



Techno-economic evaluation of coal-to-liquids (CTL) plants with carbon capture and sequestration

Hari Chandan Mantripragada^{a,*}, Edward S. Rubin^b

^a Department of Engineering and Public Policy, 129 Baker Hall, 5000 Forbes Avenue, Carnegie Mellon University, Pittsburgh, PA 15213, USA

^b Department of Engineering and Public Policy, BH 128A, 5000 Forbes Avenue, Carnegie Mellon University, Pittsburgh, PA 15213, USA

ARTICLE INFO

Article history:

Received 8 July 2010

Accepted 14 February 2011

Available online 5 March 2011

Keywords:

Coal-to-liquids

Co-production

Carbon capture and storage

ABSTRACT

Coal-to-liquids (CTL) processes that generate synthetic liquid fuels from coal are of increasing interest in light of the substantial rise in world oil prices in recent years. A major concern, however, is the large emissions of CO₂ from the process, which would add to the burden of atmospheric greenhouse gases. To assess the options, impacts and costs of controlling CO₂ emissions from a CTL plant, a comprehensive techno-economic assessment model of CTL plants has been developed, capable of incorporating technology options for carbon capture and storage (CCS). The model was used to study the performance and cost of a liquids-only plant as well as a co-production plant, which produces both liquids and electricity. The effect of uncertainty and variability of key parameters on the cost of liquids production was quantified, as were the effects of alternative carbon constraints such as choice of CCS technology and the effective price (or tax) on CO₂ emissions imposed by a climate regulatory policy. The efficiency and CO₂ emissions from a co-production plant also were compared to the separate production of liquid fuels and electricity. The results for a 50,000 barrels/day case study plant are presented.

© 2011 Elsevier Ltd. All rights reserved.

1. Introduction

The world is facing the twin problems of depleting fossil fuel reserves and increasing threat of climate change. Oil prices have been consistently increasing in the last decade and more so in the last couple of years. Particularly in the first half 2008, the price of crude oil rose close to \$150/barrel. However, for various reasons (mainly an economic downturn), late 2008 saw oil prices dropping again close to \$40/barrel (Fig. 1). Owing to such uncertainties in the availability of oil and the volatility of prices, there has been a growing interest in the production of synthetic liquid fuels. Synthetic liquids from fossil sources like coal, natural gas, oil shale, tar sands or biomass have been used in the past or are being used today. Though such endeavors have proved uneconomical in the past, if the crude oil prices are sufficiently high, these technologies have the potential to become economically feasible, although much also depends on future policies to limit greenhouse gas (GHG) emissions.

Coal-to-liquids (CTL) technology, which produces liquid transportation fuels like diesel and gasoline, is of interest to the US because of the abundance of coal reserves in this country. Previously, CTL technology was widely used in the World War II era, mostly in Germany and Japan. During the first oil shock of

1970s, there was renewed interest in CTL technology which later faded when the oil prices began to fall (Schulz, 1999; Steynberg and Dry, 2004).

In the most commonly used CTL technology, coal is first gasified to produce synthesis gas (or syngas) which, in turn, is catalytically treated in a Fischer–Tropsch (FT) process to produce different liquid fuels, mainly gasoline and diesel. These fuels are very clean in terms of criteria air pollutants such as nitrogen and sulfur oxides and aromatic hydrocarbons (Probstein and Hicks, 1985).

Two general configurations of CTL plants are shown in Fig. 2. In a typical commercial CTL plant shown in Fig. 2(a), the unconverted syngas from the FT reactor is recycled to the reactor to increase the production of liquids. In this paper, such plants are called “liquids-only” plants. Another configuration, called a “co-production” plant as shown in Fig. 2(b), is not yet commercial. Here, unconverted syngas from the FT reactor, instead of being recycled, is combusted in a combined cycle power plant to generate electricity that is sold to the grid.

2. Objectives of this study

A major concern of CTL plants is their emissions of carbon dioxide (CO₂). Coal is predominantly a carbon-rich feedstock whereas liquid fuels are richer in hydrogen. For example, the ratio of hydrogen to carbon atoms in liquid fuels is roughly from

* Corresponding author. Tel.: +1 412 268 5284; fax: +1 412 268 3757.
E-mail address: mharichandan@cmu.edu (H.C. Mantripragada).

2 to 1, whereas in coal the ratio is roughly from 1 to 1. In the process of converting coal to liquids, the excess carbon in coal is emitted in the form of CO_2 . A stream of concentrated CO_2 produced in the syngas-upgrade step is usually vented to the atmosphere (Steynberg and Dry, 2004). Subsequent combustion of the liquid fuels also generates CO_2 . As a result, coal liquids have at least twice the life-cycle CO_2 emissions compared to conventional crude oil-derived liquid fuels (Farrell and Brandt, 2006; Jaramillo, 2007).

The plant-level CO_2 in a CTL process can be offset using carbon capture and storage (CCS) technology, in which the captured CO_2 is compressed and transported to a deep geological formation, where it is sequestered. An analysis of technical and economic effects of CCS options on the CTL process is the major focus of this paper.

There have been a few recent studies dealing with techno-economic evaluation of CTL plants (Bridwater and Anders, 1991; Neathery et al., 1999; NETL/DoE, 2007a; Steynberg and Nel, 2004; SSEB, 2006; Williams et al., 2006). Although most studies consider co-production facilities, such plants are still at a conceptual stage; all of the commercial FT-based synthetic liquid production plants are the liquids-only configuration (Steynberg and Dry, 2004).

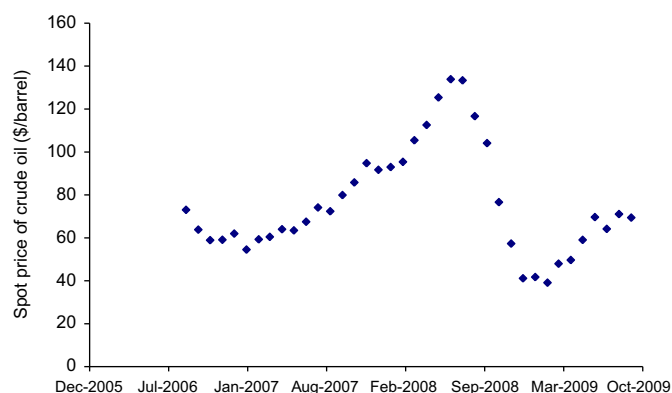


Fig. 1. Monthly crude oil prices.
[Source: www.eia.doe.gov].

