Storing syngas lowers the carbon price for profitable coal gasification

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Integrated gasification combined cycle (IGCC) electric power generation systems with carbon capture and sequestration have desirable environmental qualities, but are not profitable when the carbon dioxide price is less than approximately $50 per metric ton. We examine whether an IGCC facility that operates its gasifier continuously but stores the syngas and produces electricity only when daily prices are high may be profitable at significantly lower CO_2 prices. Using a probabilistic analysis, we have calculated the plant-level return on investment (ROI) and the value of syngas storage for IGCC facilities located in the US Midwest using a range of storage configurations. Adding a second turbine to use the stored syngas to generate electricity at peak hours and implementing 12 hours of above ground high pressure syngas storage significantly increases the ROI and net present value. Storage lowers the carbon price at which IGCC enters the US generation mix by approximately 25%.

Introduction

Producing electricity from coal-derived synthesis gas (syngas) in an integrated gasification combined cycle (IGCC) facility can improve criteria pollutant performance over other coal-fueled technologies such as pulverized coal (PC) facilities [1-5] and can be implemented with carbon capture and sequestration.

Previous studies have shown that IGCC with carbon capture and sequestration (CCS) has the potential for CO_2 control at costs comparable to those of other low-carbon generation technologies [4-6]. Using a water gas shift and Selexol process [7], IGCC facilities can achieve 85 to 90 percent CO_2 reduction with emission rates near 95 kg CO_2/MWh [5].

Adding CO_2 capture and storage incurs an energy penalty, estimated by previous work as 14 percent for an IGCC and 24 percent for a PC plant [5]. CCS has been calculated to increase the cost of delivered electricity by 44 percent for IGCC facilities and 78 percent for PC [5].

In addition to lowering CO_2 emissions, IGCC facilities with CCS have increased environmental advantages over traditional coal combustion technologies because of lower levels of criteria pollutant emissions, reduced water usage and lower amounts of solid waste. Criteria emissions control with IGCC+CCS is cost effective because most clean-up occurs in the syngas, that has higher pressure, lower mass flow and higher pollution concentration than stack exhaust gases [3, 8, 9].

Particulate emissions from existing IGCC units are below 0.001 lb/million Btu versus about 0.015 lb/mmBtu from modern PC units [8]. Current NOx emissions at IGCC facilities are 0.06-0.09 lb/mmBtu versus 0.09-0.13 lb/mmBtu for PC facilities. Further reductions of NOx emissions can be achieved at IGCC facilities with SCR: 0.01 lb NOx/mmBtu has been demonstrated commercially in an IGCC unit in Japan [8]. IGCC mercury removal is in the range of 95-99 percent, versus 85-95 percent mercury removal for PC facilities using advanced control

[2]. IGCC facilities have SO_2 emissions of 0.015-0.08 lb/mmBtu compared to 0.08-0.23 lb SO_2/mmBtu for PC, depending on the age of the facility and type of coal [8]. Additionally, IGCC facilities have lower emissions of other byproducts such as chloride, fluoride, cyanide than PC facilities [1], and IGCC uses 20-35 percent less water than PC [8]. For the same coal feed, an IGCC produces 40-50 percent less solid waste than a PC and the fused slag can be more easily disposed of than can fly ash [8].

Although IGCC is generally preferred on an environmental basis, in the absence of carbon pricing mechanisms PC facilities are capable of producing electricity at a significantly lower cost than IGCC facilities. A high carbon price is required in order for IGCC facilities to be economically competitive with PC facilities.

There are currently eight gasification facilities operating worldwide producing about 1.7 GW of electricity from coal or petcoke feedstock [10], and in all of these facilities the syngas is used immediately after it is produced. Without storage capabilities, the gasifier must be sized to fit the syngas end-use (such as a gas turbine or chemicals process) and the operation of the two systems must be coupled. For IGCC designs where the air separation unit is not fully integrated with the turbine [11, 12], adding the capability to store syngas decouples the gasifier from the turbine, allowing the gasifier and turbine to be sized and operated independently, thereby providing valuable flexibility in the way the facility is configured and operated. One example is using syngas storage to generate peak electricity. Syngas storage provides a means to continuously operate the gasifier at the most efficient sustained production rate, but to sell electricity only when daily electricity prices are highest, thereby maximizing profits and enhancing plant-level economics over a non-storage IGCC facility while operating the gasifier at the same capacity factor. When used in this manner, diurnal syngas storage at an IGCC facility can increase profits and return on investment and lower the carbon price at which IGCC enters the US generation mix. Here we examine the value of implementing diurnal storage to produce peak power from an IGCC facility.

Engineering-economic model

We modeled a non-storage (baseline) IGCC facility producing 270 net MW from one gasifier. Although facilities such as the Wabash River IGCC in Indiana operate with a spare gasifier (we term this 1+1), Wabash River was built as a government demonstration project, and new commercial plants are likely to be constructed with no spare (1+0). This baseline facility represents the lowest capital cost IGCC facility that would reasonably be built and operated [13] and has a capital cost of $415 million or $1,540/kW (Table 1). While this capital cost does not reflect recent increases, our storage results are not invalidated at higher capital costs. We also examined facilities with configurations of 1+1, 3+1 and 3+0 and reached similar conclusions as for the 1+0 analysis; due to economies of scale, syngas storage adds greater value to larger sized facilities including those constructed with spare gasifiers despite increased capital costs (see online supporting information).

Syngas storage systems were analyzed using the same gasifier size and configuration as the baseline scenario with the addition of a syngas storage process block and additional peaking turbine (Figure 1). The syngas storage block consists of the equipment necessary to accept syngas from the gasifier, store the volume of syngas in a vessel at a given pressure, and then supply syngas at the mass flow rate and pressure required by the peaking turbine. Here, the design of the syngas storage scenario is conceptual and likely not optimal, and is provided to outline the potential benefits of such a system and to consider whether syngas storage can increase the profitability of an IGCC facility. The supporting information contain a description of the compression and storage process block used in the analysis and outline technical considerations that may need to be addressed during the design process to successfully implement such a system in a real-world application. Storage vessels considered were low pressure gasometers, high pressure cylindrical bullets and gas spheres, excavated rock caverns and salt caverns.

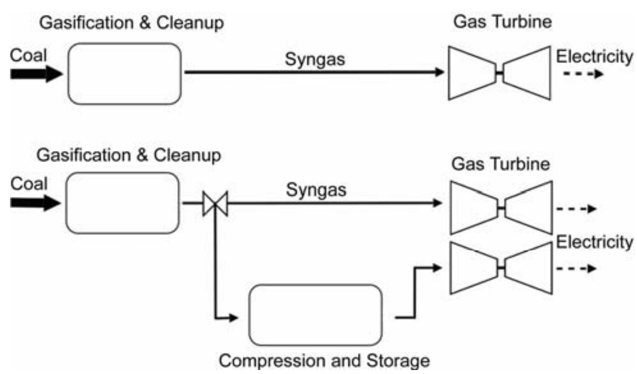

Figure 1. Baseline facility (top) and syngas storage scenario (bottom)

We calculated the return on investment (ROI) and net present value (NPV) for the baseline and syngas storage scenarios using both historical and forecasted prices for inputs (coal) and outputs (electricity) for a hypothetical IGCC facility located in the US Midwest.

The sensitivity of the ROI in each scenario to uncertainty and variability in design parameters, costs and prices was examined probabilistically. The value of adding diurnal syngas storage to produce peak electricity was quantified by comparing the ROI to that of a baseline IGCC facility producing electricity from syngas with no storage capabilities. The ROI for an IGCC facility with carbon capture and sequestration and syngas storage capabilities was calculated under a range of possible future carbon prices and compared to that of a baseline facility with no storage to quantify how storage affects the carbon price at which IGCC enters the US generation mix.

Capital and operating cost distributions for the gasification, cleanup and power block sections in the baseline facility are based on the Integrated Environmental Control Model (IECM) version cs 5.21 [14]. The baseline facility includes the process blocks shown in Table 1 (a more complete list of the processes and parameters is included in the supporting information).

Table 1. Baseline 270 MW_e net Facility Configuration and Parameters [14]

Process Block (mean capital cost $2005)	Components	Size / Description
Gasifier ($138.5M)	1 train GE/Texaco gasifier 0 spare train gasifier Coal handling Low temperature gas cooling Process condensate treatment	260 tons/hr syngas output
Air Separation Unit ($93.5M)	1 train	max output: 11,350 lb-mol/hr
Cold-gas Cleanup ($32.5M)	Hydrolyzer Selexol Claus plant Beavon-Stretford tail gas plant	98.5% efficiency 98% H_2S efficiency 95% efficiency 99% efficiency
Power Block ($150.8M)	Gas combustion turbine Heat recovery steam generator Steam turbine HRSG feedwater system	GE 7FA CCGT 510 MW (gross) combined cycle/turbine 9000 Btu/kWh
Fuel	Illinois #6 coal	HHV: 10,900 Btu/lb

Point estimates from IECM were converted into triangular distributions using assumptions of $\pm$5 percent, following capital cost estimates reported in the literature [15-17]. The distributions, rather than point estimates, were used as inputs into the engineering-economic models. Cost data from IECM are in 2005 constant dollars.

