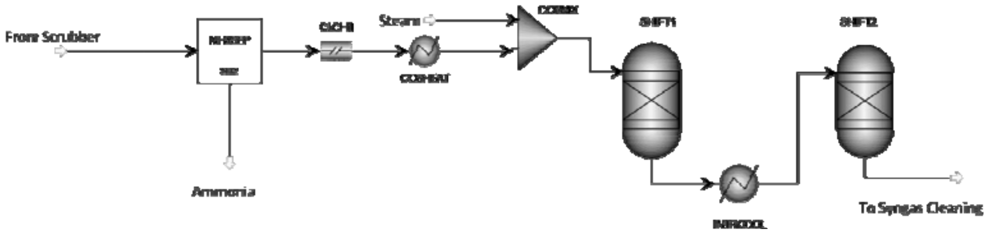

Figure 3: Syngas Treatment and Shift Reactions.

3.4 H_2S and CO_2 removal

The syngas must now pass through complex cleaning processes to produce a syngas that is acceptable for combustion and hydrogen production.The syngas is cooled before entering H_2S and CO_2 removal (40° C, 28.6 bar). These two processes are very complex and involved and were out of the scope of this investigation. To model these processes, SEP units were used with an absorbing liquid consisting of the proprietary Selexol, to remove 99.97% of the acid gas and 90% of the carbon dioxide (split fractions in acid gas = H_2 (0.000126), CO (0.000607), CO_2 (0.183), H_2S (0.9997), COS (0.45) and split fractions in CO_2 rich = H_2 (0.02), CO (0.03), CO2 (0.915),O_2 (0.04), N_2 (0.02), CH_4 (0.025), Ar (0.025)). The cleaned syngas is then sent to a fuel splitter that splits 4% of the gas to fuel for the acid plant. In addition, the remaining syngas is split between hydrogen production and electricity production (split fractions = syngas to SCOT (0.04), syngas to turbines (0.48), syngas to H_2 refinement (0.48)) which is shown in Figure 4.

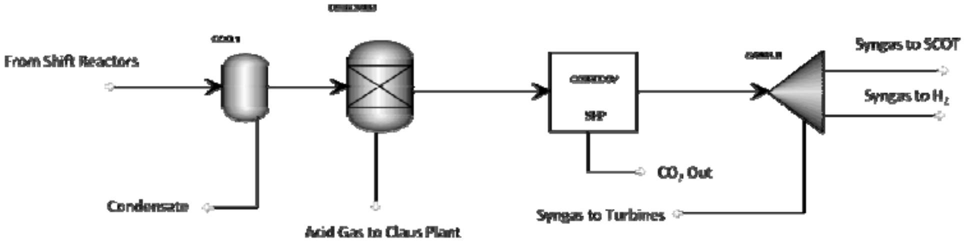

Figure 4: H_2S and CO_2 removal.

3.5 Hydrogen Refinement

The fuel gas sent to hydrogen refinement is first sent to a methanator to convert the remaining carbon monoxide in the fuel gas to methane. The reaction that takes place in the unit is (50°C, 27.6 bar,CO fractional conversion = 1):

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$

This will ultimately improve the quality of the final project. The product gas is then sent to a Pressure Swing Adsorption unit that will remove any impurities from the product gas (split fraction in hydrogen = 0.95). This unit can be operated to produce a saleable product ranging from 98 to 99.9% hydrogen (Figure 5).

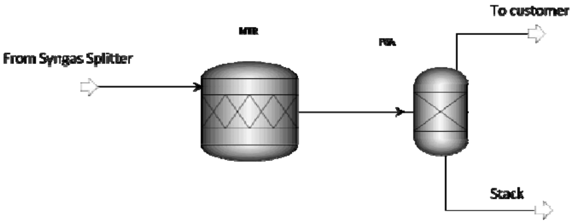

Figure 5: Hydrogen Refinement.