There is little work available, however, on the effects of CO_2 emission constraints on the cost of co-production plants. Also, there is a lack of detailed economic assessments of a CTL plant which systematically analyze the uncertainties in key parameters affecting the cost of coal liquids, including the effects of possible future carbon constraints.

To address these issues, a comprehensive techno-economic assessment model of CTL plants, capable of incorporating CCS and co-production options, has been developed. It is used in this paper to study:

- the performance and cost of liquids-only and co-production plant configurations;
- the effects of uncertainty and variability of important process parameters on the cost of liquids production;
- the effects of carbon constraints on the cost of CTL plants and
- the efficiency and CO_2 emissions advantages of co-production over separate production of liquid fuels and electricity.

While a “life-cycle” perspective of this type is beyond the scope of this paper, other studies have examined these broader dimensions (Jaramillo, 2007; Marano and Ciferno, 2001).

3. Process performance model

All major components of a CTL plant, including coal gasifiers, gas clean-up system, FT synthesis and power generation were modeled using the Aspen Plus process simulator software (Aspen Technology, 2007). For a given plant capacity and specified operating conditions the model calculates all plant mass and energy flows. The major plant sections are briefly discussed below.

3.1. Syngas generation and clean-up

The gasification process involves reactions of coal with water (steam or slurry) and oxygen supplied from an air separation unit (ASU). In this paper, a General Electric (GE) entrained flow gasification technology is modeled. Coal slurry and 95% pure oxygen are fed into the gasifier to produce syngas. The hot syngas

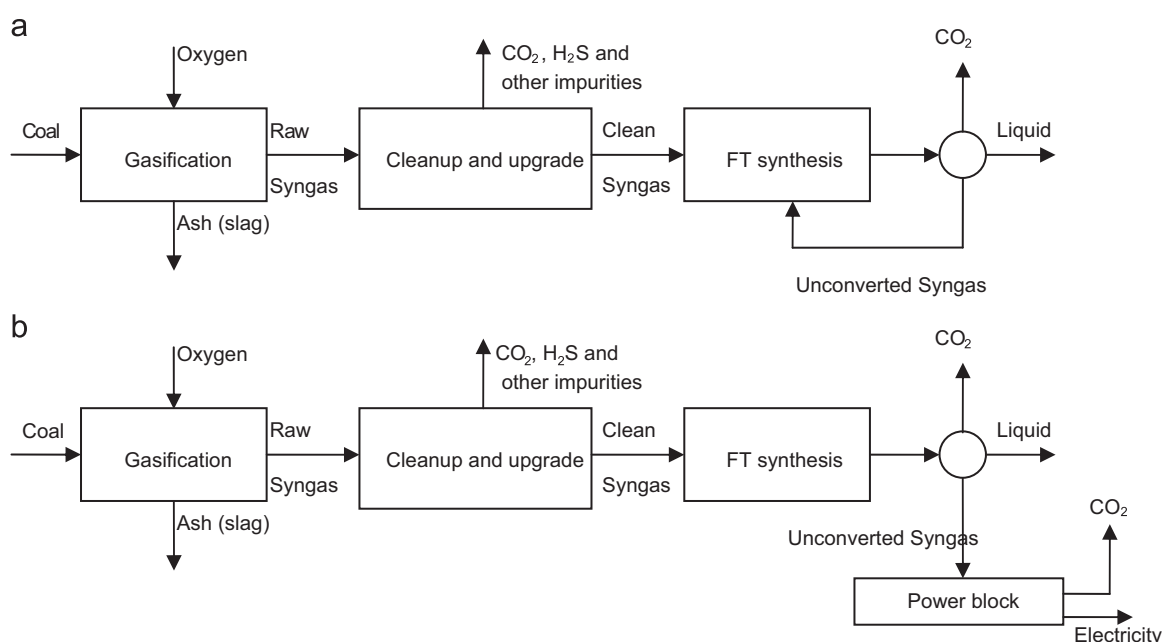


Fig. 2. (a) Liquids-only and (b) co-production configurations of coal-to-liquids plants.

Table 1
Coal and gasifier characteristics used in this case study.

Coal characteristics (as-received basis)	Gasifier conditions	Syngas composition (vol%)	
Illinois#6	GE		
Bituminous	Slurry-feed	CO	33.9
HHV—26 MJ/kg	1316 °C	H ₂	33.6
Ultimate analysis	5.6 MPa	CO ₂	15.3
(wt%)			
Ash—9.7	O ₂ /C—0.47 mol/mol	H ₂ O	14.7
Carbon—63.7	H ₂ O/C—0.55 mol/mol	CH ₄	0.1
Hydrogen—4.5		H ₂ S	0.6
Moisture—11.1		Other	1.8
Other—11			

is cooled in a radiant cooling section to produce high pressure steam, which is used for electricity generation. We note that most previous CTL studies assume a quench cooling gasifier design, which has no scope for this kind of heat recovery for power generation. Considering the products in a typical gasifier product gas, the following reactions are likely to take place (Holt and Alpert, 2001):

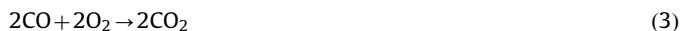
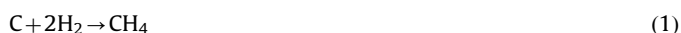


Table 1 shows the gasifier and coal characteristics as well as the composition of syngas at the gasifier exit, calculated by the model.

The syngas from the gasifier is cooled and then cleaned in a high-efficiency Selexol process in which H₂S and CO₂ impurities are separated from syngas (NETL/DoE, 2007a; Gray and Tomlinson, 1990). This clean-up step, combined with polishing, is necessary since FT catalysts are sensitive to sulfur poisoning and the presence of CO₂ is detrimental to FT reactions. Because of this, CO₂ capture is an intrinsic part of a CTL process, although normally the captured CO₂ is vented to the atmosphere.

3.2. Fischer–Tropsch synthesis

The clean syngas is fed into a low-temperature (250 °C) slurry-based FT reactor (LTFT) using Fe-based catalyst in which the syngas is converted to hydrocarbons of different chain-lengths (Eq. (8)).