The syngas produced by the gasification process is composed primarily of carbon monoxide and hydrogen and is characterized by a low energy density, typically ranging from 150-280 Btu/scf. Because of the lower energy density, larger volumes of syngas than of natural gas are required to produce electricity in a gas turbine. Syngas storage vessels thus need to be large, have high working pressures, or have these in combination. Although hydrogen is known to embrittle metals, the concentrations and partial pressures of hydrogen typically found in syngas do not appear to require any special preventative measures [18-22] for syngas storage options used in this analysis. An additional potential problem resulting from the hydrogen content of syngas is that atomic hydrogen can diffuse through most metals [23]. However industrial experience with syngas and analogies with other industrial practices suggests that excessive diffusion and leakage of syngas through a storage chamber wall is not an issue for diurnal and relatively short-term storage [24].

We restrict consideration of storage options to compressed gas technology since it is the most relevant large-scale stationary storage method for syngas production facilities and is less expensive than alternatives such as liquefaction. Compressed gas storage is the simplest storage solution, as the only required equipment is a compressor and a pressure vessel [25]. Operating parameters, capital and operating costs were examined for compressors and different storage vessels including high pressure spheres and cylindrical ‘bullets’ common for liquefied propane and compressed natural gas storage, low pressure gasometers, underground salt caverns and excavated rock caverns.

Capital costs for compressors, which are required for all storage options, were obtained from the literature [23, 25, 26], and cost distributions were constructed from these data. Compressor capital costs were found to scale linearly with the size of the compressor. The distribution of the capital cost for a given size compressor, reflecting the range of cost uncertainty, was used as an input to the engineering economic models when compression was required.

Capital costs for storage vessels were compiled from the literature and from industry professionals (the supporting information contain physical details, capital costs and cost distribution calculations for storage vessels). From a regression analysis and prediction interval derived from these data, cost distributions were constructed and used as inputs in the model. The capital cost distributions suggest a salt cavern is preferred if it is available because it is the lowest cost. However, because salt and rock caverns are geographically sparse [27], this analysis considers the general case where neither is available.

We modeled an IGCC plant located in the US Midwest, using prices for Illinois number 6 coal (HHV 11,350 Btu/lb, sulfur content of 3.2 percent by weight [28]). We used both historic coal data and price forecasts from the Energy Information Administration (EIA) to account for the variability in coal prices (supporting information discuss the price distributions considered in the analysis). We modified the EIA Annual Energy Outlook (AEO) forecasts for year 2007 coal prices with a factor to account for EIA's historical error in forecasting price data [29-31] (see supporting information for details). The 2005-2006 coal prices have a mean of \$1.51/MMBtu and standard deviation of \$0.1. The 2007 EIA forecast including the historical accuracy factor has a mean value of \$1.73/MMBtu, 15 percent higher than the mean historical 2005-06 prices.

To estimate revenue, we obtained historical locational marginal price (LMP) data for electricity from September 1, 2005 to September 1, 2006 for nodes in the Midwest ISO region [32].

We examined syngas storage scenarios with 4, 8 and 12 hours storage. Storage size (measured in hours) and compressor size were selected to accommodate 100 percent of the output of the gasifier for the number of hours indicated (that is also the period the peaking turbine can generate electricity from stored syngas). We fixed syngas storage pressure at 63 bar for all storage scenarios, requiring a 5,600 kW compressor for both charging and discharging the storage vessel. This storage pressure results in a required storage vessel volume of 17,000 m^3, 34,000 m^3 and 51,000 m^3 for 4, 8 and 12 hours of syngas storage, respectively. In the present model, the directly-fed and storage-fed gas turbines are the same size. Other arrangements may be more profitable (for example, choosing a different size peaking turbine or optimizing the storage pressures and volumes), but we wish to determine here only whether storing syngas for sale at peak times significantly increases profitability.

For each of the storage options (0, 4, 8, and 12 hours), the gasifier operates at maximum output at every hour (260 tons/hr), up to its availability. At every hour, the facility operator must decide how much electricity to produce from the IGCC turbine and from the peaking turbine. A profit maximizing operator stores syngas during hours with the lowest LMP and operates both turbines at hours with the highest LMPs. This storage scheme is illustrated for the case of 8 hours of storage, shown over two days in Figure 2. In the Midwest ISO over the year examined, the day-ahead and real-time hourly markets exhibited a correlation of 0.81, 0.77, and 0.74 for the 4, 8, and 12 hours of lowest LMPs respectively. It is thus a reasonable approximation for this analysis that the operator could use the day ahead LMPs to operate the storage scheme in real time.

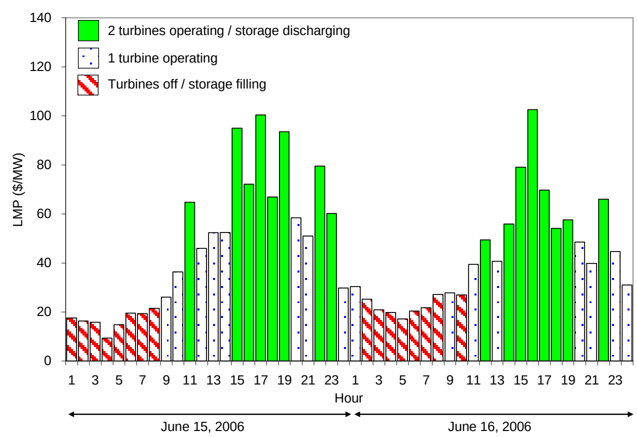

Figure 2. Storage scheme for 8 hours of syngas storage to produce peak electricity. At times of low price, the gasifier output fills storage. During high price periods, both the gasifier and stored syngas supply turbines. At intermediate prices, the gasifier output is fed to one turbine and the storage volume is unchanged.

The annual return on investment for the baseline and storage scenario is calculated as:

$$\mathrm{ROI} = \frac{\text{annual revenue}}{\text{total levelized annual expenses}} \tag{1}$$

where the annual revenue is the sum over every hour *i* of each day *j* in the year of the hourly amount of electricity produced by the IGCC turbine (MW_1) and the peaking turbine (MW_2) times the selling price of electricity at the hour (LMP) and the facility availability:

$$\text{annual revenue} = \sum_{j=1}^{365} \sum_{i=i}^{24} \left[\text{LMP}_i \cdot \left(\text{MW}_{1i} + \text{MW}_{2i} \right) \cdot \text{availability} \right]_j \tag{2}$$

and where the levelized annual expenses are the sum of the annual operating and maintenance costs and the annualized principal and debt service on the capital cost [33]:

$$\text{total levelized annual expenses} = \sum_{\substack{\text{gasifier} \\ \text{cleanup} \\ \text{air separation} \\ \text{turbine} \\ \text{compressor} \\ \text{storage}}} \begin{pmatrix} \text{annualized capital expenses} + \\ \text{annualized O \& M expenses} \end{pmatrix} \tag{3}$$

where annualized capital expenses = capital costs × (amortization factor × debt percentage)

and where annualized O&M expenses = fixed annual costs ($/yr) + (variable O&M ($/hr) × 8760 (hr/yr) × availability)

Because the levelized annual expenses are distributions, the resulting probabilistic ROI is also a distribution.

We caution that, as for construction of any peaking plant, each additional plant using this method would lower peak electricity prices, lowering the incentive for building additional plants.

Results with no carbon price

The ROI and NPV were calculated for the baseline IGCC facility and for the IGCC facility with diurnal storage; the value of adding storage to an IGCC facility was calculated by calculating the difference in economic performance.

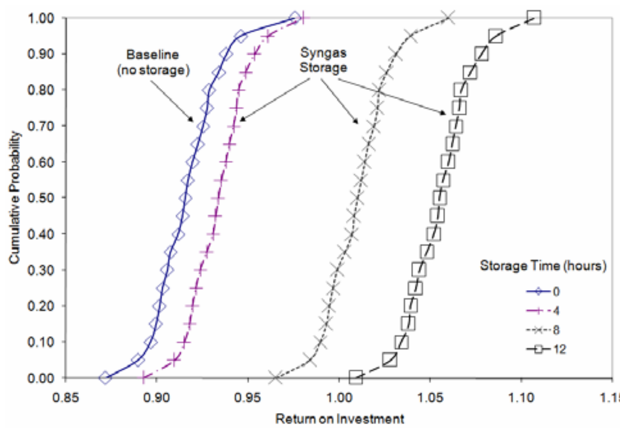

Figure 3. ROI for syngas storage scenario using a 1+0 IGCC facility with 80% availability, Cinergy node, 100% debt financing at 8% interest rate, economic and plant life of 30 years (amortization factor 0.0888), 2007 EIA AEO coal price forecast with accuracy factor, 63 bar storage pressure.

The mean ROI for the baseline 1+0 facility with no storage is 0.92 (Figure 3), suggesting that this IGCC facility would not be constructed under the assumed operating and financial parameters. The addition of 4, 8 and 12 hours of syngas storage increases the mean ROI by 2, 9 and 14 percentage points, respectively.

The NPV shows similar increases with storage; with 12 hours of syngas storage, the facility realizes increased revenue from producing and selling peak power and the NPV is $68 million ($160 million more than the baseline IGCC facility with no syngas storage). Since the magnitude of the NPV increase depends on the nodal LMPs, we have modeled locating the facility at a number of nodes in the Midwest ISO. Storage increases the NPV for all nodes examined (Figure 4).