3.6 H_2S Treatment

To determine the amount of elemental sulphur that can be produced from the acid gas, a SCOT/CLAUS plant was implemented. The acid gas that is removed from the Selexol plant is first mixed with compressed atmospheric air (isentropic, discharge pressure = 1.6 bar, isentropic efficiency = 0.89), the mixture enters the Claus Furnace. In the furnace, approximately a third of the entering H_2S is converted to SO_2 by the following reaction (309° C, 0 bar, H_2S fractional conversion = 0.333):

$$2H_2S + 3O_2 \rightarrow 2H_2O + 2SO_2$$

The product gas then moves to the CLAUS reactor. The Claus reactor then converts the SO_2 into elemental sulphur by the following reaction (140° C, 0 bar , H_2S fractional conversion = 0.97):

$$2H_2S + SO_2 \rightarrow 2H_2O + 3S$$

The product stream is then sent to a separator, where the solid elemental sulphur is removed (solids split fractions sulphur = 1). The remaining gas phase components are sent through *ClChng* block before proceeding to the SCOT reactor for tail gas treatment. Before entering the SCOT reactor, the separated gas stream is first mix with a small amount of clean syngas. The SCOT unit then converts any SO_2 and COS into H_2S by the following reactions (149° C, 1.48 bar, SO_2 fractional conversion = 1, COS fractional conversion = 1):

$$SO_2 + 3H_2 \rightarrow 2H_2O + H_2S$$

$$COS + H_2 \rightarrow H_2S + CO$$

The product stream then goes to a separation unit that separates the acid gas from all of the other components (split fractions in off-gas = H_2 (1), CO (1), CO_2 (1), O_2 (1), N_2 (1), H_2O (1), CH_4 (1), AR (1)). The acid gas is recycled back into the CLAUS section by being mixed with the incoming acid gas from the Selexol plant (Figure 6).

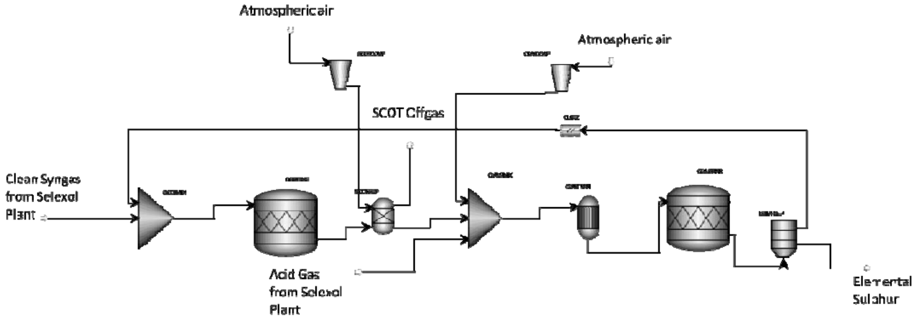

Figure 6: Acid gas treatment plant

3.7 Fuel Gas Pre-treatment

The fuel gas must be pre-treated before it can be used for power generation. The clean syngas is first mixed with steam in a standard MIXER unit. This moistened air is first reheated (28.5 bar) before it is subsequently mixed with nitrogen from the Air Separation Unit in another MIXER unit. The addition of these two streams will reduce flame temperatures in the combustor and ultimately lower thermal NO_x formation. The amount of steam and nitrogen added was based on recommendations from Ordorica-Garcia (2003). Just prior to the combustion reactions, the moist treated syngas is combined with outside air. The air is assumed to be at 15°C and 1 atm and is supplied at the rate recommended by Ordorica-Garcia (2003). The air passes through a COMPR block that compresses the air to a discharge pressure of 15.7 atm. The compressor operates isentropically, with an isentropic efficiency of 0.875 (Figure 7).

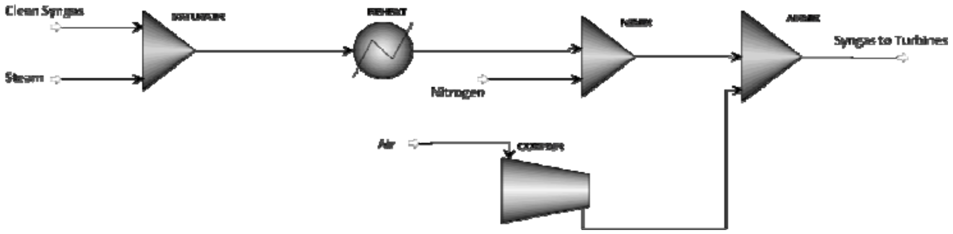

Figure 7: Fuel Gas Pre-Treatment

3.8 Combustion and Power Generation

The syngas is then sent to a FSPLIT block where the syngas is split equally between the two combustor feeds. The syngas is then combusted in the two combustor blocks. The combustor blocks are RGIBBS reactor blocks and are used for the same reasons as for the modelling of the gasifier. The combustors are operated by an adiabatic temperature restricted equilibrium (heat duty = 0, pressure = -0.008 bar, multiphase equilibrium, temperature restricted). The following reactions are present in the combustors:

$$2CO + O_2 \leftrightarrow 2CO_2$$
$$2H_2 + O_2 \leftrightarrow 2H_2O$$
$$CH_4 + 2O_2 \leftrightarrow CO_2 + 2H_2O$$
$$2H_2S + 3O_2 \leftrightarrow 2SO_2 + 2H_2O$$
$$4NH_3 + 3O_2 \leftrightarrow 2N_2 + 6H_2O$$

These hot product gases are then expanded in two turbines. The turbines operate isentropically, with an isentropic efficiency of 0.885 (discharge pressure = 1.02 bar). In practice, to compress the combustion air, the turbine and air compressor operate on the same shaft. To compensate for this loss of power, a mechanical efficiency of 52% was imposed on each turbine. The hot exhaust gases are combined before being sent to a heat exchanger that reduces the temperature of the gases before they are released out of the stack (105° C, 1.01 bar). The heat recovered is combined with the heat from the convective and radiant coolers from the gasifiers and sent to the Heat Recovery Steam generator. The heat recovery would be rather more complex in practice and would involve many heat exchangers, but for simulation purposes, a simple HEATER block was used (Figure 8).

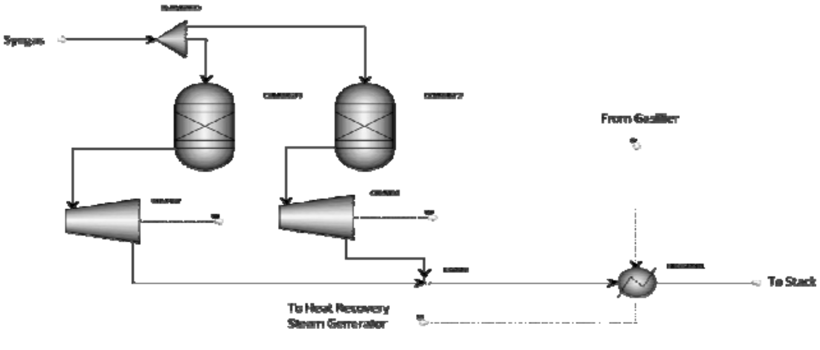

Figure 8: Fuel Combustion and Power Generation.

3.9 Heat Recovery Steam Generator

The heat recovery steam generator is a complex series of heat exchangers that essential takes liquid water and heats it with the heat recovered from the exhaust gases of the gas turbines to produce superheated steam that is used to power three turbines. There are three turbine pressures produced, low, intermediate, and high. The high pressure (HP) steam is at 124 bars, the intermediate pressure (IP) is at 22.2 bar, and the low pressure (LP) turbine is at 3.79 bars. The condensate is split 6% to the LP steam, 9% to the IP steam, and 85% to the HP steam, this was done based on recommendations from Ordorica-Garcia (2003). The steam exiting the high pressure turbine is reused in the intermediate pressure turbine, and the steam exiting the intermediate pressure turbine is then reused in the low pressure turbine. The maximum power potential can be recovered when the steam is recycled in this way (Figure 9).

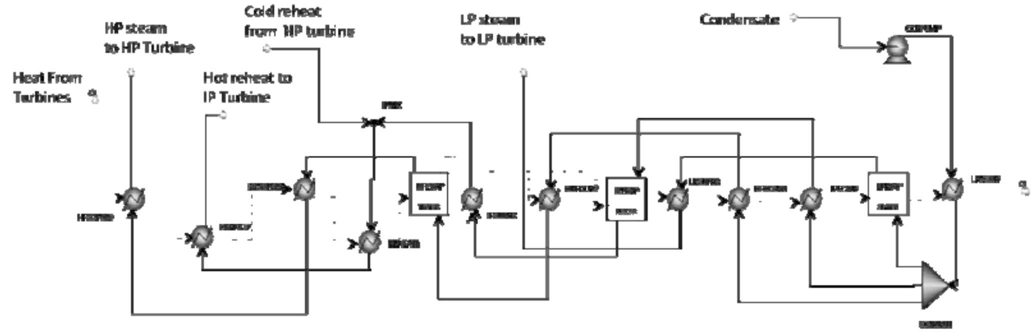

Figure 9: Heat Recovery Steam Generator.

