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Polygeneration of SNG, Hydrogen, Power, and Carbon Dioxide from Texas Lignite

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Disclaimer

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Introduction

The intent of this study is to investigate the feasibility of siting a lignite conversion plant in Texas at the minemouth of the Wilcox lignite deposit. The concept is to coproduce at least three products: electric power, hydrogen or substitute natural gas (SNG), and carbon dioxide. The electric power would be sold to the grid, the hydrogen would be sent by pipeline to the Gulf Coast petroleum refineries, the SNG would be sold as a natural gas supplement, and the carbon dioxide would be pipelined to the West Texas oil fields for enhanced oil recovery (EOR). EOR provides an economically attractive option for sequestering CO_2, and thus reducing greenhouse gas emissions from the lignite conversion. If natural gas prices continue to remain high in the future, there may be an opportunity for petroleum refiners to use low cost Texas lignite in place of natural gas to provide the hydrogen necessary for their refining operations. Also, lignite could be used to produce SNG as a natural gas supplement, and electric power could also be generated from the lignite and dispatched to the Texas grid. In the longer term, since SNG uses the same infrastructure as natural gas, SNG could be an attractive alternative as a hydrogen carrier for fuel cell based transportation systems. Finally, the West Texas oil fields continue to need carbon dioxide for EOR applications and carbon dioxide produced as a by product of Texas lignite conversion represents a potentially valuable resource close to the oil fields.

Site Selection

In the 1970s, concerns over a potential shortage of natural gas fostered considerable interest in the production of substitute natural gas from coal. A number of large-scale demonstration projects were planned. Of these projects, only one was ever built, in Beulah, North Dakota. The increased availability of North American natural gas in the 1980s and 1990s ended interest in large-scale production of SNG from coal. However, small-scale SNG production from LPG and naphtha has found a niche market in Japan and elsewhere. These systems provide back-up fuel for natural gas based power generation.

The increased demand for natural gas has resulted both in higher gas prices and more gas imports, a trend that is anticipated to continue. The Energy Information Administration (AEO 2004) predicts the wellhead natural gas price will rise to between $4.40 and $4.94 per million BTU by 2025, up from a 2002 price of $2.95. Much of the predicted future demand is anticipated to be supplied by imports of liquefied natural gas (LNG). Recent spot prices for natural gas have been volatile, ranging between $1.70 and $8.00 per million BTU and have averaged $5.81 per million BTU through August 2004. Therefore, the economics of SNG production may again be attractive, particularly if produced from low cost feedstocks and co-producing high valued by-products such as electricity.

Dakota Gasification Company's Beulah plant still produces about 170 million SCFD of SNG from lignite. In addition, it has expanded operations to co-produce ammonia, ammonium sulfate, phenol, and cresylic acid. In 2000, the plant began exporting carbon dioxide for use in enhanced oil recovery. Currently, about 95 million SCFD of CO_2 produced at the plant is transported via a 205 miles long pipeline to EnCana Corporation's Weyburn oil field in southern Saskatchewan. The CO_2 is used for tertiary oil recovery, resulting in 5,000 bbl/day of incremental oil production or an additional 130 to 140 million barrels of oil over the life of the project. The initial investment for this project was $1.3 billion (Canadian) by EnCana for field facilities and $100 million (U.S.) by Dakota Gasification for the pipeline and supporting facilities. Annual net revenue generated by the sale of the CO_2 is between $15 and 18 million. The Weyburn field is the subject of a long-term monitoring program to assess the final deposition of the CO_2 being injected in this project.

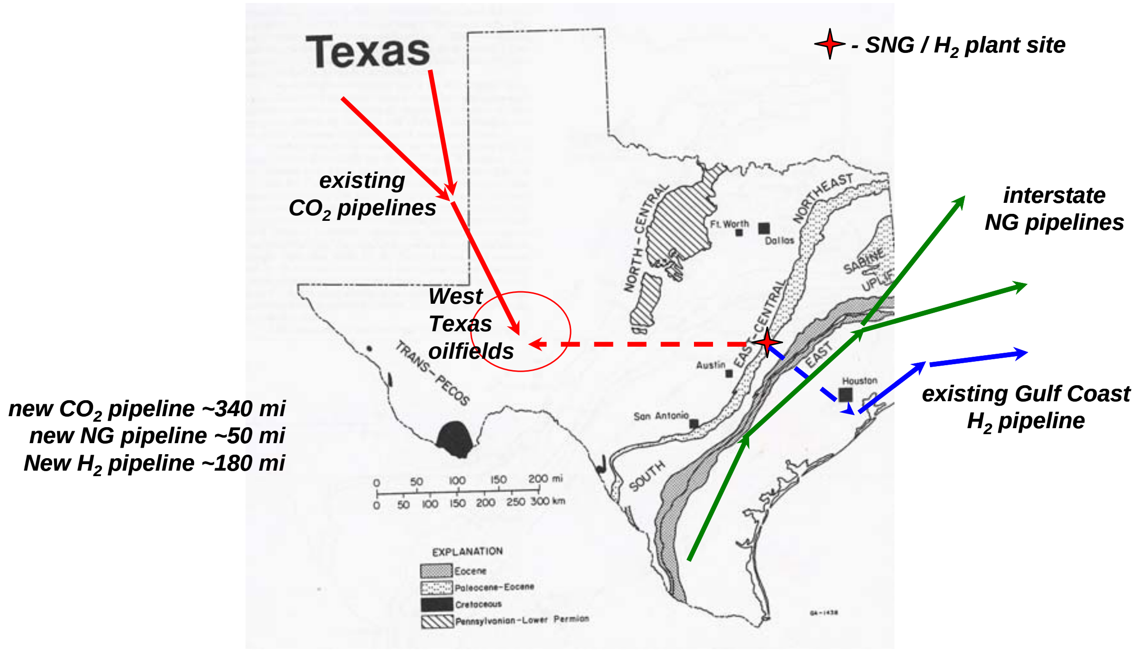
An alternative to SNG production from coal is the production of hydrogen. Currently, there is strong demand for hydrogen for petroleum refining, where it is used in hydrotreating and hydrocracking operations for the production of low sulfur transportation fuels. New requirements for ultra-low sulfur gasoline (2005) and diesel fuel (2008) mandated by the U.S. Environmental Protection Agency have resulted in the construction of new steam methane reformers for the production of hydrogen from natural gas. It is anticipated that the trend toward zero sulfur fuels will continue beyond 2010, and thus, the opportunity exists to produce hydrogen from low-cost coals as an alternative to natural gas. Longer term, hydrogen demand within the U.S. could significantly expand if hydrogen is one day used to power fuel cell vehicles.