Besides the FT reactions, other reactions also take place in the FT reactor. The water gas shift (WGS) reaction (Eq. (2)) is highly relevant where Fe catalysts are used. From Eq. (8), it can be ascertained that the H₂/CO ratio of the reactants of the FT reaction should be 2. However, the coal-derived syngas here (Table 1) has a ratio of approximately 1.0. In other processes the H₂ content of the syngas can be increased by adding a WGS reactor. However, in slurry-based LTFT reactors using Fe catalysts, the WGS reaction takes place within the FT reactor, producing the necessary H₂ required for the FT reactions (Steynberg and Nel, 2004). For

slurry-based LTFT reactors, the minimum H₂/CO ratio required for the syngas input stream is 0.67. Since this value is exceeded in the present case, no additional WGS reactor is required (Gray and Tomlinson, 1990). Within the FT reactor, the extent of WGS reaction with respect to FT reactions depends on the inlet H₂/CO ratio, with lower H₂/CO ratio leading to a higher WGS reaction rate (Raje and Davis, 1997). The CO₂ produced in the FT reactor by the WGS reaction is separated from the unconverted syngas and other gaseous products using a second Selexol system. The temperature of FT reactor is maintained by generating medium pressure steam which is used for power generation.

FT products contain a large number of hydrocarbon chains ranging from carbon numbers 1 to more than 30. The distribution of chain lengths depends on a parameter called the chain growth probability (α) and can be represented by the Anderson–Schulz–Flory (ASF) equation as shown in Eq. (9). The chain growth probability depends on the operating parameters of the FT reactor such as temperature and the catalyst used. For the LTFT reactor using Fe-based catalyst considered here, the value of α is assumed to be 0.9 (Steynberg and Dry, 2004)

$$W_n = n(1-\alpha)^2 \alpha^{n-1} \quad (9)$$

The performance model calculates the conversion of clean syngas in the FT reactor using Eqs. (2) and (8) and the distribution of carbon numbers in the products using Eq. (9).

The gaseous products from the FT reactor include unconverted syngas, CO₂ and lighter hydrocarbons (C1–C4). In the liquids-only configuration, these gases are separated from the liquids and recycled back to the FT reactor after separating CO₂ and reforming the light hydrocarbons to CO and H₂. A small fraction of this recycled gas is vented to the atmosphere in order to maintain the quantity of inert species in the recycle loop. In the liquids-only plant with CCS, a small release of carbon emissions occurs because of this purge gas.

3.3. Power generation block

The unconverted syngas from the FT reactor is either recycled back to the reactor (liquids-only configuration) or combusted in a gas turbine (co-production configuration) to generate electricity. The gas turbine modeled here is a GE 7FB design, which operates at a pressure ratio of 18.5 and a firing temperature of 1395 °C (2550 °F). Output from a simple cycle power plant operating on natural gas is 185 MW. Output from a combined cycle plant is 280 MW at a heat rate of 6256 kJ/kWh (LHV). Before the syngas is combusted in the gas turbine, it is mixed with nitrogen from the ASU in order to control NO_x emissions while also increasing the mass flow rate through the turbine. After dilution with N₂, the fuel entering the gas turbine combustion chamber has a lower heating value between 4.5 and 4.8 MJ/m³, as in an IGCC plant (NETL, 2007b).

The hot exhaust gases from the gas turbine are cooled in a heat recovery steam generator (HRSG), which generates high pressure superheated steam (9.8 MPa, 538 °C). The HRSG also reheats the intermediate pressure (2.1 MPa) steam to 538 °C. This steam is expanded in a 3-stage (9.8, 2.1 and 0.28 MPa) steam turbine with intermediate reheating to generate additional electricity. As noted earlier, steam also is generated from raw gas cooling at the gasifier exit and from cooling of the FT reactor. For the co-production cases, steam from this waste heat recovery is sent to the HRSG and used to generate electricity. After supplying all auxiliary load requirements (i.e., power for the ASU, CO₂ capture unit, and in the CCS case, CO₂ compression), net electricity generated is sold to the grid.

In the liquids-only plant, high-pressure and medium-pressure steam from waste heat recovery are sufficient to generate power to meet all auxiliary load requirements of the plant, including the compressor requirements in the CCS case. For the liquids-only plant without CCS, the option of generating additional steam power from waste heat as a co-product for export was not included in our results because the amount of such power is small compared to the net power generated in a co-production plant and thus would not have a significant effect on key results. Although there is potential to generate additional electricity by combusting the gases in the FT recycle loop, this would incur additional costs for a combined cycle power plant. Since sufficient steam power is available to meet all plant requirements, this option is not considered in this paper, consistent with current commercial practice.

3.4. CO₂ capture and sequestration

For both the liquids-only and co-production configurations, concentrated CO₂ streams are available from the gasification and FT synthesis sections. When CCS is employed, this CO₂ is compressed to 150 bar, transported and sequestered deep underground. Since CO₂ is captured as part of the process, the energy penalty of CCS is the additional compressor energy. Additionally, in a co-production configuration, combustion of unconverted syngas in the gas turbine results in more CO₂ emissions, which can be captured using a post-combustion capture process. Thus, two configurations of the co-production plant are modeled:

- Capturing only the pre-combustion (Selexol-based) CO₂ from the gasification and FT units (99% capture), but not the CO₂ from the gas turbine exhaust gases. This case is referred to as CCS(a) in this paper.
- Capturing both pre-combustion (Selexol) and post-combustion (MEA, 90% capture) CO₂. In this case, called CCS(b), exhaust CO₂ from the gas turbine is captured using an amine-based (MEA) chemical absorption process. Thus for this configuration, the energy penalty of CCS also includes the energy required for post-combustion CO₂ capture. That includes the low-pressure steam needed for solvent regeneration, plus electricity for pumps and other equipment (Rao and Rubin, 2002). The two concentrated CO₂ streams from the plant are then compressed and transported to a geological sequestration site.

For the liquids-only configuration, electricity is produced from waste heat in an amount that is just sufficient to meet the internal requirements of the plant, with no export to the grid. For the case with CCS in a liquids-only plant, more electricity is produced internally to handle the additional compressor energy requirements. However, since only waste heat is used, the efficiency of a liquids-only plant does not change.