3.10 Steam Turbines

After leaving the heat recovery steam generator the different steam streams enter their respective turbines. High pressure steam enters the high pressure turbine at 124 bar and leaves at a discharge pressure of 22.95 bars. The intermediate pressure turbine takes in steam at 22.95 bar and discharges it at 3.79 bars. The low pressure turbine takes in steam at 3.79 bars and discharges it at 38 mmHg. All turbines operate isentropically with an isentropic efficiency of 0.92 (Figure 10).

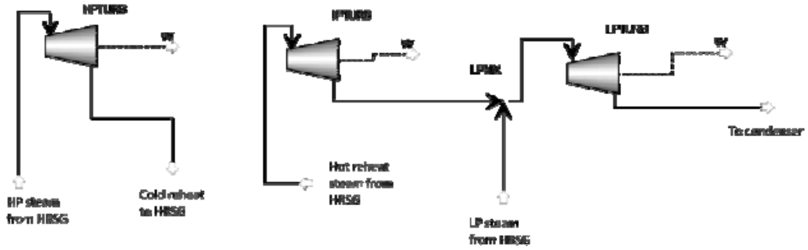

Figure 10: Steam Turbines.

3.11 Simulation Results

3.11.1 Mass Balance

The entire plant mass balance can be seen in Table 3. Selected stream compositions can be seen in Table 4.

Table 3: Overall Mass Balance (Based on a 50% swing cut)

Input	kg/hr	Output	kg/hr
Petroleum Coke	175,000	Slag	6,027
Oxygen	169,656	Soot	6,167
Nitrogen	529,460	Waste water	199,099
Air (GT)	2,661,100	Offgas	103,180
Air (claus)	5,007	Sulphur	2,321
Water (slurry)	93,751	Stackout	3,276,450
Water (steam)		Hydrogen Sold	12,850
Cycle	689,254	PSA Refuse	25,956
Gas Conditioning	47,084	Carbon Dioxide Rich	388,097
Shift Reaction	348,069	Steam Cycle	689,254
Total	4,718,381		4,709,401

Table 4: Selected Stream Compositions

Stream	Clean syngas mole %	CO2 Rich mole %	Acid Gas mole %	SCOT off-gas mole %	Fuel gas mole %	Flue gas mole %
H2	91.3	3.2	732 ppm	19.1	23.6	Trace
CO	0.6	341 ppm	25 ppm	0.2	0.2	Trace

CO2	5.2	96.7	85.8	73.1	1.4	0.4
O2	Trace	Trace	-	32 ppm	Trace	13.5
N2	1.5	514 ppm	-	4.8	65.5	77.2
H_2O	0.4	0	-	2.5	9.1	8.1
H_2S	7 ppm	-	14.1	-	2 ppm	Trace
Ar, CH_4, others	0.9	398 ppm	242 ppm	0.2	0.2	0.8

3.11.2 Power Balance

The total power generated and consumed can be seen in Table 5.

Table 5: Overall Power Balance

Input	Petroleum Coke	1709	MW_{th}
Outputs	Gas Turbines	269	MW_e
	Steam Turbine	293	MW_e
Gross Output		561	MW_e
Plant Consumption		124	MW_e
Net Output		437	MW_e
Efficiency (HHV)		25.6	%

4.0 Economic Analysis

4.1 Capital Costs

The calculations for the direct capital costing section were based on a combination of theoretical formulas from literature and scaled-up values from previously established plants (results shown in Table 6). The theoretical calculations were based on the report written by Frey and Rubin (1990), for the US Department of Energy. Process scaling was used to calculate the major process areas for which theoretical calculations were not directly available. Scaling factors were determined using ratios comparing input feeds for the different units. For instance, the input air feed rate from the process simulation was compared to the air feed rate of the Tampa Electric plant to determine the scaling factor for the air separation unit.