Hydrogen is typically not transported long distances; however, a 300 mile long hydrogen pipeline is in operation along the U.S. gulf coast, providing hydrogen to the large concentration of petroleum refineries and petrochemical plants in this region. The pipeline connects Texas City, TX with Baton Rouge, LA, and services over 50 customers. The hydrogen supply was recently expanded and the pipeline currently moves about 560 million SCFD of hydrogen.

One potential drawback to the production of SNG or hydrogen from coal is the co-production of large quantities of CO_2, a greenhouse gas. Geological sequestration is one option for mitigating CO_2 emissions from coal conversion. This may be an especially attractive option if the CO_2 is used for enhanced oil recovery as is being done at the Weyburn field using CO_2 produced at Dakota Gasification. There are a number of other locations within the U.S. where low-priced coals are located near oil fields that currently employ CO_2 EOR, including Wyoming and West Texas. The Permian Basin in West Texas currently utilizes about 1,200 million SCFD of CO_2 and is the largest CO_2 EOR operation in the world. The bulk of the CO_2 currently used there is produced from natural CO_2 reservoirs located in northern New Mexico and southern Colorado and shipped via pipeline to West Texas. The remaining CO_2 comes from gas plants located in West Texas. About 160,000 incremental barrels of oil are produced per day due as a result of CO_2 EOR. The Wyoming oil fields currently utilize between 150 and 175 million SCFD, all of it supplied by gas plants. In addition to the Weyburn field in Saskatchewan, oil production in the Williston basin of North Dakota could be improved through CO_2 EOR, though no projects are currently in operation there. The production and reserves of surface-mineable coal located near these oil fields are given below:

	Production MM tons	Reserves MM tons	Price $/ton
Texas Lignite	50	10,000	14.00
North Dakota Lignite	30	7,200	8.50
WY Sub-bituminous	380	22,000	6.50

For this study, the coal conversion plant was sited in Texas at a minemouth location above the Wilcox lignite seam. This is shown below:



The Texas site was selected because it fulfilled all major requirements. It is roughly 340 miles from the West Texas oil fields, 20 miles from interstate natural gas pipelines that run to the U.S. East Coast and Midwest, and 180 miles from the existing U.S. gulf coast hydrogen pipeline. In addition, electric power is already produced at minemouth locations along the seam. The only drawback of Texas lignite relative to the other coals identified above is its relative higher price. Other locations such as North Dakota and Wyoming may have better economics but are not located in high hydrogen demand areas of the U.S. The analysis presented below for SNG would be generally applicable to these other locations; however, product yields and cost would be different.

Plant Designs

This analysis has investigated two overall conceptual configurations. The first of these uses Texas Lignite to produce electric power, SNG, and carbon dioxide in a polygeneration facility. The second polygeneration configuration uses the Texas lignite to produce electric power, hydrogen, and carbon dioxide.

Conversion of Texas lignite using gasification technology presents some challenges because of the high moisture content of the as-received (AR) coal. The AR lignite contains 30 weight percent moisture and 11.8 percent oxygen (see Table 1). Because of this the calorific value of the coal is low at 7868 Btu per pound on a high heating value (HHV) basis.

Seven gasification systems were examined to convert the lignite into synthesis gas. These are listed in Table 2. The single-stage slurry feed system with heat recovery represents a GE/Texaco type process. This gasifier type is in operation at the Polk Power Station in Florida. Feeding the coal to the gasifier in this system requires that the coal be slurried with water. Assuming that the slurry can contain 66 percent lignite by weight and, because the lignite already contains 30 percent moisture, the overall solids content of the slurry is only 46 percent by weight. This means that 54 weight percent of the input to the gasifier is water. This results in a total carbon content in the feed slurry of only 29 weight percent. Because of this low carbon content and high water content, the overall clean cold gas efficiency of the gasification system is only 60 percent on an HHV basis. This means that only 60 percent of the energy content of the input lignite resides in the clean synthesis gas. It is assumed that the capacity factor for this system would be 85 percent. That is, the gasifier is on stream producing synthesis gas for 310 days per year. Also, it is assumed that the carbon utilization is 95 percent. That is, 5 percent of the input carbon in the lignite resides in the slag.

Table 1. Wilcox Lignite Analysis

	AF	MAF	As Fed	MF	AR
Carbon	58.14	72.47	55.38	63.19	44.24
Hydrogen	4.89	6.09	4.65	5.31	3.72
Nitrogen	0.96	1.2	0.92	1.05	0.73
Sulfur	0.77	0.96	0.73	0.84	0.59
Oxygen	15.47	19.28	14.73	16.81	11.77
Mineral/Matter	11.78	–	11.22	12.80	8.96
Moisture	8.00	–	12.36	14.10	30
Total	**100.00**	**100**	**100.00**	**114.10**	**100.00**
HHVBTU#	10341	12890	9851	11240	7868

Table 2. Summary of Cases Analyzed

Gasifier Type Case Number	Single-Stage Slurry Feed Heat Recovery –	Single-Stage Slurry Feed Quench –	Two-Stage Slurry Feed Heat Recovery and Quench	Single-Stage Dry Feed Heat Recovery –	Single-Stage Dry Feed Quench	Single-Stage Dry Feed Quench Advanced
Hydrogen/Power/CO_2	√	√	√	√	√	√
SNG/Power/CO_2	√	√	√	√	√	√
Capacity Factor %	85	85	85	85	85	90
Carbon Utilization %	95	95	95	95	95	99
Percent Solids in Slurry	46	46	46	NA	NA	NA
Feed Coal Moisture Wt %	30	30	30	12	12	8
Cold Gas Efficiency (% HHV)	60	60	69	77.5	77.5	80

The single-stage slurry gasifier with quench represents a GE/Texaco system with full quench in place of a radiant gas cooler for heat recovery. Because the lignite is fed using a water slurry, the same issues pertain as in the prior case.