4. Process cost model

The results from the performance model are then input to a cost model, which calculates the total capital cost (US \$/barrel/day) and operating costs (US M\$/year) of the plant, as well as the overall cost of the liquid product (US \$/barrel). Models for the direct cost of all the process sections except the FT process are obtained from Integrated Environmental Control Model (IECM) for an IGCC plant (IECM-cs, 2007). New cost models for the FT process were developed through a regression of cost data from recent literature (NETL/DoE, 2007a; Sseb, 2006). In this model, the liquid product from the FT reactor system is considered to be

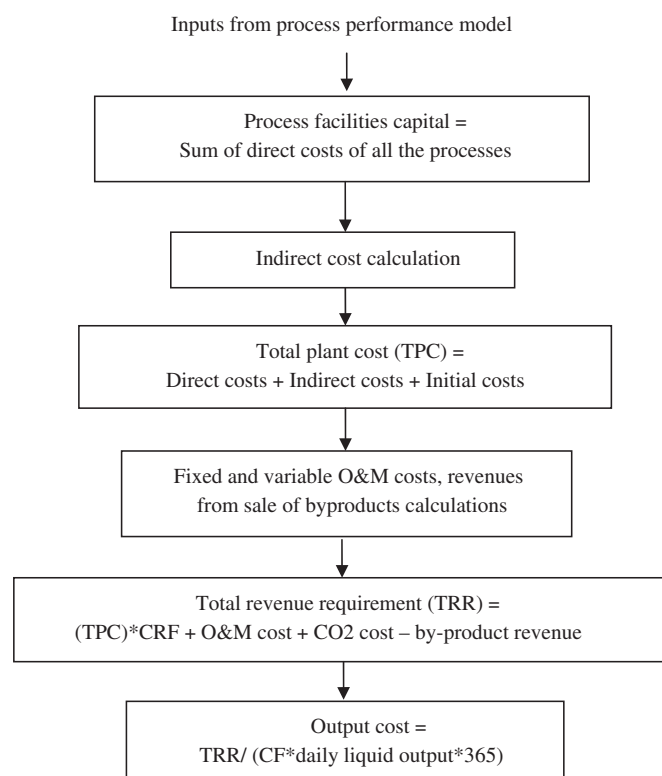


Fig. 3. Flowchart depicting the process cost model (EPRI, 1986).

equivalent to crude oil. Thus, the subsequent production of various refined products (which requires further refinery processing) is outside the scope of the CTL process model. All costs are expressed in constant (levelized) 2007 US dollars. The cost calculation methodology is shown in Fig. 3.

5. Case study results

This section demonstrates the application of the techno-economic assessment models of the CTL plants described above. Both liquids-only and co-production configurations are modeled for a plant producing 50,000 barrels/day of liquids from Illinois#6 bituminous coal. Plants with and without CCS are considered. Through an uncertainty and sensitivity analysis, the important factors that affect the cost of liquid fuel production from coal, including the plant size, price of coal, economic assumptions, technical factors and carbon constraints are studied. All costs are reported in constant 2007 USD.

5.1. Liquids-only plant without CCS

The process performance and cost models were applied to the liquids-only configuration. Table 2 shows the main results for performance and cost of the plant, both with and without CCS.

For the base case plant without CCS, the amount of coal needed to produce 50,000 barrels/day of liquid fuel product is about 19 kilo-tonnes/day. The resulting emissions of CO₂ are close to 24.7 kilo-tonnes/day. The overall plant efficiency, calculated as the energy content of liquid products per unit input energy (based on higher heating value), is about 56%. For this plant without CCS, the capital cost is \$91,900 per daily barrel and the cost of product liquid is about \$76/barrel.

Table 2
Performance and cost results from the techno-economic models, for liquids-only and co-production configurations, with and without CCS (all costs in constant 2007 US dollars).

Parameter	Liquids-only		Co-production		
	No CCS	With CCS	No CCS	With CCS	
				CCS(a)	CCS(b)
Coal consumption (10 ³ tonnes/day, as-received)	19.05		23.57		
CO ₂ emissions (10 ³ tonnes/day)	24.69		35.39	8.01	0.80
Net power output (MW)	–	0.29	910 ^a	835	765
Efficiency (% higher heating value)	56.1	56.1	53.3	52.2	51.3
Specific capital cost (10 ³ \$/daily barrel)	91.9	93.1	115.8	117.1	125.9
Cost of liquid product (\$/barrel, \$0/tonne CO ₂)	76.1	81.8	58.5	69.4	78.2
Cost of liquid product (\$/barrel, \$25/tonne CO ₂)	88.4	81.8	76.2	73.4	78.6

^a Net power = gross power output of 1200 MW (gas turbine power = 420 MW, steam turbine power = 780 MW), less auxiliary power requirement of 290 MW.

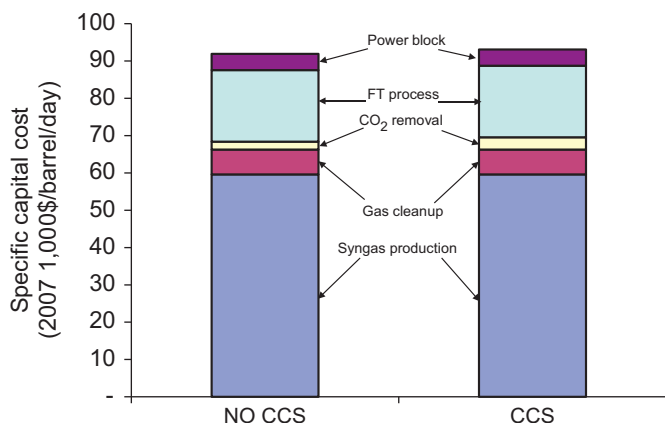


Fig. 4. Distribution of capital cost for a liquids-only plant. Syngas production section accounts for close to 60% of the capital cost, followed by the FT section.

5.1.1. Details of capital and O&M costs

Fig. 4 shows the breakdown of capital cost among the major sections of the plant. Syngas production contributes more than 60% of the capital cost, followed by the FT process (about 20%). Oxygen production and gasification trains account for the bulk of costs in the syngas generation section. Cryogenic air separation units (ASU) used to produce oxygen are highly capital intensive and also have a significant energy penalty. The maximum capacity of ASUs currently available is about 2725 tonnes/day of oxygen, so that multiple trains of ASUs are needed for large applications, which reduces economies of scale. Also, the cost of ASUs increases exponentially with increasing purity of oxygen. Oxygen purity needs to be high (95%) for FT applications since the presence of inerts such as N₂ hampers the FT reaction and decreases its efficiency. Similarly, gasification equipment also has high capital costs, with limited economies of scale before multiple trains are required by size limitations.

The contribution of capital and operating cost components to the liquid product cost is depicted in Fig. 5. It is clear that capital cost component is the biggest contributor, accounting for 60% of the total cost. Operating costs occur in the form of consumables like coal and are the next major contributor to cost.