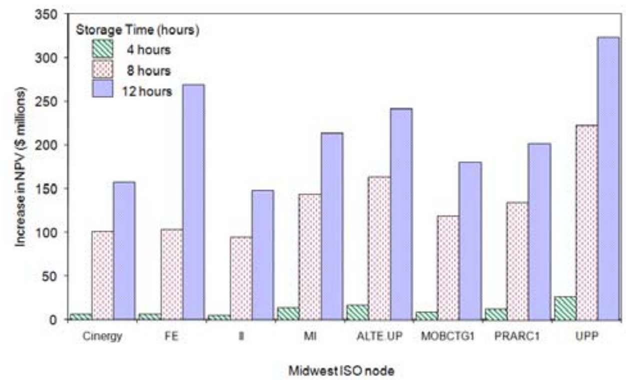

Figure 4. Increase in NPV from adding a diurnal syngas storage scheme. Parameters as in figure 3.

The sensitivity of the analysis to variations in the parameters was analyzed. The ROI for the 12 hour storage scenario is sensitive to the gasifier availability, structure of the financing, price of coal, and capital costs of the turbines, gasifier, air separation unit, and cleanup processes. The gasifier availability and the financing are the most important parameters over which the facility developer or operator has control. In addition to a plot of the sensitivity analysis, a closed-form solution of the increase in ROI using mean prices for peak and off peak LMP prices is included in the supporting information. We caution that the mean prices necessary for the closed form solution do not capture the 'peakiness' of the price duration curves, as the gains from using syngas storage depend on the differences in electricity prices at peak and off peak hours for every hour the facility is operated. In addition, we have examined the sensitivity to engineering integration choices such as the form of NOx control. For example, if steam injection is used, resulting in a plant efficiency decrease of 5% [36] when the second turbine is run, the addition of 12 hour storage increases ROI by 11% instead of by 14% with no steam injection. Other NOx control options are discussed in the supporting information.

Results with a carbon price

We have examined the implications of using diurnal syngas storage scheme at an IGCC facility in a regulatory environment with a carbon tax or carbon allowance price. A carbon price will increase the price of electricity and the revenue received by the IGCC plant. Because storage adds value and increases the ROI for an IGCC facility, the carbon price at which IGCC enters the generation mix may be lowered. The method for examining the hypothesis was to 1) re-examine the baseline (no storage) scenario with the addition of carbon capture, transport and storage process and costs; 2) increase the Midwest ISO LMP prices by redispatching the existing generation with the addition of a carbon price using heat rates and CO_2 emission factors from the US EPA's eGRID database [34]; and 3) plot the facility ROI versus the carbon tax and examine the hurdle rate crossover.

The 1+0 baseline IGCC facility was modified to include a carbon capture, transport and storage process from IECM, consisting of a water gas shift process, Selexol CO_2 capture and transport process. Appropriate adjustments to the performance and the capital and operating costs were made to the engineering economic model (the supporting information provide a comprehensive list of the processes, financial and operating parameters for the 1+0+CCS scenario).

Adding CCS increases capital costs, and incurs an energy penalty, increasing coal consumption and decreasing net electricity produced. The 1+0+CCS facility has a net output of 238 MW and a capital cost of $2,380/kW (compared to $1,540/kW for the 1+0 scenario). Implementing diurnal syngas storage with the 1+0+CCS scenario significantly improves the plant level ROI and NPV of the IGCC facility (Figure 5).

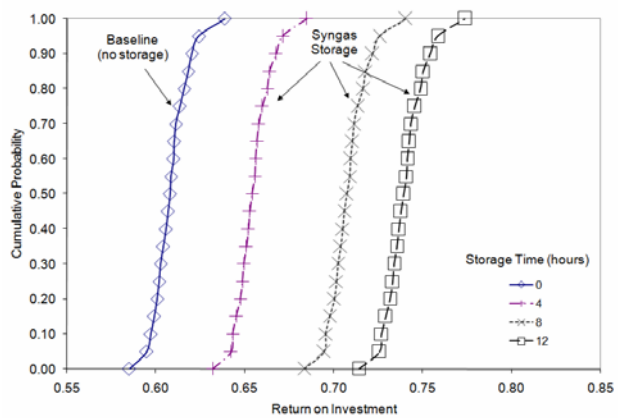

Figure 5. ROI for a 1+0 facility with carbon capture, transport and storage: 1+0 gasifier train, 80% availability, Cinergy node, 100% debt financing at 8% interest rate, economic and plant life of 30 years, 2007 EIA AEO coal price forecast with accuracy factor.

The mean ROI for the baseline 1+0+CCS facility with no storage under the assumed operating and financial parameters is 0.61. This ROI is about 30 percentage points lower than the case without CCS due to the increased capital costs and energy penalty associated with carbon capture and storage process. The addition of 4, 8 and 12 hours of syngas storage increases the mean ROI by 5, 10 and 13 percentage points, respectively (although the facility is not profitable, in the absence of special circumstances such as selling the CO_2 for enhanced oil recovery).

A carbon price will increase the price of electricity by an amount dependent on the types of generation units used. Using eGRID [34] data, we constructed a dispatch curve for the Midwest ISO, using the rate of carbon emission per kWh for each generator in the ISO reported in eGRID (figure 6, where values are shown for three carbon prices). Using these updated dispatch curves, the nodal prices incorporating the price of carbon were calculated. These prices and the hourly Midwest ISO load in each hour of the year examined were used to estimate the revenue received by the IGCC plant with storage for each hour.

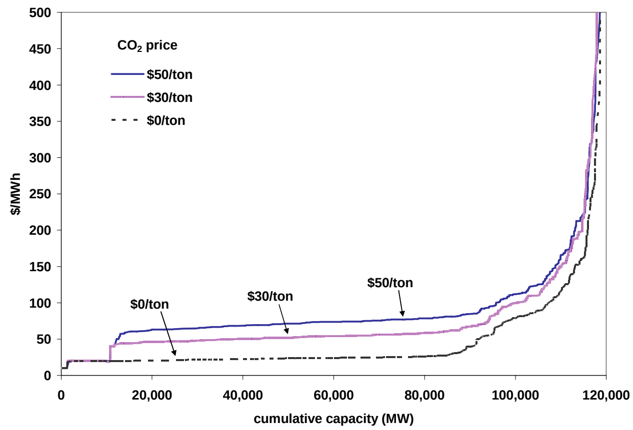

Figure 6. Midwest ISO price curves at a range of carbon prices. Redispatch analysis using eGRID data for each generator.

Using LMP data incorporating a carbon price, we reran the analysis and examined the implications of a carbon price on the ROI of an IGCC facility with syngas storage and 90% carbon capture (paying the carbon price for the remainder). Figure 7 plots the mean values of the ROI for the syngas storage scenario versus carbon price and shows that storage lowers the carbon price at which a given investment hurdle rate is achieved.

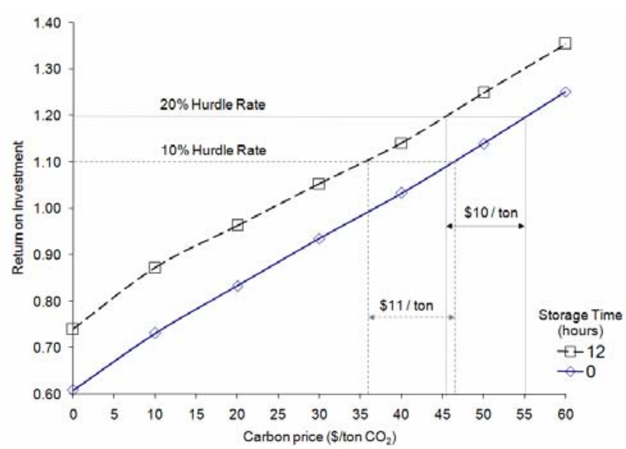

Figure 7. Effects of carbon price on return on investment for syngas storage facilities

A 20 percent ROI is attained at approximately $45/ton CO_2 with 12 hours of high pressure above ground storage, versus $55/ton for a facility with no syngas storage. Because 12 hours of syngas storage increases the ROI for the IGCC facility, the carbon tax at which a 20 percent hurdle rate is achieved is lowered by about $10/ton CO_2. The reduction in price depends on the required hurdle rate. Although private firms may require higher hurdle rates in order to undertake projects, for comparison the return on equity that state regulators currently guarantee varies from 9.45 to 12.0 percent [35]. Using this range as the hurdle rate, storage lowers the carbon price from ~ \$45/ton to ~ \$35/ton CO_2. At a lower hurdle rate of 10 percent, 12 hours of syngas storage lowers the required carbon price by \$14/ton CO_2, from \$47 to \$36/ton.

Discussion

Producing peak electricity from diurnally stored syngas in gas turbines, while operating the gasifier at a constant output, increases firm-level profits for an IGCC facility despite the additional capital cost. Storage decouples the operation of the gasifier from the turbine and allows the facility to produce electricity when it is most valuable. Storing syngas in gas spheres at a pressure of 60 bar would add 25% to the land area of the IGCC plant modeled. Other configurations, optimized storage parameters, lower fuel costs through long term contracts or more sophisticated financing arrangements may further increase profitability. Syngas storage can lower the CO_2 price at which IGCC enters the generation mix by approximately \$10/ton, speeding deployment. However, the ability of even a small fraction of generators to employ syngas storage to increase profitability is likely to lead to earlier deployment of commercial IGCC at significant scale.