The two-stage slurry feed gasifier represents a ConocoPhillips E-Gas type system. This gasifier is operating at Wabash, Indiana. In this system the coal is injected using a water slurry into two stages of the gasifier. In the first stage the coal slurry is gasified with oxygen and the hot gases from this stage rapidly dry the second stage coal slurry. The unconverted char is then separated from the gasifier effluent and this dried char is recycled back to the first stage. Therefore this gasifier system can be thought of as being intermediate between a single stage slurry system and a dry feed system. This is exemplified by the much higher clean cold gas efficiency (69 percent versus 60 percent) of this system. A preliminary analysis of both a heat recovery and a quench version of this two-stage system was undertaken.

Three dry feed gasification systems were analyzed. The single-stage dry feed heat recovery case represents a Shell type system with a waste heat boiler. This system is operating at the NUON IGCC plant in the Netherlands. In this system the as-received lignite must be dried before feeding to the lock hoppers. If the lignite is not dried it will bridge and block the pressurized lock hoppers. In this analysis it is assumed that the lignite is dried from 30 to 12 weight percent moisture. The resulting carbon in the feed lignite is then 55.38 percent (see Table 1).

The single-stage dry feed quench system analyzed represents a Shell type gasifier but with the waste heat boiler section eliminated and replaced by full water quench of the gasifier effluent. This quench configuration is not commercially available but, because the system is much less expensive without the waste heat boiler it is assumed that it could be available if it proved suitable for certain applications. In these two dry feed Shell type gasifiers, cooled synthesis gas is recycled to the gasifier exit to cool the effluent synthesis gas to below the ash fusion temperature before the gas enters the waste heat boiler. In these dry feed systems the clean cold gas efficiency is increased to 77.5 percent on an HHV basis. It was assumed that the capacity factor remained at 85 percent and the carbon utilization remained at 95 percent.

The single-stage dry feed advanced quench gasification system analyzed in this study represents a GSP type gasifier. The GSP process was formerly known as the Babcock Borsig Power (BBP) Noell process. Future Energy GmbH acquired the intellectual property rights, the test plant facilities, real estate and buildings, and the entire patent stock from the insolvent BBP in December 2002. In Schwarze Pumpe Germany, the GSP process was used to gasifier lignite until 1991. This gasifier has a capacity of about 700 tons per day (130 MW thermal). Currently it is being used to gasify waste oils to produce synthesis gas for a methanol plant.

The GSP gasifier is a dry feed system. It is an oxygen-blown entrained gasifier with a so called “cooling screen” wall. This concept is similar to that of the Shell process where a membrane wall with pressurized water or steam is used to cool the gasifier inside surface so that a constantly forming liquid slag layer forms the refractory lining. This is different from the other gasifier systems like GE/Texaco and E-Gas where a brick refractory lining is used. In this case the lignite feed is dried to 8 weight percent moisture before being sent to the pressurized lock hoppers. The resulting carbon content of the feed is then 58.14 percent. The clean cold gas efficiency in this case is 80 percent. It is assumed in this case that the capacity factor has increased to 90 percent and the carbon utilization has increased to 99 percent.

After an initial screening study of the various gasification systems it was concluded that, because of the low cold gas efficiency of the single-stage slurry feed gasifiers when used to gasify the lignite, they were not suitable for processing this coal. However, the slurry feed gasifiers could be suitable for gasifying high moisture low rank coals if it were possible to remove the inherent moisture by drying and then to slurry the dried coal with water and feed this slurry to the gasifier before the coal could reabsorb the moisture. Because of this uncertainty, only six cases were analyzed in detail and these are shown in Table 3. The two-stage slurry quench (E-Gas type), the single-stage dry feed quench (Shell type), and the single-stage advanced dry feed quench (GSP type) gasifiers were analyzed in the two configurations to produce either SNG or hydrogen, along with power and carbon dioxide.

Production of SNG, Carbon Dioxide, and Power

Figure 1 shows the schematic of the three cases that convert the lignite into SNG, power, and carbon dioxide. In the case using the two-stage slurry gasifier with quench (Case 1S) the as-received lignite is slurried with water and fed with oxygen into the gasifier. In the two cases where the dry feed gasifiers are used (Cases 2S and 3S) the lignite is dried under nitrogen using some of the fuel gas and the dried lignite is pneumatically conveyed to the gasifier using either nitrogen or carbon dioxide. The raw synthesis gas after water quench is sent to a raw gas shift unit where the hydrogen to carbon monoxide molar ratio is adjusted to three to one to be compatible with methanation. The shift effluent is cooled and passed to the activated carbon reactor to remove mercury. The synthesis gas is then sent to sulfur removal where a concentrated stream of hydrogen sulfide is produced. This is sent to a Claus SCOT combination for sulfur recovery. After sulfur removal the gas is sent to a bulk carbon dioxide removal system. The recovered carbon dioxide is then dehydrated and compressed

Table 3. Cases Analyzed

Case	Description
1 S	Two-stage slurry quench-SNG/Power/CO_2
2 S	Single-Stage dry quench-SNG/Power/CO_2
3 S	Advanced single-stage dry quench-SNG/Power/CO_2
1 H	Two-stage slurry quench H_2/Power/CO_2
2 H	Single-stage dry quench H_2/Power/CO_2
3 H	Advanced single-stage dry quench-H_2/Power/CO_2