5.1.2. Effect of a carbon price on the cost of liquid products

In the future, it is likely that there will be an implicit or explicit cost associated with CO₂ emitted into the atmosphere. To see the effect of a carbon price on the product cost, a price of \$25/tonne CO₂ was considered, with the results shown in Table 2. The cost of liquid product increases to over \$88/barrel, an increase of 16%. Even with a

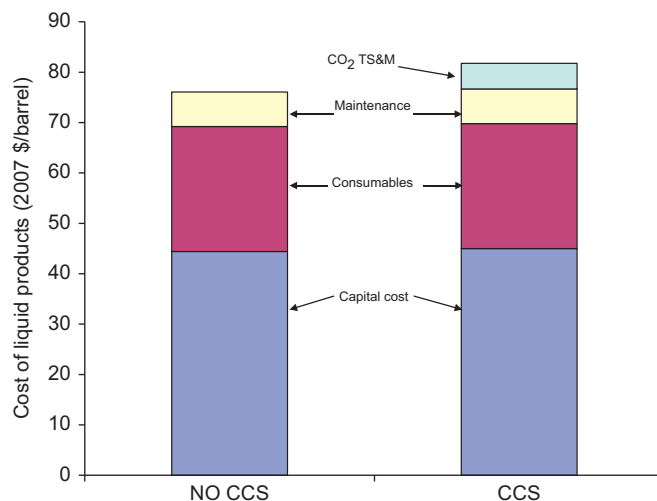


Fig. 5. Distribution of liquid product cost for a liquids-only plant. Capital cost component accounts for close to 50% of the total cost.

CO₂ price of \$25/tonne, the cost of liquids from a liquids-only plant is in the range of crude oil prices seen in the recent past.

5.2. Liquids-only plant with CCS

In a CTL plant which is designed for CCS, additional capital costs occur in the form of compressors which compress the concentrated CO₂ to a pressure of approximately 150 bar so that it is easier to transport and inject CO₂ in supercritical state. Pipeline transportation costs, geological sequestration and monitoring costs are the additional operating expenses.

With the addition of CCS, emissions are reduced to 290 tonnes/day, in the form of CO₂ and other carbon-containing components from the purge gases of FT loop. As explained previously, the additional energy required for CO₂ compression is assumed to be produced by waste heat recovery. Hence, the efficiency of the plant does not change with the addition of CCS, although some additional capital expense is required. Total capital cost increases to \$93,100 per daily barrel (Table 2), an increase of 1.3%, while the output cost increases to about \$82/barrel, an increase of 7.5% from the plant without CCS. Thus, operating costs of CCS affect the cost of product liquids more than the increase in capital costs. Because of the extremely low CO₂ emissions, a price on CO₂ does not affect the product cost significantly. It can be seen in Table 2 that, even with CCS, the cost of liquid products is comparable to the recent crude oil prices. Fig. 4 shows the breakdown of capital costs among process

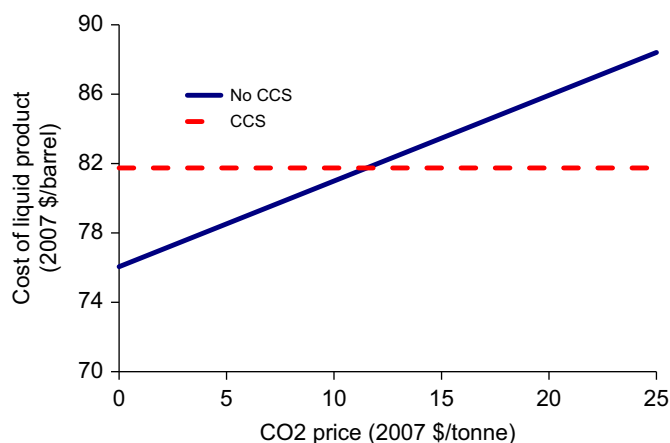


Fig. 6. Effect of carbon price on cost of liquid product. CCS becomes more economical when the CO₂ price exceeds \$12/tonne.

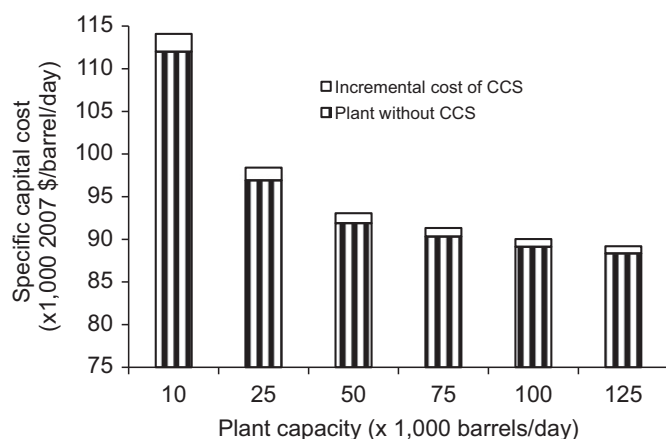


Fig. 7. Specific capital costs (\$/barrel/day) of plants with different capacities. Economy of scale is evident.

sections and Fig. 5 shows the breakdown of product cost in terms of the capital and operating cost components.

It also can be seen from Table 2 that the cost of liquid product from a plant with CCS is less than from a plant without CCS, paying \$25/tonne CO₂ emitted. Fig. 6 shows the effect of increasing CO₂ price on the cost of liquid product. For the assumptions in this case study, it can be seen that for a CO₂ price over \$12/tonne, a plant with CCS is more economical than a plant without CCS.

5.3. Sensitivity and uncertainty analysis

Because there are known economies of scale, the effect of plant capacity on the capital cost and liquid product cost of a CTL plant was first analyzed. The plant capacity was varied from 10,000 to 125,000 barrels/day. The variation in specific capital cost is depicted in Fig. 7. The specific capital cost of a 125,000 barrels/day plant is about 40% less than that of a 10,000 barrels/day plant, which shows the economy of scale exhibited by CTL plants. Since capital cost is the dominant factor, the cost of liquid product also decreases with increasing plant capacity. However, the effect of scale on the specific capital cost shows a decreasing trend, as shown in Fig. 8. For example, the specific capital cost of a 100,000 barrel/day plant is only 5% less than a 50,000 barrel/day plant.