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Storing syngas lowers the carbon price for profitable coal gasification

ADAM NEWCOMER AND JAY APT

Supporting Information

Table S1. 1+0+ccs scenario operating and financial parameters

IECM cs version 5.21 (February 2, 2007)

Operating parameters			Financial parameters		
Overall Plant					
Base GE Quench			Year Costs Reported	2005	
Cold gas cleanup			Constant Dollars		
CO2 Capture: Sour Shift + Selexol			Discount Rate (Before Taxes)	8.00E-02	fraction
Slag: landfill			Fixed Charge Factor (FCF)	8.88E-02	fraction
Sulfur: sulfur plant					
			Inflation Rate	0	%/yr
Capacity Factor	80	%	Plant or Project Book Life	30	years
			Real Bond Interest Rate	8	%
Gross Plant Size	297.7	MWg	Real Preferred Stock Return	0	%
Net Plant Size	238.1	MW	Real Common Stock Return	0.1	%
Net Electrical Output (MW)	238.1		Percent Debt	99.99	%
Total Plant Energy Input (MBtu/hr)	2781		Percent Equity (Preferred Stock)	0	%
Gross Plant Heat Rate, HHV (Btu/kWh)	9343		Percent Equity (Common Stock)	1.00E-02	%
Net Plant Heat Rate, HHV (Btu/kWh)	11680				
			Federal Tax Rate	35	%
Net Plant Efficiency, HHV (%)	29.17		State Tax Rate	4	%
Ambient Air Temperature	77	°F	Property Tax Rate	2	%
Ambient Air Pressure	14.7	psia	Investment Tax Credit	0	%
Ambient Air Humidity	1.80E-02	lb H2O/lb dry air	Construction Time	0.25	years
			Operating Labor Rate	24.82	$/hr
Coal			Water Cost	0.8316	$/1000 gal
Illinois #6			Sulfur Byproduct Credit	68.64	$/ton
Heating Value	1.09E+04	btu/lb	Sulfur Disposal Cost	10	$/ton
Carbon	61.2	wt% as received	Selexol Solvent Cost	2.32	$/lb
Hydrogen	4.2		Claus Plant Catalyst Cost	565.8	$/ton
Oxygen	6.02		Beavon-Stretford Catalyst Cost	218.6	$/cu ft
Chlorine	0.17		Slag Disposal Cost	13.07	$/ton
Sulfur	3.25		Limestone Cost	19.64	$/ton
Nitrogen	1.16		Lime Cost	72.01	$/ton
Ash	11		Ammonia Cost	248.2	$/ton
Moisture	13		Urea Cost	412.4	$/ton
			MEA Cost	1293	$/ton
Plant Inputs	Flow Rate (tons/hr)		Activated Carbon Cost	1322	$/ton
Coal	127.6		Caustic (NaOH) Cost	624.7	$/ton
Oil	0.3479		High Temperature Catalyst Cost	60.1	$/cu ft
Other Fuels	3.04E-02		Low Temperature Catalyst Cost	300.5	$/cu ft
Other Chemicals, Solvents & Catalyst	2.39E-03		Glycol Cost	2.356	$/lb
Total Chemicals	2.39E-03		Bulk Reagent Storage Time	60	days
Oxidant	109.3		The following apply to all process blocks		
Process Water	48.63		General Facilities Capital	15	%PFC
			Engineering & Home Office Fees	10	%PFC
Plant Outputs	Flow Rate (tons/hr)		Project Contingency Cost	15	%PFC
Slag	16.38		Booster Pump Operating Cost	1.5	%PFC
Ash Disposed	0		Pre-Production Costs		
Other Solids Disposed	0		Months of Fixed O&M	1	months
Particulate Emissions to Air	1.39E-03		Months of Variable O&M	1	months
Captured CO2	254.2		Misc. Capital Cost	2	%TPI

By-Product Ash Sold	0	
By-Product Gypsum Sold	0	
By-Product Sulfur Sold	4.066	
By-Product Sulfuric Acid Sold	0	
Total Solids & Liquids	274.6	

Inventory Capital (gasifier)	1	%TPC
Inventory Capital (other processes)	0.5	%TPC
Maint. Cost Allocated to Labor	40	% total
Administrative & Support Cost	30	% total labor
TCR Recovery Factor	100	%
Number of Operating Jobs	6.67	jobs/shift
Number of Operating Shifts	4.75	shifts/day
Royalty Fees	0.5	%PFC
Process Contingency Cost		
gasifier	11.77	%PFC
turbine	8.006	%PFC
air separation	5	%PFC
sulfur removal	8.348	%PFC
co2 capture	5	%PFC
Total Maintenance Cost		
gasifier	3.707	%TPC
turbine	1.5	%TPC
air separation	2	%TPC
sulfur removal	2	%TPC
co2 capture	2	%TPC

Plant Energy Requirements	Value
Total Generator Output (MW)	510.5
Air Compressor Use (MW)	208.6
Turbine Shaft Losses (MW)	6.036
Gross Plant Output (MWg)	297.7
Misc. Power Block Use (MW)	5.954
Air Separation Unit Use (MW)	31.77
Gasifier Use (MW)	4.343
Sulfur Capture Use (MW)	3.291
Claus Plant Use (MW)	0.4343
Beavon-Stretford Use (MW)	1.321
Water-Gas Shift Reactor Use (MW)	-11.52
Selexol CO2 Capture Use (MW)	24.01
Net Electrical Output (MW)	238.1

Gasifier Area		
Number of Operating Trains	1	
Number of Spare Trains	0	
Gasifier Temperature	2450	°F
Gasifier Pressure	615	psia
Total Water or Steam Input	0.5566	mol H2O/mol C
Oxygen Input from ASU	0.4945	mol O2/mol C
Total Carbon Loss	3	%
Sulfur Loss to Solids	0	%
Coal Ash in Raw Syngas	0	%
Percent Water in Slag Sluice	0	%
Raw Gas Cleanup Area		
Particulate Removal Efficiency	100	%
Power Requirement	1.362	% MWg

Syngas output	vol%	Syngas Out (tons/hr)
Carbon Monoxide (CO)	30.64	109.4
Hydrogen (H2)	32.92	8.478
Methane (CH4)	0.261	0.5338
Ethane (C2H6)	0	0
Propane (C3H8)	0	0
Hydrogen Sulfide (H2S)	0.975	4.237
Carbonyl Sulfide (COS)	4.10E-02	0.314
Ammonia (NH3)	8.00E-03	1.74E-02
Hydrochloric Acid (HCl)	4.80E-02	0.2231
Carbon Dioxide (CO2)	18.52	103.9
Moisture (H2O)	14.86	34.13
Nitrogen (N2)	0.864	3.086
Argon (Ar)	0.872	4.442
Total	100	268.8

GE Gasifier Process Area Costs	Capital Cost (M$)
Coal Handling	23.86
Gasification	38.5
Low Temperature Gas Cooling	19.64
Process Condensate Treatment	9.929
General Facilities Capital	13.79
Eng. & Home Office Fees	9.193
Project Contingency Cost	13.79
Process Contingency Cost	10.93
Interest Charges (AFUDC)	-4.003
Royalty Fees	0.4596
Preproduction (Startup) Cost	5.672
Inventory (Working) Capital	1.396
Total Capital Requirement (TCR)	143.2

Variable Cost Component	O&M Cost (M$/yr)
Oil	0.838
Other Fuels	2.04E-02
Water	0.2836
Slag Disposal	1.501

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	2.009
Maintenance Labor	1.966
Maintenance Material	2.949
Admin. & Support Labor	1.193

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	8.117	4.862
Annual Variable Cost (excluding coal)	2.64	1.582
Total Annual O&M Cost	10.76	6.44
Annualized Capital Cost	14.86	8.898
Total Levelized Annual Cost	25.62	15.34

Gas Turbine/Generator	
Gas Turbine Model	GE 7FA

Power Block Plant Costs	Capital Cost (M$)
Gas Turbine	54.81

No. of Gas Turbines	1	
Total Gas Turbine Output	202.6	MW
Fuel Gas Moisture Content	33	vol %
Turbine Inlet Temperature	2420	°F
Turbine Back Pressure	2	psia
Adiabatic Turbine Efficiency	95	%
Shaft/Generator Efficiency	98	%
Air Compressor		
Pressure Ratio (outlet/inlet)	15.7	ratio
Adiabatic Compressor Efficiency	70	%
Combustor		
Combustor Inlet Pressure	294	psia
Combustor Pressure Drop	4	psia
Excess Air For Combustor	171.1	% stoich.
Heat Recovery Steam Generator		
HRSG Outlet Temperature	250	°F
Steam Cycle Heat Rate, HHV	9000	Btu/kWh
Steam Turbine		
Total Steam Turbine Outlet	95.13	MW
Power Block Totals		
Power Requirement	2	% MWg

Heat Recovery Steam Generator	17.27
Steam Turbine	25.86
HRSG Feedwater System	3.611
General Facilities Capital	15.23
Eng. & Home Office Fees	10.16
Project Contingency Cost	15.23
Process Contingency Cost	8.111
Interest Charges (AFUDC)	-4.309
Royalty Fees	0.5078
Preproduction (Startup) Cost	3.307
Inventory (Working) Capital	0.7515
Total Capital Requirement (TCR)	150.6

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	1.636
Maintenance Labor	0.9018
Maintenance Material	1.353
Admin. & Support Labor	0.7612