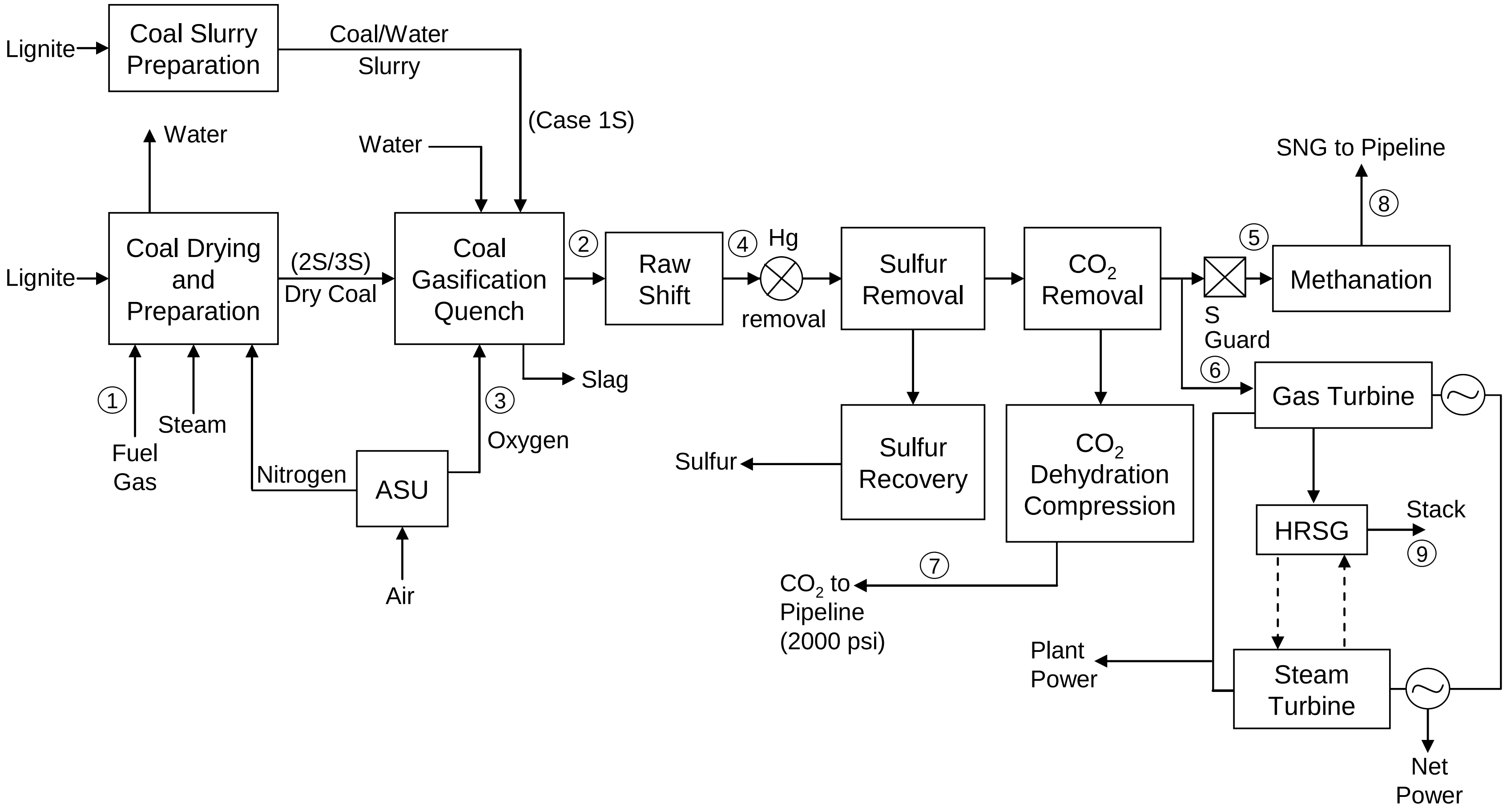


Figure 1. Texas Lignite to Power and SNG

to 2000 psi and sent to a pipeline. To protect the methanation catalyst, the synthesis gas with sulfur and carbon dioxide removed is sent to a sulfur polishing reactor to remove the last traces of hydrogen sulfide before being sent to the three stage methanation reactor. In this reactor the carbon monoxide and hydrogen are combined to produce methane. The SNG from the methanation reactor, at about 20 bars, is then sent to battery limits to be compressed for delivery to the natural gas pipeline. Some of the synthesis gas exiting the bulk carbon dioxide removal system is sent to a gas turbine where electric power is generated. The hot effluent from the gas turbine is sent to a heat recovery seam generator (HRSG) where the high pressure steam is used in a steam turbine to generate additional electric power. Some of this power is used in the plant and the net power is sold.

Table 4 summarizes the results of the three SNG cases. In Case 1S, the synthesis gas is produced using a two-stage slurry feed quench gasification system. Feed rate is 7,500 TPD of as-received lignite containing 30 weight percent moisture. The products from this plant are 32 MMSCFD of SNG, 7,753 TPD of carbon dioxide, and 255 MW of net electric power. The overall HHV efficiency of the process from lignite to products is 44.6 percent. The capital cost of the plant is estimated to be $760 MM (see Table 5 for the capital cost breakdown). The operating and maintenance cost, less the lignite, is estimated to be $39 MM per annum and the lignite cost is $33 MM per annum. The lignite is assumed to cost $14 per ton on an as-received basis. The assumed capacity factor for the plant is 85 percent. The required selling price (RSP) of the SNG was calculated using a discounted cash flow (DCF) analysis with the financial parameters shown in Table 6. Because three products are produced from the plant, power, carbon dioxide and SNG, it was necessary to fix the value of two of the products and calculate the RSP of the third. It is assumed that the value of the electric power is $35.6 per MWH and the value of the carbon dioxide is $12 per ton. With these values and the financial parameters assumed in the DCF analysis, the RSP of the SNG for this case is calculated to be $6.90/MMBTU.

In Case 2S, the synthesis gas is produced using a single-stage dry feed quench gasification system. The lignite dried to 12 weight percent moisture is conveyed to the high pressure gasifier using carbon dioxide as carrier gas so that nitrogen will not dilute the SNG product. The dried lignite feed rate is 5,990 ton per day (TPD). The products from this plant are 34 MMSCFD of SNG, 7,418 TPD of carbon dioxide, and 236 MW of net electric power. The assumed capacity factor for the plant is 85 percent. The overall HHV efficiency of the process from lignite to products is 45.3 percent. The capital cost of the plant is estimated to be $743 MM (see Table 5 for the capital cost breakdown). The operating and maintenance cost, less the lignite, is estimated to be $38 MM per annum and the lignite cost is $33 MM per annum. Because the lignite is dried to 12 percent moisture the cost of the lignite on an as-fed basis is now $17.53 per ton. The required selling price (RSP) of the SNG was calculated using a discounted cash flow (DCF) analysis with the financial parameters shown in Table 6. As in Case 1S, it is assumed that the value of the electric power is $35.6 per