Table 6: Direct capital costs for major process areas of IGCC power plant

Major Process Area	Calculation Basis	Capital Cost (2009 USD)
Air Separation Unit (ASU)	Tampa Electric Polk Plant	104,450,000
Petcoke Gasification	Tampa Electric Polk Plant	290,175,000
Gas Cooling and Scrubbing	Frey and Rubin (1990)	18,117,000
Selexol Plant	Frey and Rubin (1990)	48,803,000
Claus Plant	Frey and Rubin (1990)	2,785,000
Boiler Feedwater System	Frey and Rubin (1990)	8,250,000
Process Condensate System	Frey and Rubin (1990)	18,306,000
Power Generation	Tampa Electric Polk Plant	66,815,000
PSA Unit	NREL Biomass Plant	13,302,000
General Facilities	Frey and Rubin (1990)	100,765,000
Total Direct Capital (TDC)		**671,768,000**

For indirect capital costing calculations (results shown in Table 7), textbook values provided by Peters et al (2003) were used. Summing up all capital costs, the total capital investment (TCI) was determined to be \$1.07 billion dollars (calculated in 2009 US dollars). Considering an annual net electricity production of 3,460 GW·h, the capital costs can be represented as 2,448 USD/kW.

Table 7: Total capital costs including indirect and working capital values.

Project Requirement	Cost (2009 USD)	Assumptions
Engineering and Supervision	92,885,000	14% of total direct capital costs
Construction and Contractor	94,540,000	10% of the fixed capital investment
Contingency	94,540,000	10% of the fixed capital investment

Total Indirect Costs (TIC)	281,965,000	
Total Direct Costs (TDC)	671,768,000	
Fixed Capital Investment (FCI)	953,733,000	
Working Capital	116,850,000	10% of the total capital investment
Total Capital Investments (TCI)	1,070,583,000	
Capital Investment (USD/kWe net)	2,448	annual net electricity production of 3,460 GW·h

4.2 Operating Costs

A number of assumptions were made when calculating the operating costs of the plant (results shown in Table 8). For instance, for the calculation of operating labour, an hourly wage of $30/hr for each employee was assumed, based on US Ministry of Labor data for power plant production occupations (US MOL 2009). It was also assumed that 14 employees would be required to operate the plant, and that the plant would be in production 24 hours/day for 330 days out of the year, allowing for a one month outage period. These assumptions on plant requirements were made based on a master's thesis paper prepared for the University of Waterloo (Ordorica-Garcia 2003).

Table 8: Total operating costs and cost of electricity/hydrogen for IGCC.

Operations Requirement	**Cost (2009 USD/yr)**
Operating Labour (OL)	3,325,000
Supervisory and Clerical Labour	500,000
Administrative Costs	500,000
Plant Overhead	1,665,000
Maintenance and Repairs	18,410,000
Total Fixed Operating Cost	24,400,000
Fuel	66,305,000
Catalysts and Chemicals	2,760,000
Total Variable Operating Costs	69,065,000
Amortized Capital Costs	22,115,000
Total Operating Costs	115,580,000
Cost of Electricity (US¢/kW·h)	3.10
Cost of Hydrogen Production (USD/kg)	1.24

Supervisory and administrative labour costs were calculated as being 15% of operating labour, while plant overhead was represented by 50% of OL (Peters et al. 2003). Maintenance and repairs costs were calculated using a list of factors applied to each unit that would require maintenance. These factors were taken from a master's thesis paper prepared for North Carolina State University (Akunuri 1999). Fuel costs were calculated using pet coke feed rates, which were taken from the simulation. Lower heating values and spot prices for petroleum coke, which were attained from literature (IEA 2010), were also used in the fuel cost calculations. The cost for catalysts and chemicals were assumed to be equivalent to 15% of the maintenance and repairs costs (Peters et al. 2003). A straight-line depreciation calculation was used to determine amortized capital costs for the plant. A 30-year operating life for the plant was assumed for the purposes of this calculation.