The breakeven CO₂ price at which CCS becomes more economical for different output capacities was also calculated. It was found that the breakeven CO₂ price is not highly sensitive to the

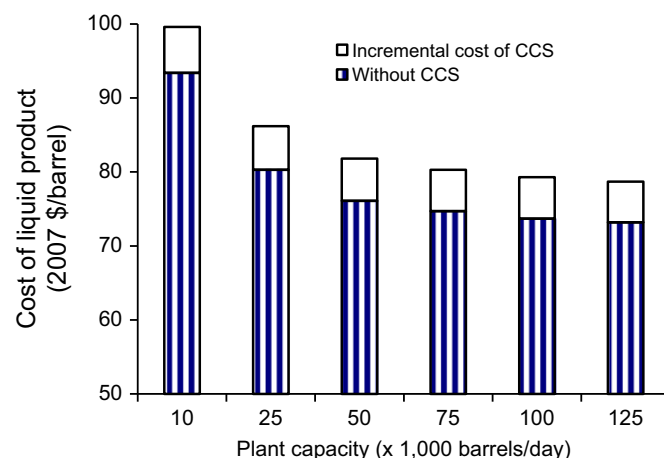


Fig. 8. Cost of liquid product (\$/barrel) of plants with different capacities. Since capital cost component is the major contributor to the product cost, there is economy of scale even here.

plant output capacity. In this range of plant output capacities, the breakeven CO₂ price varies from \$11/tonne for the largest plant to \$14/tonne CO₂ emitted for the smallest plant considered. Beyond these CO₂ prices, a plant with CCS becomes more economical than a plant without CCS. The reason for the relative insensitivity of breakeven CO₂ price to plant output capacity is that the added variable cost of the CO₂ tax is a relatively small portion of the total plant cost, which is dominated by the capital cost component.

In addition to the deterministic results shown above, a probabilistic analysis was conducted to examine the impacts of uncertainty and/or variability in key operating and cost parameters. The uncertainty analysis employed a Monte Carlo simulation technique (Morgan and Henrion, 1990), implemented within the Matlab code for the CTL plant model. Each selected parameter was assigned a range of values and a probability distribution based on values derived from the literature, as discussed below. Using the random number generator within Matlab, 20,000 simulations were performed for each case to generate cumulative probability distributions for cost results. The ranges and probability distributions given to each cost parameter are shown in Table 3.

The assumptions and data leading to these ranges are as follows:

- For the direct capital cost of each plant section, a triangular distribution ranging from $\pm 25\%$ of the deterministic value was assumed, in the absence of more detailed data, similar to other cost estimation methods (NETL, 2007b). For the indirect components of capital cost, ranges are adapted from Chen and Rubin (2008).
- To amortize capital costs, a wide range of 5–20% for the capital recovery factor (CRF) was assumed (with a triangular distribution about the nominal value of 15%). This reflects variability rather than uncertainty, i.e., a range of investment perspectives, with the lower values representing significant public sector subsidies or investment and the high end a private sector investment.
- A CTL plant combines elements of petroleum refinery and a process plant using coal. The capacity factor of U.S. petroleum refineries has been very high (about 90%) over the past two decades (EIA, 2009) but because of other components like coal gasifiers, a slightly lower capacity factor of 85% is used for the base case. The uncertainty or variability in levelized capacity factor over the life of the plant is denoted by a triangular

Table 3
Assumed uncertainty distributions for model cost parameters.

Parameter	Base case value	Distribution
<i>Capital cost Parameters</i>		
Direct capital cost (DC)	Calculated from the cost model	Triangular (–25%, base case, 25%)
General facilities capital (GFC)	15% of DC	Triangular (10%, 15%, 20%)
Engineering and home office (EHO)	10% of DC	Triangular (7%, 10%, 12%)
Process contingency	25% of DC	Triangular (10%, 25%, 40%)
Project contingency	15% of DC	Triangular (10%, 15%, 20%)
Royalty charges	10% of DC	Triangular (7%, 10%, 12%)
<i>Financial parameters</i>		
Capital recovery factor (CRF)	0.15	Triangular (5%, 15%, 20%)
<i>Plant operation</i>		
Capacity factor	0.85	Triangular (0.75, 0.85, 0.95)
<i>Operating costs</i>		
Coal price	\$60/tonne	Uniform (20, 100)
CO ₂ transport cost	\$5/tonne CO ₂	Uniform (1.3, 10.4)
CO ₂ storage cost	\$5/tonne CO ₂	Uniform (0.65, 10.4)
Sequestration monitoring cost	\$0.25/tonne CO ₂	Uniform (0.13, 0.39)

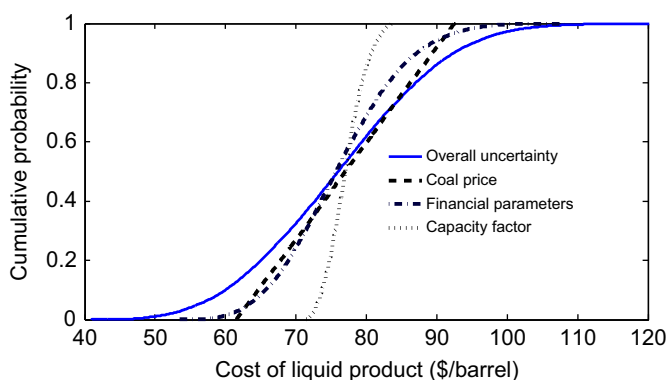


Fig. 9. Cumulative probability distributions of cost of liquid product (2007 \$/barrel) for plant without CCS. Capital cost assumptions, particularly the capital recovery factor, cause the maximum uncertainty in the output cost.

distribution with minimum, most likely and maximum values of 75%, 85% and 95%, respectively.

- Mine-mouth coal prices since 1950 (EIA, 2009) show a very wide variation ranging from \$20 to \$100/tonne for bituminous coals, with no systematic pattern. Thus, a uniform distribution about the current price of about \$60/tonne for Illinois#6 was assumed.
- The operating costs of CCS can vary over a wide range depending on the distance transported and the characteristics of the storage site. The ranges shown in Table 3 were adopted from the IPCC (2005). The values used for deterministic calculations are approximate averages of these costs.

Fig. 9 shows the effect of the uncertainties on the product cost for the case without CCS. Uncertainties in capital cost parameters, together with the wide variation in CRF, are the cause of most of the variation in product cost. The 90% confidence interval of product cost is \$56–\$97/barrel for a plant without CCS, compared to the deterministic estimate of \$76/barrel. A similar analysis shows a 90% confidence interval of and \$62–\$104/barrel for the CTL plant with CCS, compared to the deterministic estimate of

\$82/barrel. The variation for both plant configurations is almost by a factor of 2, showing how important the economic assumptions are in estimating the profitability of a CTL plant.

The effect of the uncertainties mentioned above on the breakeven CO₂ price where CCS becomes the more economical option was also studied. The 90% confidence interval for the breakeven CO₂ price was found to be between \$6.4 and \$19.4/tonne CO₂ emitted, for the 50,000 barrels/day plant capacity considered in this case study. This is a much wider range of values than the variation due only to plant capacity effects, as noted earlier. In general, a higher breakeven CO₂ price is needed under conditions when variable costs are a smaller portion of the total cost. Thus, uncertainty in key model parameters contributes more to the variation in breakeven CO₂ price than the plant output capacity.