Table 4. Summary of Results: SNG and Power

	Two-Stage/Slurry/Q Case 1S	**Single-Stage/Dry/Q Case 2S**	**Single-Stage/Dry/Q/Advanced Case 3S**
Coal Input (TPD AF)	7500	5990	5707
SNG Out (MMSCFD)	32	34	39
Power Net (MW)	255	236	244
Overall Efficiency (% HHV)	44.6	45.3	49.4
Capital ($ MM)	760	743	734
O & M ($ MM)	39	38	38
Coal Cost (MM$/Yr) ($/T)AF	33/14	33/17.53	35/18.4
CO_2 Value ($/T)	12	12	12
CO_2 Captured (TPD)	7753	7418	7724
Power Value ($/MWH)	35.6	35.6	35.6
RSP SNG ($/MMBtu) (HHV)	6.90	6.73	5.00

Table 5. Capital Cost Summary: SNG and Power

	$MM (2004)		
	Case 1S	**Case 2S**	**Case 3S**
Coal Handling/Drying	28	47	45
Gasification	102	87	70
Air Separation	83	75	76
Sulfur Removal/Recovery	22	22	23
Shift	22	20	20
CO_2 Removal/Compression	43	47	46
Methanation	31	33	36
Power Generation/Distribution	191	182	189
WW Treatment	13	13	13
Balance of Plant	44	40	41
Total Installed Cost	**579**	**566**	**559**
Home Office (8.4%)	49	48	47
Fee (2%)	11	11	11
Contingency (15%)	96	94	93
Total Plant Cost	**735**	**719**	**710**
Total Capital (Inc. ND Capital)	**760**	**743**	**734**

Table 6. Discounted Cash Flow Analyses Assumptions

Initial Plant Output 50% (Year 1) 90% (Year 2)
Debt: Equity = 67:33
Required Selling Price (RSP) in constant dollars necessary for 15% ROE (Current $)
Debt: 16 years @ 8% interest
General inflation 3%
Escalation in accordance with EIA projects
Depreciation 16 years with double declining balance
Federal and state income tax (Fed 34%) (State 6%)
Local tax and insurance 2% of depreciable capital
Project life 25 years

MWH and the value of the carbon dioxide is $12 per ton. With these values and the financial parameters assumed in the DCF analysis, the RSP of the SNG for this case is calculated to be $6.73/MMBTU.

In Case 3S, the synthesis gas is produced using a single-stage advanced dry feed quench gasification system. The stream numbers on Figure 1 refer to the material balance for this case. The material balance is shown for selected stream flows for this case and is summarized in Table 7. Again the coal is conveyed into the gasifier using carbon dioxide and the feed rate is 5,707 TPD of dried lignite containing 8 weight percent moisture. The products from this plant are 39 MMSCFD of SNG, 7,724 TPD of carbon dioxide, and 244 MW of net electric power. The assumed capacity factor for the plant is 90 percent and the carbon utilization is assumed to be 99 percent. The overall HHV efficiency of the process from lignite to products is 49.4 percent. The capital cost of the plant is estimated to be $734 MM (see Table 5 for the capital cost breakdown). The operating and maintenance cost, less the lignite, is estimated to be $38 MM per annum and the lignite cost is $35 MM per annum. Because the lignite is dried to 8 percent moisture the cost of the lignite on an as-fed basis is now $18.40 per ton. The required selling price (RSP) of the SNG was calculated using a discounted cash flow (DCF) analysis with the financial parameters shown in Table 6.

As in Case 1S, it is assumed that the value of the electric power is $35.6 per MWH and the value of the carbon dioxide is $12 per ton. With these values and the financial parameters assumed in the DCF analysis, the RSP of the SNG for this case is calculated to be $5.00/MMBTU.

Production of Hydrogen, Carbon Dioxide and Power

Figure 2 shows the schematic of the three cases that convert the lignite into hydrogen, power, and carbon dioxide. In the case using the two-stage slurry gasifier with quench (Case 1H), the as-received lignite is slurried with water and fed with oxygen into the gasifier. In the two cases where the dry feed gasifiers are used (Cases 2H and 3H), the lignite is dried under

Table 7. Power and SNG Production Noel Quench Gasifier (Case 3S)

Selected Flows, Pound Moles/Hour									
	1	2	3	4	5	6	7	8	9
	Fuel Gas To Drying	Quenched Output	Oxygen To Gasifier	Shifted Gas	Clean Gas To Methatn	Turbine Fuel	CO2 To Pipeline	SNG To Pipeline	Stack From HRSG
CH4	0.1	2		2	1	0.8		4,076	0
H20	0	43,842		0	0	0		0	10,218
H2	1,135	9,962		24,019	12,427	10,217		443	0
CO	380	22,064		8,006	4,159	3,419		2	0
CO2	8	2,044		16,101	89	73	14,629	212	3,493
N2	11	240	101		125	102		126	94,850
H2S		114		114					0
NH3									0
O2			10,138						12,818
Total	1,535	78,267	10,239	48,243	16,801	13,813	14,629	4,859	121,380
T, Deg F	309	400	555	110	309	309		488	260
P, atm	1.0	30.0	30.6	27.1	25.0	34.7	136	20.0	1.00