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	4.651	2.79
Total Annual O&M Cost	4.651	2.79
Annualized Capital Cost	13.25	7.946
Total Levelized Annual Cost	17.9	10.74

Syngas Input	Syngas In (tons/hr)	Heated Syngas In (tons/hr)
Carbon Monoxide (CO)	5.471	5.471
Hydrogen (H2)	15.97	15.97
Methane (CH4)	0.5338	0.5338
Ethane (C2H6)	0	0
Propane (C3H8)	0	0
Hydrogen Sulfide (H2S)	5.30E-03	5.30E-03
Carbonyl Sulfide (COS)	3.16E-03	3.16E-03
Ammonia (NH3)	1.74E-02	1.74E-02
Hydrochloric Acid (HCl)	2.23E-01	2.23E-01
Carbon Dioxide (CO2)	13.37	13.37
Water Vapor (H2O)	33.37	76.94
Nitrogen (N2)	3.086	3.086
Argon (Ar)	4.442	4.442
Oxygen (O2)	0	0
Total	76.5	120.1

Air Separation		
Oxidant Composition		
Oxygen (O2)	95	vol %
Argon (Ar)	4.234	vol %
Nitrogen (N2)	0.7657	vol %
Final Oxidant Pressure	580	psia
Maximum Train Capacity	1.14E+04	lb-moles/hr
Number of Operating Trains	1	integer
Number of Spare Trains	0	integer
Unit ASU Power Requirement	210.4	kWh/ton O2
Total ASU Power Requirement	10.67	% MWg

Air Separation Plant Costs	Capital Cost (M$)
Process Facilities Capital	66.33
General Facilities Capital	9.949
Eng. & Home Office Fees	6.633
Project Contingency Cost	9.949
Process Contingency Cost	3.316
Interest Charges (AFUDC)	-2.757
Royalty Fees	0.3316
Preproduction (Startup) Cost	2.266
Inventory (Working) Capital	0.4809
Total Capital Requirement (TCR)	96.5

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	2.009
Maintenance Labor	0.7694

Maintenance Material	1.154	
Admin. & Support Labor	0.8337	

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	4.767	2.859
Total Annual O&M Cost	4.767	2.859
Annualized Capital Cost	8.492	5.093
Total Levelized Annual Cost	13.26	7.952

Sulfur Removal

Hydrolyzer (or Shift Reactor)		
COS to H2S Conversion Efficiency	98.5	%
Sulfur Removal Unit		
H2S Removal Efficiency	98	%
COS Removal Efficiency	33	%
CO2 Removal Efficiency	0	%
Max Syngas Capacity per Train	2.50E+04	lb-mole/hr
Number of Operating Absorbers	3	
Power Requirement	1.106	% MWg
Claus Plant		
Sulfur Recovery Efficiency	95	%
Max Sulfur Capacity per Train	1.00E+04	lb/hr
Number of Operating Absorbers	3	
Power Requirement	1.46E-01	% MWg
Tailgas Treatment		
Sulfur Recovery Efficiency	99	%
Power Requirement	0.4438	% MWg
Sulfur Sold on Market	90	%

Sulfur Removal Plant Costs	Capital Cost (M$)
Sulfur Removal System - Hydrolyzer	0
Sulfur Removal System - Selexol	13.29
Sulfur Recovery System - Claus	7.057
Tail Gas Clean Up - Beavon-Stretford	4.584
General Facilities Capital	3.739
Eng. & Home Office Fees	2.493
Project Contingency Cost	3.739
Process Contingency Cost	2.14
Interest Charges (AFUDC)	-1.062
Royalty Fees	0.1246
Preproduction (Startup) Cost	0.8692
Inventory (Working) Capital	0.1852
Total Capital Requirement (TCR)	37.16

Variable Cost Component	O&M Cost (M$/yr)
Makeup Selexol Solvent	7.77E-02
Makeup Claus Catalyst	3.36E-03
Makeup Beavon-Stretford Catalyst	4.90E-03
Sulfur Byproduct Credit	1.761
Disposal Cost	2.85E-02

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	2.009
Maintenance Labor	0.2963
Maintenance Material	0.4445
Admin. & Support Labor	0.6917

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	3.442	2.064
Annual Variable Cost	-1.647	-0.9878
Total Annual O&M Cost	1.795	1.08E+00
Annualized Capital Cost	3.27	1.961
Total Levelized Annual Cost	5.065	3.038

CO2 Capture

Water-Gas Shift Reactor

CO to CO2 Conversion Efficiency	95	%
COS to H2S Conversion Efficiency	98.5	%
Steam Added	0.99	mol H2O/mol CO
Maximum Train CO2 Capacity	1.50E+04	lb-moles/hr
Number of Operating Absorbers	2	integer
Number of Spare Absorbers	0	integer
Thermal Energy Credit	3.87	% MWg

Water Gas Shift Process Area Costs	Capital Cost (M$)
High Temperature Reactor	1.536
Low Temperature Reactor	1.722
Heat Exchangers	25.87
General Facilities Capital	4.369
Eng. & Home Office Fees	2.913
Project Contingency Cost	4.369
Process Contingency Cost	1.456
Interest Charges (AFUDC)	-1.211
Royalty Fees	0.1456
Preproduction (Startup) Cost	0.9396
Inventory (Working) Capital	0.2112
Total Capital Requirement (TCR)	42.32

Variable Cost Component	O&M Cost (M$/yr)

Water	9.25E-02

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	0.3013
Maintenance Labor	0.3379
Maintenance Material	0.5068
Admin. & Support Labor	0.1917

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	1.338	0.8023
Annual Variable Cost	9.25E-02	5.55E-02
Total Annual O&M Cost	1.43E+00	0.8578
Annualized Capital Cost	3.72E+00	2.234
Total Levelized Annual Cost	5.154	3.091

Selexol

CO2 Product Stream		
Number of Compressors	3	
Product Pressure	2000	psig
CO2 Compressor Efficiency	80	%
Transport & Storage		
Storage Method:	Geologic	
CO2 Removal Efficiency	95	%
H2S Removal Efficiency	94	%
Max Syngas Capacity per Train	3.20E+04	lb-mole/hr
Number of Operating Absorbers	2	
Number of Spare Absorbers	0	
Power Requirement	8.065	% MWg

Selexol (CO2) Process Area Costs	Capital Cost (M$)
Absorbers	7.809
Power Recovery Turbines	1.936
Slump Tanks	0.7871
Recycle Compressors	3.467
Flash Tanks	1.675
Selexol Pumps	1.589
Refrigeration	3.073
CO2 Compressors	11.95
Final Product Compressors	1.23
Heat Exchangers	3.702
General Facilities Capital	5.582
Eng. & Home Office Fees	3.722
Project Contingency Cost	5.582
Process Contingency Cost	3.722
Interest Charges (AFUDC)	7.063
Royalty Fees	0.1861
Preproduction (Startup) Cost	2.651
Inventory (Working) Capital	0.2791
Total Capital Requirement (TCR)	66

Variable Cost Component	O&M Cost (M$/yr)
CO2 Transport	3.086
CO2 Storage	9.719

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	0.6025
Maintenance Labor	1.116
Maintenance Material	1.675
Admin. & Support Labor	0.5157

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	3.909	2.345
Annual Variable Cost	1.28E+01	7.68E+00
Total Annual O&M Cost	1.67E+01	10.02
Annualized Capital Cost	5.81E+00	3.484
Total Levelized Annual Cost	22.52	13.51

CO2 Transport

Total Pipeline Length	62.14	miles
Net Pipline Elevation Change (Plant->Injection)	0	feet
Number of Booster Stations	0	integer
Compressor/Pump Driver	Electric	
Booster Pump Efficiency	75	%
Pipeline Region	Midwest US	

CO2 Transport Process Area Costs	Capital Cost (M$)
Material Cost	5.195
Labor Costs	16.73
Right-of-way Cost	2.91
Miscellaneous Costs	7.642
Interest Charges (AFUDC)	-0.9311
Total Capital Requirement (TCR)	31.54

Design Pipeline Flow (% plant cap)	100	%
Actual Pipeline Flow	1.78E+06	tons/yr
Inlet Pressure (@ power plant)	2000	psia
Min Outlet Pressure (@ storage site)	1494	psia
Average Ground Temperature	42.08	°F
Pipe Material Roughness	1.80E-03	inches
Pipe Size	10	inches

Fixed Cost Component	O&M Cost (M$/yr)
Total Fixed Costs	0.31

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	0.31	0.1859
Annual Variable Cost	0.00E+00	0.00E+00
Total Annual O&M Cost	3.10E-01	0.1859
Annualized Capital Cost	2.78E+00	1.665
Total Levelized Annual Cost	3.086	1.851

Table S2. 3+1 Scenario operating and financial parameters

IECM cs version 5.21 (February 2, 2007)