The sensitivity of cost results to process performance parameters has not been explicitly considered in this paper. Inclusion of such additional uncertainties or variability would further broaden the distribution function results in Fig. 9. Note, however, that since the capital cost models are based on process parameter values, the cost uncertainties in Table 3 may be considered to implicitly reflect performance model uncertainties to some degree. Future studies will explicitly address the sensitivity of results to process performance parameters (in addition to cost parameters), as well as sensitivity to different combinations of technology options (such as gasifier choice) and fuel feedstock (such as different coal ranks).

5.4. Analysis of co-production plant configuration without CCS

For a CTL plant with co-production of electricity the unconverted syngas and the FT product gases are combusted in a combined cycle power plant, as discussed earlier. The performance and cost results for the base case assumptions were shown in Table 2.

Without CCS, the coal needed to produce 50,000 barrels/day of liquid fuel output is about 23.6 kilo-tonnes/day, 24% more than that of the liquids-only case. The emissions of CO₂ are 35.4 kilo-tonnes/day, 43% higher than the liquids-only case. The overall plant efficiency is slightly less than the liquids-only case. The gas turbine power plant generates close to 420 MW while the total steam power from the combined cycle power plant plus waste heat recovery in the gasification and FT sections is about 780 MW. After supplying auxiliary loads totaling 290 MW, the co-production plant without CCS produces more than 900 MW of net electricity which can be sold to the grid for additional revenue.

The capital cost of this plant, normalized on only the liquid fuel product, is \$115,900/barrel/day, about 26% higher than the liquids-only plant without CCS. However, if the by-product electricity is sold to the grid at \$80/MW h, the cost of liquid product falls to \$58/barrel without a CO₂ price and \$76/barrel with a CO₂ price of \$25/tonne CO₂. In both cases, these product costs are lower than for the corresponding liquids-only plants. Thus, despite the increase in capital cost, the revenue generated by electricity co-production produces cheaper liquid fuels than liquids-only plants.

5.5. Co-production plant with CCS

Table 2 also shows the performance and cost results when CCS is applied to a co-production plant in one of two ways.

Applying CCS only to the pre-combustion stage reduces the CO₂ emissions to 8000 tonnes/day, while the net electricity output falls to 830 MW (net efficiency of 52.2%). The capital cost increases only by 1%, similar to the liquids-only case, because the only additional equipment required are the CO₂ compressors. If

the electricity is sold at \$80/MWh, the cost of liquid product is \$69/barrel without a CO₂ price and \$73/barrel with a price of \$25/tonne CO₂ emitted. Hence, pre-combustion CCS is cheaper than paying a CO₂ price of \$25/tonne when the electricity is sold at \$80/MWh.

With the addition of the post-combustion capture from the exhaust gases of the gas turbine, CO₂ emissions can be reduced to 800 tonnes/day. The energy penalty for CO₂ capture and compression reduces the net electricity output is reduced to 765 MW and the efficiency of the plant decreases by 2% points, compared to the case without CCS.

The total plant capital cost increases by about 10% compared to a co-production plant without CCS. This is a much higher increase compared to the plant with only pre-combustion CO₂ capture. The liquid product cost is now \$78/barrel without a CO₂ price and almost the same with a CO₂ price of \$25/tonne, when electricity is sold at \$80/MWh. Here, unlike the liquids-only case, a CO₂ price of \$25/tonne is not enough to make CCS a cost-effective option when electricity is sold at \$80/MWh. The revenue generated by electricity compensates for the cost of CO₂ emissions. The CO₂ price would have to increase to \$35/tonne

5.6. Co-production vs. liquids-only

Fig. 10 shows the effect of the electricity price on the cost of product liquids for cases with and without CCS, with a CO₂ price of \$25/tonne. Electricity prices at which co-production breaks even with liquids-only plant (based on the cost of liquid products) are indicated with arrows. For all the cases, co-production plants become cheaper than liquids-only plants when the selling price of electricity is in the range \$40–80/MWh. This price range corresponds to current market prices of electricity in the U.S., which can be expected to grow when there are carbon constraints. At higher prices, co-production becomes increasingly favorable.

In terms of the CCS options, for the CCS(a) case where only the pre-combustion CO₂ is captured and there is a CO₂ price of \$25/tonne, CCS is cheaper than no CCS, across the range of electricity selling prices considered here.

On the other hand, for the CCS(b) case where both pre-combustion and post-combustion CCS are used, CCS becomes more costly than paying a CO₂ price of \$25/tonne, if the electricity selling price exceeds about \$45/MWh. Above that price, the

revenue generated by electricity sales exceeds the cost of paying for the additional CO₂ capture, so that capture is not economical.

Thus, the economic feasibility of CCS for a co-production plant depends on various factors such as the extent of CO₂ capture, the cost of CO₂ emitted and the electricity selling price. It was found that, for an electricity price of \$100/MWh, CCS(b) is cheaper only if the CO₂ price exceeds \$35/tonne.

6. Discussion and conclusion

The performance and cost results presented in this paper indicate that while liquids-only plants are more thermodynamically efficient, the most economical way that liquids can be produced from coal (in terms of the cost of liquid product) is in a co-production plant that also generates electricity for sale. The cost of liquid product from either a liquids-only plant or a co-production plant was seen to be comparable to the crude oil prices of the past several years. However, CTL plants are also highly capital intensive, with the capital cost component accounting for about half the total product cost. The case studies presented have showed that the capital requirement for a 50,000 barrels/day liquids-only CTL plant is on the order of \$5 billion. It was also seen that when uncertainties in cost parameters were considered, the cost of liquid product can vary by nearly a factor of two. Thus, investments in CTL plants carry a significant financial risk.

Though co-production plants are much costlier to build than liquids-only plants, the added electricity sales revenue can significantly lower the cost of the liquid product at current market prices of electricity. It was seen that the liquid product cost from a co-production plant decreases rapidly with increasing electricity price. With future carbon constraints, electricity prices can be expected to increase. Thus, co-production plants can help reduce the financial risks of a CTL plant.

Plant-level CO₂ emissions are high but can be greatly reduced by using the carbon capture and storage (CCS) technology. The incremental capital costs of CCS are minimal for a current liquids-only plant because CO₂ capture is already included in the base plant process. Even with CCS, the liquid product costs are comparable to recent crude oil prices.