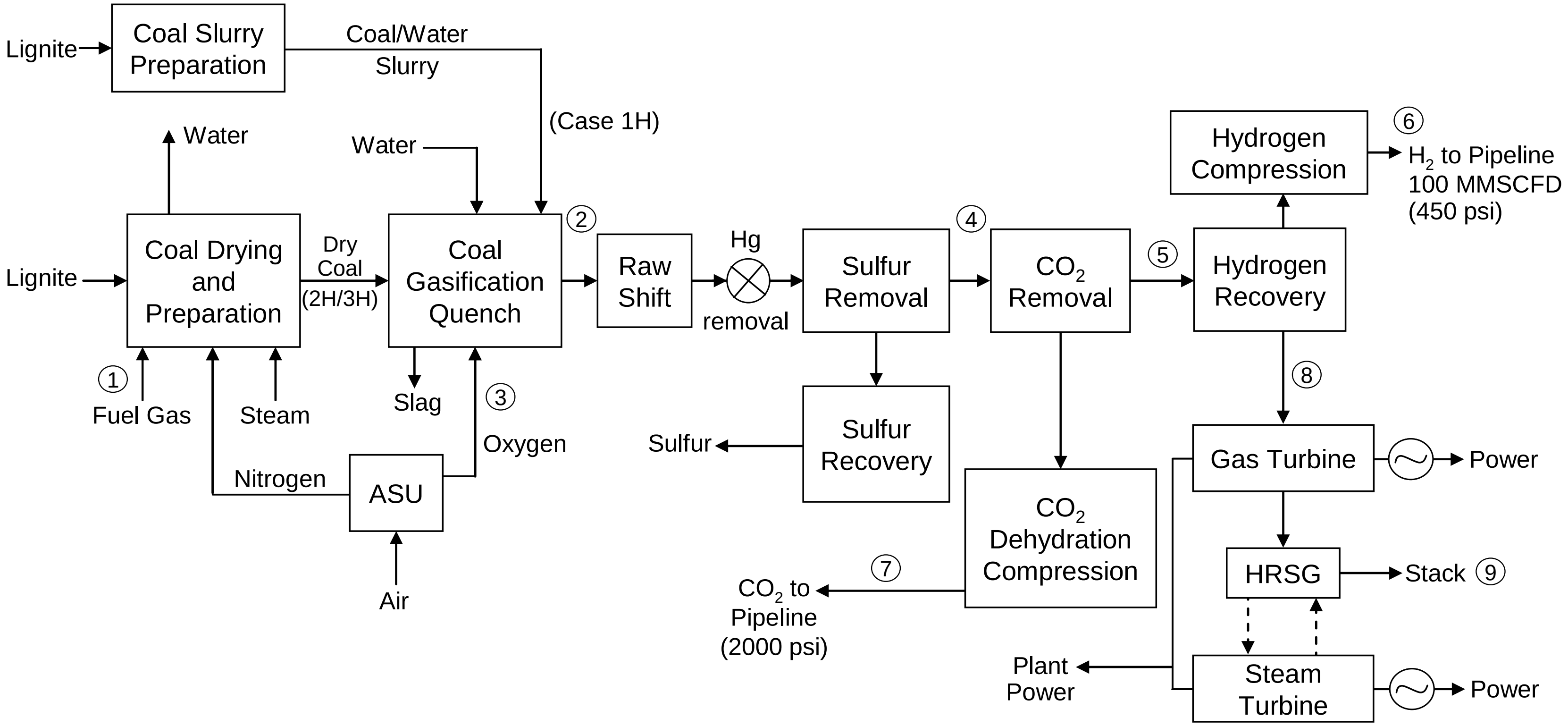


Figure 2. Lignite to Electric Power and Hydrogen

nitrogen using some of the fuel gas and the dried lignite is pneumatically conveyed to the gasifier using nitrogen. The raw synthesis gas after water quench or scrub is sent to a raw gas shift unit where much of the carbon monoxide is converted into hydrogen. The shift effluent is cooled and passed to the activated carbon reactor to remove mercury. The synthesis gas is then sent to sulfur removal where a concentrated stream of hydrogen sulfide is produced. This is sent to a Claus SCOT combination for sulfur recovery. After sulfur removal, the gas is sent to a bulk carbon dioxide removal system. The recovered carbon dioxide is then dehydrated and compressed to 2000 psi and sent to a pipeline. The synthesis gas with sulfur and carbon dioxide removed is sent to a polymer membrane separation system followed by a PSA unit where the required amount of hydrogen is removed. A membrane system is used in this case to maintain the system pressure of the remaining synthesis gas that will be used for power generation. This synthesis gas is sent to a gas turbine where electric power is generated. The hot effluent from the gas turbine is sent to a heat recovery seam generator (HRSG) where the high pressure steam is used in a steam turbine to generate additional electric power. Some of this power is used in the plant and the net power is sold.

Table 8 summarizes the results for the three hydrogen production cases. In Case 1H, the synthesis gas is produced using a two-stage slurry feed quench gasification system. Feed rate is 6,852 TPD of as-received lignite containing 30 weight percent moisture. The products from this plant are 100 MMSCFD of hydrogen, 9,202 TPD of carbon dioxide, and 224 MW of net electric power. The overall HHV efficiency of the process from lignite to products is 46.5 percent. The capital cost of the plant is estimated to be $709 MM (see Table 9 for the capital cost breakdown). The operating and maintenance cost, less the lignite, is estimated to be $36 MM per annum and the lignite cost is $30 MM per annum. The lignite is assumed to cost $14 per ton on an as-received basis. The assumed capacity factor for the plant is 85 percent. The RSP of the hydrogen was calculated using a DCF analysis with the financial parameters shown in Table 6. As in the previous cases, it is assumed that the value of the electric power is $35.6 per MWH and the value of the carbon dioxide is $12 per ton. With these values and the financial parameters assumed in the DCF analysis, the RSP of the hydrogen for this case is calculated to be $5.94/MMBTU or $0.80 per kilogram.

In Case 2H, the synthesis gas is produced using a single-stage dry feed quench gasification system similar to Case 2S except that nitrogen is used to convey the coal to the gasifier. Feed rate is 5,158 TPD of dried lignite containing 12-weight percent moisture. The products from this plant are 100 MMSCFD of hydrogen, 8,468 TPD of carbon dioxide, and 189 MW of net electric power. The assumed capacity factor for the plant is 85 percent. The overall HHV efficiency of the process from lignite to products is 47.1 percent. The capital cost of the plant is estimated to be $666 MM (see Table 9 for the capital cost breakdown). The operating and maintenance cost, less the lignite, is estimated to be $34 MM per annum and the lignite cost is $28 MM per annum. Because the lignite is dried to 12-percent moisture the cost of the lignite on an as-fed basis is now $17.53 per ton. The RSP of the hydrogen was calculated using a DCF analysis with the financial parameters shown in Table 6. As in Case 1H, it is assumed that the value of the electric power is $35.6 per MWH and the value

Table 8. Summary of Results: Hydrogen and Power

	Two-Stage/ Slurry/Q Case Number 1H	**Single-Stage/ Dry/Q Case Number 2H**	**Single-Stage/ Dry/Q/Advanced Case Number 3H**
Coal Input (TPD AF)	6852	5158	4665
Hydrogen Out (MMSCFD)	100	100	100
Power Net (MW)	224	189	193
Overall Efficiency (% HHV)	46.5	47.1	49.2
Capital ($ MM)	709	666	650
O & M ($ MM/Yr) (less coal)	36	34	34
Coal Cost ($MM/YR)/($/T/AF)	30/14	28/17.53	28/18.4
CO_2 Value ($/T)	12	12	12
CO_2 Captured (TPD)	9202	8468	8279
Power Value ($/MWH)	35.6	35.6	35.6
RSP Hydrogen ($/MMBtu) (HHV)	5.94	6.25	5.20
RSP Hydrogen ($/kg)	0.80	0.84	0.70

of the carbon dioxide is $12 per ton. With these values and the financial parameters assumed in the DCF analysis, the RSP of the hydrogen for this case is calculated to be $6.25/MMBTU or $0.84 per kilogram.