Operating parameters

Overall Plant		
Base GE Quench		
Cold gas cleanup		
Slag: landfill		
Sulfur: sulfur plant		
Capacity Factor	80	%
Gross Plant Size	945.3	MWg
Net Plant Size	813.8	MW
Net Electrical Output (MW)	813.8	
Total Plant Energy Input (MBtu/hr)	8035	
Gross Plant Heat Rate, HHV (Btu/kWh)	8500	
Net Plant Heat Rate, HHV (Btu/kWh)	9873	
Net Plant Efficiency, HHV (%)	34.56	
Ambient Air Temperature	77	°F
Ambient Air Pressure	14.7	psia
Ambient Air Humidity	1.80E-02	lb H2O/lb dry air
Coal		
Illinois #6		
Heating Value	1.09E+04	btu/lb
Carbon	61.2	wt% as received
Hydrogen	4.2	
Oxygen	6.02	
Chlorine	0.17	
Sulfur	3.25	
Nitrogen	1.16	
Ash	11	
Moisture	13	
Plant Inputs	Flow Rate (tons/hr)	
Coal	368.6	
Oil	1.191	
Other Fuels	0.1042	
Other Chemicals, Solvents & Catalyst	5.38E-03	
Total Chemicals	5.38E-03	
Oxidant	315.8	
Process Water	140.5	
Plant Outputs	Flow Rate (tons/hr)	
Slag	47.31	
Ash Disposed	0	
Other Solids Disposed	0	
Particulate Emissions to Air	4.02E-03	
Captured CO2	0	
By-Product Ash Sold	0	

Financial parameters

Year Costs Reported	2005	
Constant Dollars		
Discount Rate (Before Taxes)	8.00E-02	fraction
Fixed Charge Factor (FCF)	8.88E-02	fraction
Inflation Rate	0	%/yr
Plant or Project Book Life	30	years
Real Bond Interest Rate	8	%
Real Preferred Stock Return	0	%
Real Common Stock Return	0.1	%
Percent Debt	99.99	%
Percent Equity (Preferred Stock)	0	%
Percent Equity (Common Stock)	1.00E-02	%
Federal Tax Rate	35	%
State Tax Rate	4	%
Property Tax Rate	2	%
Investment Tax Credit	0	%
Construction Time	0.25	years
Operating Labor Rate	24.82	$/hr
Water Cost	0.8316	$/1000 gal
Sulfur Byproduct Credit	68.64	$/ton
Sulfur Disposal Cost	10	$/ton
Selexol Solvent Cost	2.32	$/lb
Claus Plant Catalyst Cost	565.8	$/ton
Beavon-Stretford Catalyst Cost	218.6	$/cu ft
Slag Disposal Cost	13.07	$/ton
Limestone Cost	19.64	$/ton
Lime Cost	72.01	$/ton
Ammonia Cost	248.2	$/ton
Urea Cost	412.4	$/ton
MEA Cost	1293	$/ton
Activated Carbon Cost	1322	$/ton
Caustic (NaOH) Cost	624.7	$/ton
The following apply to all process blocks		
General Facilities Capital	15	%PFC
Engineering & Home Office Fees	10	%PFC
Project Contingency Cost	15	%PFC
Pre-Production Costs		
Months of Fixed O&M	1	months
Months of Variable O&M	1	months
Misc. Capital Cost	2	%TPI
Inventory Capital (gasifier)	1	%TPC
Inventory Capital (other processes)	0.5	%TPC
Maint. Cost Allocated to Labor	40	% total

By-Product Gypsum Sold	0	
By-Product Sulfur Sold	11.75	
By-Product Sulfuric Acid Sold	0	
Total Solids & Liquids	59.06	

Plant Energy Requirements	Value
Total Generator Output (MW)	1538
Air Compressor Use (MW)	579.2
Turbine Shaft Losses (MW)	19.17
Gross Plant Output (MWg)	945.3
Misc. Power Block Use (MW)	18.91
Air Separation Unit Use (MW)	91.77
Gasifier Use (MW)	12.87
Sulfur Capture Use (MW)	6.143
Claus Plant Use (MW)	0.4343
Beavon-Stretford Use (MW)	1.321

Administrative & Support Cost		30	% total labor
TCR Recovery Factor		100	%
Number of Operating Jobs		6.67	jobs/shift
Number of Operating Shifts		4.75	shifts/day
Royalty Fees		0.5	%PFC
Process Contingency Cost			
	gasifier	11.77	%PFC
	turbine	8.006	%PFC
	air separation	5	%PFC
	sulfur removal	8.348	%PFC
Total Maintenance Cost			
	gasifier	3.707	%TPC
	turbine	1.5	%TPC
	air separation	2	%TPC
	sulfur removal	2	%TPC

Gasifier Area

Number of Operating Trains	3	
Number of Spare Trains	1	
Gasifier Temperature	2450	°F
Gasifier Pressure	615	psia
Total Water or Steam Input	0.5566	mol H2O/mol C
Oxygen Input from ASU	0.4945	mol O2/mol C
Total Carbon Loss	3	%
Sulfur Loss to Solids	0	%
Coal Ash in Raw Syngas	0	%
Percent Water in Slag Sluice	0	%
Raw Gas Cleanup Area		
Particulate Removal Efficiency	100	%
Power Requirement	1.362	% MWg

Syngas output	vol%	Syngas Out (tons/hr)
Carbon Monoxide (CO)	30.64	316.1
Hydrogen (H2)	32.92	24.49
Methane (CH4)	0.261	1.542
Ethane (C2H6)	0	0
Propane (C3H8)	0	0
Hydrogen Sulfide (H2S)	0.975	12.24
Carbonyl Sulfide (COS)	4.10E-02	0.9072
Ammonia (NH3)	8.00E-03	5.02E-02
Hydrochloric Acid (HCl)	4.80E-02	0.6446
Carbon Dioxide (CO2)	18.52	300.2
Moisture (H2O)	14.86	98.61
Nitrogen (N2)	0.864	8.914
Argon (Ar)	0.872	12.83
Total	100	776.6

GE Gasifier Process Area Costs	Capital Cost (M$)
Coal Handling	68.93
Gasification	148.6
Low Temperature Gas Cooling	52.52
Process Condensate Treatment	18.76
General Facilities Capital	43.32
Eng. & Home Office Fees	28.88
Project Contingency Cost	43.32
Process Contingency Cost	33.99
Interest Charges (AFUDC)	-12.57
Royalty Fees	1.444
Preproduction (Startup) Cost	16.88
Inventory (Working) Capital	4.384
Total Capital Requirement (TCR)	448.5

Variable Cost Component	O&M Cost (M$/yr)
Oil	2.869
Other Fuels	6.99E-02
Electricity	3.472
Water	0.8192
Slag Disposal	4.335

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	2.009
Maintenance Labor	6.5
Maintenance Material	9.749
Admin. & Support Labor	2.553

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	20.81	3.647
Annual Variable Cost (excluding coal)	11.57	2.026
Total Annual O&M Cost	103.9	18.2
Annualized Capital Cost	39.85	6.983
Total Levelized Annual Cost	143.7	25.19

Gas Turbine/Generator

Gas Turbine Model	GE 7FA	
No. of Gas Turbines	3	
Total Gas Turbine Output	659.3	MW
Fuel Gas Moisture Content	33	vol %
Turbine Inlet Temperature	2420	°F
Turbine Back Pressure	2	psia

Power Block Plant Costs	Capital Cost (M$)
Gas Turbine	164.4
Heat Recovery Steam Generator	51.82
Steam Turbine	77.74
HRSG Feedwater System	8.511
General Facilities Capital	45.38

Adiabatic Turbine Efficiency	95	%
Shaft/Generator Efficiency	98	%
Air Compressor		
Pressure Ratio (outlet/inlet)	15.7	ratio
Adiabatic Compressor Efficiency	70	%
Combustor		
Combustor Inlet Pressure	294	psia
Combustor Pressure Drop	4	psia
Excess Air For Combustor	171.1	% stoich.
Heat Recovery Steam Generator		
HRSG Outlet Temperature	250	°F
Steam Cycle Heat Rate, HHV	9000	Btu/kWh
Steam Turbine		
Total Steam Turbine Outlet	286	MW
Power Block Totals		
Power Requirement	2	% MWg

Eng. & Home Office Fees	30.25
Project Contingency Cost	45.38
Process Contingency Cost	24.22
Interest Charges (AFUDC)	-12.84
Royalty Fees	1.513
Preproduction (Startup) Cost	9.502
Inventory (Working) Capital	2.239
Total Capital Requirement (TCR)	448.1

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	1.636
Maintenance Labor	2.686
Maintenance Material	4.03
Admin. & Support Labor	1.297

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	9.648	1.691
Total Annual O&M Cost	9.648	1.691
Annualized Capital Cost	46.51	8.15
Total Levelized Annual Cost	56.16	9.84

Air Separation

Oxidant Composition		
Oxygen (O2)	95	vol %
Argon (Ar)	4.234	vol %
Nitrogen (N2)	0.7657	vol %
Final Oxidant Pressure	580	psia
Maximum Train Capacity	1.14E+04	lb-moles/hr
Number of Operating Trains	2	integer
Number of Spare Trains	0	integer
Unit ASU Power Requirement	210.4	kWh/ton O2
Total ASU Power Requirement	9.708	% MWg

Air Separation Plant Costs	Capital Cost (M$)
Process Facilities Capital	181.5
General Facilities Capital	27.22
Eng. & Home Office Fees	18.15
Project Contingency Cost	27.22
Process Contingency Cost	9.073
Interest Charges (AFUDC)	-7.544
Royalty Fees	0.9073
Preproduction (Startup) Cost	5.821
Inventory (Working) Capital	1.316
Total Capital Requirement (TCR)	263.6

Fixed Cost Component	O&M Cost (M$/yr)
Operating Labor	2.009
Maintenance Labor	2.105
Maintenance Material	3.158
Admin. & Support Labor	1.234