The final results for the cost of electricity and hydrogen production are displayed at the bottom of Table 8. With a total annual operating cost of $115,580,000, the cost of electricity for the plant is 3.10 US¢/kW·h. This value is based on the maximum electricity production (no hydrogen is produced). An increase in this cost could be expected if hydrogen was also being produced at the plant. This value of 3.10 US¢/kW·h of electricity produced, is relatively inexpensive compared to a previously reported value of 4.47 US¢/kW·h for an IGCC plant without CO_2 capture (Ordorica-Garcia 2003). This is most likely due to the lower fuel costs associated with pet coke as compared to coal. The cost for hydrogen production was determined to be $1.24/kg, based on the amount of hydrogen produced per year and the annual operating costs. The production rates of hydrogen are based on the case where there is a 50:50 split of hydrogen production to electricity production. Other values reported for the cost of hydrogen production from the gasification of coal have been lower than the cost yielded from this simulation. A cost of $1.10/kg was reported by the US Department of Energy (DOE), for a theoretical coal to hydrogen plant (Padro 1999). This cost could easily be reached, if hydrogen production were increased beyond 50%.

4.3 Operating Strategies

To determine the maximum level of profits that can be made from this endeavour, the case of 100% power production will be compared to the case of a 50:50 split between hydrogen and power production. For the 100% power production case, the profit margin is calculated by comparing the cost of producing the electricity to the price that the electricity could be sold back to the refinery or other interested parties.

For December 2009, the US DOE reports the average price paid for power for industrial purposes was 6.52 US¢/kW·h (US DOE, 2010). This represents a profit of 3.42 US¢/kW·h, which works out to a total profit of 127,580,000 USD/yr. At this rate, the payback period for the plant would be approximately 8.3 years. If a 50:50 split of hydrogen to power production were used, then only 50% of the value presented above could be expected to come from power production. This value would therefore be 63,790,000 USD/yr without the profit from hydrogen production factored in. The cost for hydrogen, purchased on an industrial scale, is around 2.20-3.10 USD/kg (US DOE, 2002). Assuming the low range for the market cost of hydrogen, the difference in prices represents a profit of 0.96 USD/kg. Based on the hydrogen flowrates yielded in the simulation, a profit of 89,685,000 USD/yr can be expected. With both of these values combined, the total profit for this particular case would be 153,475,000 USD/yr. With the increased amount of profit for this case, the payback period is effectively lowered to 6.9 years. Please see Table 9 for a complete summary of results based on the two operating strategies.

Table 9: Economic indicators for both operating strategies.

	100% Power Production	**50:50 Hydrogen to Power Production**
Annual Profits (USD/yr)	127,580,000	153,475,000
Payback Period	8.3	6.9

4.4 Possible Plant Locations

A few geographic locations lend themselves to the economical operation of an IGCC facility based on the plant requirements, which are abundant water supply and availability of petroleum coke as a feedstock.

Southeastern Louisiana (USA) is possibly the best location for an IGCC power plant and this region has a high density of oil refineries in quite a small area. In total, there are 11 operating oil refineries along an approximately 150-mile stretch of the Mississippi River. The fact that the Mississippi River also runs through this area is a major advantage, as it can more than provide enough fresh water to the plant. Ufa is a city in Western Russia, is also a good potential location for a pet coke-powered IGCC power plant. The city itself is encompassed by three rivers, the Ufa, the Belaya, and the Dyoma and the city is also home to three oil refineries owned by the Bashneft oil company.

5.0 Conclusions

An IGCC process was successfully designed that uses 175,000 kg/hr of Petroleum Coke to produce 437 MW of power and 12,850kg/hr of hydrogen at 25.6% efficiency. A preliminary costing assessment indicates that a capital investment of 1.07 billion dollars would be required for a plant of this size. Model and costing data indicates that at 3.45¢/kW·h for electricity production and $1.38/kg of hydrogen produced, IGCC is a viable alternative to other technologies. In current market conditions for hydrogen and electricity costs, it is more beneficial to produce a split of 50:50 hydrogen and power. This operating strategy leads to larger profits when compared to 100% power production, with an expected payback period of 6.9 years. Ideal locations for a petcoke IGCC require a large supply of fresh water and availability of petcoke from multiple refineries. Locations such as Southeastern Louisiana or Ufa, Russia meet these conditions.

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Biography

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