Table 9. Capital Cost Summary: Hydrogen and Power

	($ MM 2004)		
	Case 1H	**Case 2H**	**Case 3H**
Coal Handling/Drying	27	50	47
Gasification	93	75	70
Air Separation	77	65	65
Sulfur Removal/Recovery	22	21	21
Shift	20	18	18
CO_2 Removal/Compression	50	49	46
Hydrogen Recovery	16	16	16
Power Generation/Distribution	182	164	164
WW Treatment	14	14	14
Balance of Plant	39	36	34
Total Installed Cost	**540**	**508**	**495**
Home Office (8.4%)	45	43	42
Fee (2%)	11	10	10
Contingency (15%)	89	84	82
Total Plant Cost	**685**	**645**	**629**
Total Capital (Inc. ND Capital)	**709**	**666**	**650**

In Case 3H, the synthesis gas is produced using a single-stage advanced dry feed quench gasification system. The stream numbers on Figure 2 refer to the material balance for this case. The material balance is shown for selected stream flows for this case and is summarized in Table 10. Feed rate is 4,665 TPD of dried lignite containing 8 weight percent moisture. The products from this plant are 100 MMSCFD of hydrogen, 8,279 TPD of carbon dioxide, and 193 MW of net electric power. The assumed capacity factor for the plant is 90 percent and the carbon utilization was assumed to be 99 percent. The overall HHV efficiency of the process from lignite to products is 49.2 percent. The capital cost of the plant is estimated to be $650 MM (see Table 9 for the capital cost breakdown). The operating and maintenance cost, less the lignite, is estimated to be $34 MM per annum and the lignite cost is $28 MM per annum. Because the lignite is dried to 8 percent moisture the cost of the lignite on an as-fed basis is now $18.40 per ton. The RSP of the hydrogen was

Table 10. Power and Hydrogen Production Noel Quench Gasifier (Case 3H)

Selected Flows, Pound Moles Per Hour									
	1 Fuel Gas To Drying	2 Quenched Output	3 Oxygen To Gasifier	4 Shifted Gas	5 Clean Gas To H2 Rec	6 Recovered Hydrogen	7 CO2 To Pipeline	8 Turbine Fuel	9 Stack From HRSG
CH_4	0.2	2		2	2			2	0
H_2O	7	36,368		0	67			60	11,220
H_2	1,239	8,914		23,531	23,414	11,019		11,155	0
CO	294	17,559		2,941	2,938			2,645	0
CO_2	14	1,203		15,820	140		15,679	126	2,773
O_2			8,376		0			0	13,499
N_2	220	2,200	84	2,200	2,200			1,980	96,116
H_2S		94		94					0
NH_3									0
Total	1,774	66,340	8,460	44,589	28,762	11,019	15,679	15,969	123,608
T, Deg F	309	398	555	110	85	282		100	260
P, atm	1.0	30.0	30.6	27.1	27.1	30.6	136	1.0	1.0

calculated using a DCF analysis with the financial parameters shown in Table 6. As in Case 1H, it is assumed that the value of the electric power is $35.6 per MWH and the value of the carbon dioxide is $12 per ton. With these values and the financial parameters assumed in the DCF analysis, the RSP of the hydrogen for this case is calculated to be $5.20/MMBTU or $0.70 per kilogram.

Hydrogen Cost Compared to Natural Gas Costs

Figure 3 shows the impact of natural gas prices on the resulting cost of hydrogen from steam methane reforming. If the feed natural gas price to the steam methane reformer (SMR) is $3.00/MMBTU then the resulting cost of the hydrogen produced would be about $5.50/MMBTU or $0.74/KG. If the natural gas price was $4.00/MMBTU then the resulting cost of the hydrogen from the SMR would be about $6.75/MMBTU (HHV basis) or $0.91/KG. This analysis has estimated that the cost of producing hydrogen from Texas lignite is in the range of $5.20-$6.25/MMBTU. This is assuming that the plants are configured as described to coproduce electric power and carbon dioxide for sales. These costs of hydrogen would be equivalent to using steam methane reforming for natural gas prices of between $3 and $4.00/MMBTU. EIA is forecasting that natural gas prices will be in the range $4.50-$5.00/MMBTU by 2025 but spot natural gas process for 2004 have been much higher in the range $5.75 to $6.50/MMBTU with expectations that they may well rise to above $7.00/MMBTU during 2005.

This indicates that producing hydrogen from Texas lignite at the mine mouth and pipelining this hydrogen to Gulf Coast refineries could be an attractive proposition.

Impact of the value of the Carbon Dioxide

In this study it was assumed that the carbon dioxide was valued at $12 per ton. This is equivalent to $0.70 per thousand cubic feet about midway in the usual range of carbon dioxide prices for EOR. If the carbon dioxide could only command a lower value, then the required selling prices of the other co products would have to be increased to realize the annual revenue requirement for the plant. However, it takes capital and energy to capture, dehydrate, and compress the carbon dioxide. In the configuration that produces power, SNG, and carbon dioxide, it would not be economic to produce carbon dioxide as a co product for sale if the value of the carbon dioxide were less than $5.50 per ton. Similarly, for the plant producing power, hydrogen and carbon dioxide, it would not be economic to produce carbon dioxide as a co product for sale if the value of the carbon dioxide were less than $7.75 per ton. Therefore, a value of $12 per ton makes it worthwhile to recover the carbon dioxide for sale.