Cost Component	M$/yr	$/MWh
Annual Fixed Cost	8.506	1.49
Total Annual O&M Cost	8.506	1.49
Annualized Capital Cost	27.36	4.794
Total Levelized Annual Cost	35.87	6.285

Sulfur Removal

Hydrolyzer (or Shift Reactor)		
COS to H2S Conversion Efficiency	98.5	%
Sulfur Removal Unit		
H2S Removal Efficiency	98	%
COS Removal Efficiency	33	%
CO2 Removal Efficiency	15	%
Max Syngas Capacity per Train	2.50E+04	lb-mole/hr
Number of Operating Absorbers	4	
Power Requirement	0.6499	% MWg
Claus Plant		
Sulfur Recovery Efficiency	95	%
Max Sulfur Capacity per Train	1.00E+04	lb/hr
Number of Operating Absorbers	3	

Sulfur Removal Plant Costs	Capital Cost (M$)
Sulfur Removal System - Hydrolyzer	1.478
Sulfur Removal System - Selexol	29.56
Sulfur Recovery System - Claus	13.76
Tail Gas Clean Up - Beavon-Stretford	5.789
General Facilities Capital	7.589
Eng. & Home Office Fees	5.059
Project Contingency Cost	7.589
Process Contingency Cost	4.223
Interest Charges (AFUDC)	-2.152
Royalty Fees	0.253
Preproduction (Startup) Cost	1.415
Inventory (Working) Capital	0.3753

Power Requirement	4.60E-02	% MWg
Tailgas Treatment		
Sulfur Recovery Efficiency	99	%
Power Requirement	0.1398	% MWg
Sulfur Sold on Market	90	%

Total Capital Requirement (TCR)	74.95	
Variable Cost Component	O&M Cost (M$/yr)	
Makeup Selexol Solvent	0.175	
Makeup Claus Catalyst	9.71E-03	
Makeup Beavon-Stretford Catalyst	1.42E-02	
Sulfur Byproduct Credit	5.089	
Disposal Cost	8.24E-02	
Fixed Cost Component	O&M Cost (M$/yr)	
Operating Labor	2.009	
Maintenance Labor	0.6004	
Maintenance Material	0.9007	
Admin. & Support Labor	0.783	
Cost Component	M$/yr	$/MWh
Annual Fixed Cost	4.294	0.7523
Annual Variable Cost	-4.807	-0.8424
Total Annual O&M Cost	-0.5139	-9.01E-02
Annualized Capital Cost	7.778	1.363
Total Levelized Annual Cost	7.264	1.273

Syngas and SNG Storage

Storage options for syngas and SNG are not well reported in the literature; however, both technical and economic aspects of hydrogen and natural gas storage are addressed. From these related studies, costs for syngas and SNG storage[a] can be reasonably estimated, based on the composition and properties (pressure, temperature, etc) of the gas to be stored. Costs for syngas storage in above ground and underground ground vessels are estimated based on existing estimates for natural gas and hydrogen storage options.

Above ground options include storage in existing piping infrastructure, in gasometers or in cylindrical "bullets" common for LPG, LNG and CNG storage. Underground storage options include salt caverns and excavated rock caverns. The choice of storage vessel depends on both technical and economic considerations including the composition and quantity of the gas to be stored, the charge and discharge rates, as well as capital, operating and maintenance costs.

Options for the large scale, bulk storage of gasses include compressed gas, cryogenic liquid, solids such as metal hydrides and liquid carriers such as methanol and ammonia. Metal hydride storage is an emerging technology used for storing pure gases such as hydrogen. Liquid carriers such as methanol and ammonia are also useful for a pure gas. As syngas and SNG are gas mixtures of varying compositions, depending on the gasification process, solid and liquid carrier storage options are unlikely to be feasible and are not further considered in this paper.

Cryogenic Liquid Storage. Cryogenic liquid storage has been used for large scale hydrogen storage, with the technology largely driven by the needs of space programs. Storing liquid hydrogen presents numerous engineering challenges due to its low heat of vaporization and resultant very high loss index [1]. Because the boil-off would be too high, liquid hydrogen cannot be stored in cylindrical tanks of the

[a] As used here, a storage system includes both the storage reservoir as well as the mechanism for providing mass flow during the charging or discharging, such as a compressor.

type used for LNG [2]. Spherical tanks are used for large-scale applications because this shape has the lowest surface area for heat transfer per unit volume. The National Aeronautics and Space Administration (NASA) uses liquid hydrogen tanks up to 3.8 x 10^3 cubic meters (10^6 US gallons) which are about 22 meters in diameter [1]. Liquid hydrogen storage is expensive; costs include both the spherical storage tanks as well as the facility required for cooling and liquefaction. Capital costs for liquid hydrogen storage and liquefaction facilities from a 1986 study are illustrated in Figure S1 below.

Installed liquid hydrogen tank costs

Installed hydrogen liquefaction facility costs

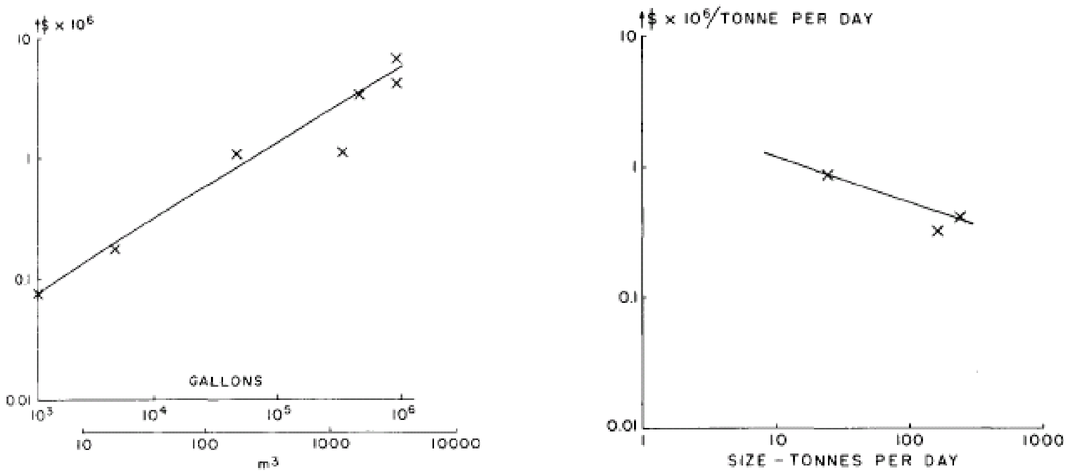

Figure S1 Capital cost of liquid hydrogen facilities [1]

From the above costs, liquid hydrogen storage capital charges, including a 15 percent ROI, are calculated to be \$1,916/tonne (\$2004)[b] [1] or approximately[c] \$350/Nm^3. Although the above study is 20 years old and steel prices have changed and high strength steel technology has improved, the reported costs are still approximately 6 to 9 times more expensive than other storage options. In addition to high costs, there are technical concerns related to liquid syngas storage. Syngas is a gas mixture and not pure gas. The chemical components that make up syngas liquefy and react at different temperatures and pressures. As such, it is unknown what technical difficulties may arise from liquefy and cryogenically storing syngas. Additionally, syngas and SNG is typically used in gaseous form for an end-use process,

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[b] Converted \$1986 Canadian to \$2004 US, using reported exchange rate of \$1C(1986) = \$0.83US(1986) and a deflator of \$1986 to \$2004 = 1.505 from http://www1.jsc.nasa.gov/bu2/inflateGDP.html

[c] Calculated using a liquid hydrogen density of 70.99g/l and STP density of 0.08988 g/l

such as combustion in a turbine. Compressing and liquefying the gas for storage (an energy consuming process), followed by expansion and vaporization for end use, is inefficient. Because of the high capital costs, technical uncertainties, and gas-to-liquid-to-gas conversion inefficiencies, liquid storage does not appear particularly suited to syngas storage, and is not further considered in this paper.

Compressed Gas Storage. Compressed gas storage is the most relevant large-scale stationary storage systems for syngas production facilities, as it can be readily used for syngas and SNG containing either hydrogen or methane. Compressed gas storage is the simplest storage solution as the only required equipment required is a compressor and a pressure vessel [2]. The main problem with compressed gas storage is the low storage density, which depends on the storage pressure. For pure hydrogen storage, several stages of compression are required because of the low density [3]. Compressed gas can be stored in high and low pressure above ground vessels, existing pipelines, and in underground cavities.

Compressors. Compressed gas storage requires a compressor to provide the necessary mass flow of gas into the storage vessel. No literature discusses syngas compression or compressor requirements for syngas service, however reasonable estimates can be drawn from literature discussing compressors for natural gas and hydrogen service. The density and molecular weight of the gas to be compressed is an important consideration for compressor choice. Centrifugal compressors, which are widely used for natural gas, are not generally suitable for pure hydrogen compression as the pressure rise per stage is very small due to the low density and low molecular weight [2, 4]. Positive displacement, reciprocating compressors may be the best choice for large-scale hydrogen compression [4], and hydrogen can be compressed using standard axial, radial or reciprocating piston-type compressors with slight modifications of the seals to take into account the higher diffusivity of the hydrogen molecules [2].