Future climate change policies will have a major impact on the CO₂ emissions from a CTL plant (Dooley and Dahowski, 2009). Under cap-and-trade regime, or an emissions tax policy, CCS is the lowest-cost option for a liquids-only plant when the CO₂ price exceeds about \$12/tonne, based on the results presented here. However, for a co-production plant with both pre-combustion and post-combustion CO₂ capture, revenue from electricity sales offsets the cost of a carbon price or tax. Thus, the CO₂ price has to be more than \$35/tonne to make CCS cost-effective for electricity prices in the range \$20–\$100/MWh.

Despite the economic and strategic benefits of CTL in a scenario of sustained high crude oil prices, the technology also poses significant environmental risks as well as resource requirements. Thus, any policy regarding large-scale implementation of CTL must take into account the economic, strategic and environmental risks and benefits of this technology.

Acknowledgment

This work was supported in part by US Department of Energy, National Energy Technology Laboratory (DOE/NETL) under the contract number DE-AC21-92MC29094. However, the authors alone are responsible for the content of this paper.

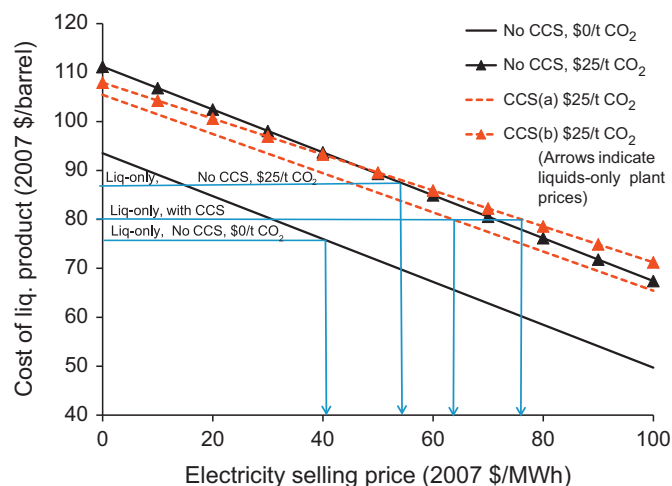


Fig. 10. Effect of selling price of electricity (cents/kWh) on the cost of liquid product from a co-production CTL plant. Arrows show the prices at which co-production breaks even with liquids-only plants in terms of cost of liquid products.

Appendix A. Supporting information

Supplementary data associated with this article can be found in the online version at [doi:10.1016/j.enpol.2011.02.053](https://doi.org/10.1016/j.enpol.2011.02.053).

References

- Aspen Technology, Inc., 2007, Ten Canal Park, Cambridge, MA 02141-2201.
- Bridwater, A.V., Anders, M., 1991. Economics of liquid fuels production by coal gasification. *Fuel* 70, 1193–1205.
- Chen, C., Rubin, E.S., 2008. CO₂ control technology effects on IGCC plant performance and cost. *Energy Policy* 37, 915–924.
- Dooley, J.J., Dahowski, R.T., 2009. Large-scale US unconventional fuels production and the role of CCS technologies in reducing their greenhouse gas emissions. *Energy Procedia* 1, 4225–4232 Elsevier Ltd.
- EIA, 2009. Data on Coal, Petroleum and Electricity. Energy Information Agency, US Dept. of Energy.
- EPRI, 1986. TAGTM – Technical Assessment Guide, vol. 1. Electricity supply, P-4463-SR, Electric Power Research Inst, Inc., Palo Alto, CA, December.
- Farrell, A.E., Brandt, A.R., 2006. Risks of oil transition. *Environmental Research Letters* 1, 014004.
- Gray, D., Tomlinson, G.C., 1990. Assessing the economic impact of indirect liquefaction process improvements: vol 1, development of the integrated indirect liquefaction model and baseline case, DoE DE91004668.
- Holt, N.A.H., Alpert, S.B., 2001. Integrated gasification combined-cycle power plants. *Encyclopedia of Physical Science and Technology*, 897–924.
- IPCC, 2005. Intergovernmental Panel on Climate Change special report on Carbon dioxide Capture and Sequestration. Cambridge University Press, New York.
- IECM-cs, 2007. Integrated Environmental Control Model, Carbon Sequestration Edition. Carnegie Mellon Univ. Center for Energy and Env. Studies.
- Jaramillo, P., 2007. A life cycle comparison of coal and natural gas for electricity generation and the production of transportation fuels. Ph.D. Thesis, Carnegie Mellon University.
- Marano, J.J., Ciferno, J.P., 2001. Life cycle greenhouse gas inventory for Fischer–Tropsch fuels, NETL.
- Morgan, G.M., Henrion, M., 1990. Uncertainty. Cambridge University Press.
- Neathery, J., Gray, D., Challman, D., Derbyshire, F., 1999. The pioneer-plant concept: co-production of electricity and added-value products from coal. *Fuel* 78, 815–823.
- NETL/DoE, 2007a. Baseline Technical and Economic Assessment of a Commercial Scale Fischer–Tropsch Liquids Facility. National Energy Technology Lab/US Dept. of Energy April.
- NETL, 2007b. Cost and Performance Baseline for Fossil Energy Plants. NETL/DoE May.
- Probststein, R.F., Hicks, R.E., 1985. *Synthetic Fuels Intl.* Student ed. McGraw Hill, Singapore.
- Raje, A.P., Davis, B.H., 1997. Fischer–Tropsch synthesis over iron-based catalysts in a slurry reactor. Reactoin rates, selectivities and implications for improving hydrocarbon productivity. *Catalysis Today* 36, 335–345.
- Rao, A.B., Rubin, E.S., 2002. A technical, economic and environmental assessment of amine-based CO₂ capture technology for power plant greenhouse gas control. *Environmental Science and Technology* 36, 4467–4475.
- Schulz, H., 1999. Short history and present trends of Fischer–Tropsch synthesis. *Applied Catalysis A: General* 186, 3–12.
- Steynberg, A., Dry, M., 2004. Fischer–Tropsch Technology, *Studies in Surface Science and Catalysis*, 152. Elsevier, Amsterdam.
- Steynberg, A.P., Nel, H.G., 2004. Clean coal conversion options using Fischer–Tropsch technology. *Fuel* 83, 765–770.
- SSEB, 2006. American Energy Security—Building a Bridge to Energy Independence and to a Sustainable Energy Future. The Southern States Energy Board, Norcross, Georgia.
- Williams, R.H., Larson, E.D., Jin, H., 2006. Comparing climate-change mitigating potentials of alternative synthetic liquid fuel technologies using biomass and coal. In: *Proceedings of Fifth Annual Conference on Carbon Capture and Sequestration*, DOE/NETL, pp. 8–11.