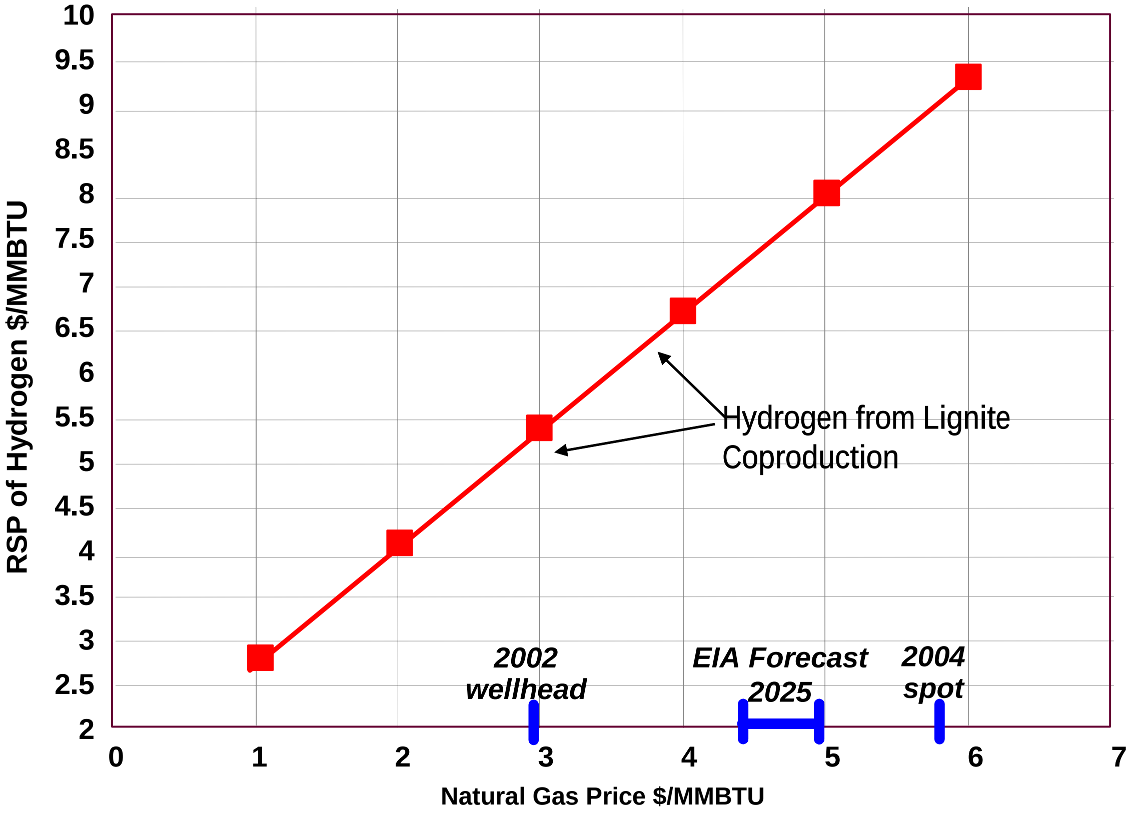


Figure 3. Hydrogen Cost versus Natural Gas Price from Steam Methane Reforming

Conclusions

This feasibility study has shown that siting a mine mouth Lignite fed gasification plant in Texas to produce hydrogen, SNG, electric power, and carbon dioxide could be economically feasible in an era of high natural gas prices.

Because of the high moisture content of the lignite the choice of gasification system becomes an important issue. If the as-received lignite is used directly, systems that use water slurry to feed the coal into the gasifier are penalized because of the very low carbon content of the resulting slurry. It may be possible to dry the lignite to a low moisture content, maybe 10 percent moisture, and then to slurry the dried lignite with water and feed the gasifier. Because most of this water is "inherent" (nonsurface water contained in the coal structure) the lignite may reabsorb this water slowly enough to allow the resulting slurry to have a much higher carbon content than if as-received lignite was used. Dry feed gasifier systems are preferred when processing lignite. However, even then, the lignite must be dried to prevent blockage in the pressurized lock hoppers. Just how dry the lignite must be is uncertain but in this analysis it was assumed that 8 to12 weight percent moisture was necessary.

Hydrogen produced from Texas lignite in a coproduction plant could be produced in the range \$5.20-\$6.20/MMBTU (HHV basis) equivalent to between \$0.70 and \$0.84 per kilogram. The actual cost depends on the gasification system and the values of the co produced electric power and carbon dioxide. This range of hydrogen costs is equivalent to hydrogen produced by steam methane reforming of natural gas if the natural gas feed price was between \$3.00 and \$4.00/MMBTU. With natural gas prices continuing to remain above \$5.00/MMBTU this concept of using Texas lignite for hydrogen production would be economically viable.

For the production of SNG form Texas lignite, the costs range from \$6.90-\$5.00/MMBTU (HHV basis). This depends on the gasification system, the value of coproduced power, and the value of the carbon dioxide. If natural gas prices remain above \$5.00/MMBTU then the configuration using the advanced dry feed gasification system would be economically viable for production of SNG. This option may be even more attractive with other low rank coals such as Wyoming subbituminous and North Dakota lignite coals that are priced lower than Texas lignite.

Production of electric power from these conceptual coproduction plants provides a valuable revenue stream. Net power to sales averaged around 240 MW. It was assumed that these plants would be base load and that the value of the electricity was \$35.6/MWH.

The opportunity to sell carbon dioxide for EOR in Texas provided another valuable revenue stream for the plants. The break even cost of recovering the carbon dioxide ranged from about $5.50 to $7.75 per ton depending on whether SNG or hydrogen was the product. In this analysis it was assumed that the value of the carbon dioxide was $12 per ton; therefore, at this value recovery was a profitable venture. With the worldwide movement towards regulation of carbon dioxide and greenhouse gases (GHG), these plants would qualify for future verifiable carbon dioxide emissions trading credits. The Chicago Climate Exchange (CCX) launched a GHG program in September 2003 and many companies signed up on a voluntary basis for trading. Carbon dioxide is trading for about $1 per ton. In Europe on 1 January 2005, the European Union Emissions Trading Program begins. Therefore, plants that are already configured for carbon dioxide recovery will have an advantage in the future.