The capital costs of compression depend on the properties of the gas to be compressed. Compressing pure hydrogen requires about three times the compressor power as natural gas and specific capital costs for large hydrogen compressors are expected to be 20 to 30 percent higher than for natural gas [5]. Compressor costs are based on the amount of work done by the compressor, which depends on the inlet pressure, outlet pressure, and flow rate [2]. Capital costs of compressors reported in the literature range from $479-$4,900/hp ($650-$6,600/kW) and are shown in Table S3.

Table S3. Small compressor capital costs [1, 2]

Size (hp)	Capital cost ($)	Cost/hp ($/hp)	Source
13	63,700	4,900	Amos
100	180,000	1,800	Amos
100	187,373	1,874	Taylor[d]
335	164,150-246,225	n/a	Amos
3,600	2,330,000	647	Amos
3,600	2,248,470	625	Amos
5,000	2,440,000	488	Amos
6,000	3,160,000	527	Amos
6,000	2,873,045	479	Taylor
38,000	20,000,000	526	Amos

Costs for large-scale, megawatt sized compression facilities for pipeline transport were developed by the International Energy Agency (IEA) [6] and are shown in Table S4.

Table S4. Compressor capital cost estimates for large (MW) pipeline compressors ($MM)

Type	Initial Pressure Facility	Booster Station
Electrical Power Generation Plant CO_2 export pipeline	$5.590 + 0.509P - 0.006 P^2$	$6.388 + 0.581P - 0.008 P^2$
Fuel Synthesis Plant Hydrogen product pipeline	$24.902 + 0.549P - 0.005 P^2$	$28.460 + 0.628P - 0.005 P^2$
CO_2 Storage Facilities	$5.590 + 0.509P - 0.006 P^2$	$6.388 + 0.581P - 0.008 P^2$
Pipeline Branch CO_2	$6.388 + 0.581P - 0.008 P^2$	$6.388 + 0.581P - 0.008 P^2$
Natural Gas and Hydrogen	$28.460 + 0.628P - 0.005 P^2$	$28.460 + 0.628P - 0.005 P^2$

where P is the compressor power in MW

[d] Taylor figures converted from $1986 Canadian to $2004 US. Using $1C(1986) = $0.83US(1986) and a deflator of $1986 to $2004 = 1.505 from http://www1.jsc.nasa.gov/bu2/inflateGDP.html

The costs developed by the IEA are significantly higher than the costs reported in Table S3. For example, the IEA estimate for the 38,000hp (28 MW) compressor listed in Table S3 is about $36 million, or 1.8 times higher than the cost reported by Amos. Because of this difference, care should be taken to choose the appropriate cost estimated based on the size of the compressor when estimating compressor capital costs.

The largest operating cost for compressors is the energy required to compress the gas[2]. The exact energy requirements for compression depend on the desired final pressure. The theoretical work for isothermal compression of ideal gas from pressure p_1 to p_2 is given by:

$$W_{1,2} = p_1 V_1 \ln\left(\frac{p_2}{p_1}\right) \tag{1}$$

where V_1 is the volume of the gas at pressure p_1. Figure S2 illustrates the work required to compress a gas from an initial pressure, p_1, to a higher pressure, p_2.

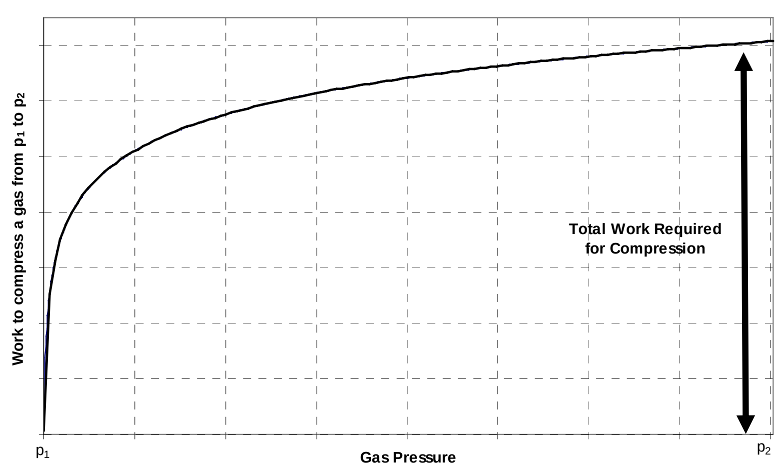

Figure S2. Work to compress an ideal gas from P_1 to P_2

Because of the logarithmic relationship, the work and electricity consumption of the compressor is highest in the low-pressure range, and a high final storage pressure requires minimal power compared to the initial compression of the gas.

The physical parameters necessary for the model are related to the compression of the gas for storage. Compression increases the pressure and changes the volumetric density of the gas. The volumetric density of a gas mixture varies with the pressure of the gas. The ideal gas law can be used to determine the relationships between compression and pressure of a gas to first order. Some gases may vary significantly from the ideal gas law, particularly at high pressures, and may be more accurately described by cubic equations of state. To determine how the volumetric density varies with pressure, pure methane, syngas[e] and SNG[f] gases were modeled in Aspen using the ideal gas law, as well as the more accurate, Soave-Redlich-Kwong (SRK), and Peng-Robinson equations of state [7]. The results of the models are illustrated in Figure S3.

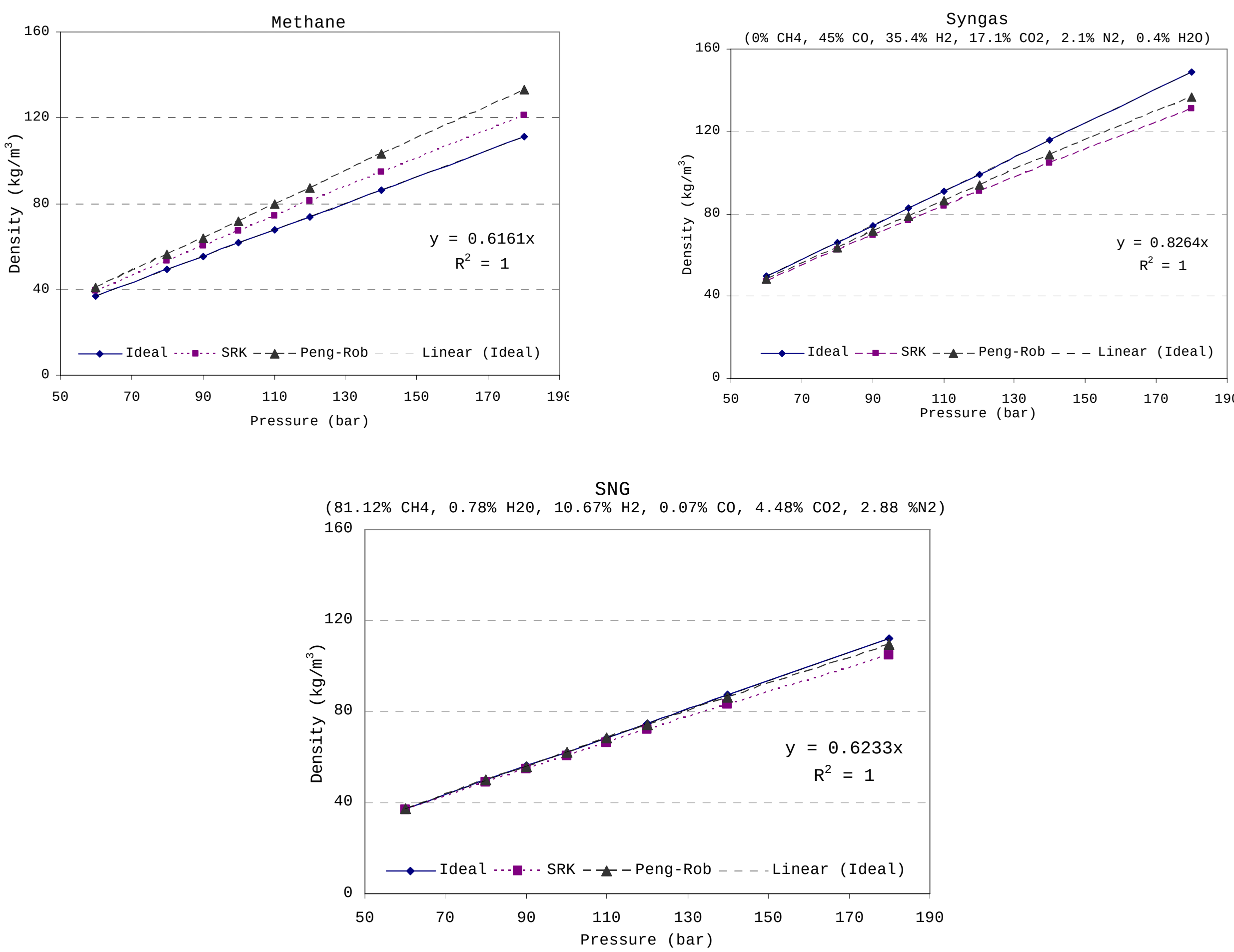

Figure S3. Volumetric density versus pressure for three different gas mixtures using three different equations of state

[e] Composition by weight: 0% CH_4, 45% CO, 35.4% H_2, 17.1% CO_2, 2.1% N_2, 0.4% H_2O

[f] Composition by weight 81.12% CH_4, 0.78% H_20, 10.67% H_2, 0.07% CO, 4.48% CO_2, 2.88% N_2

For each of the fuels modeled, the volumetric density varies linearly with pressure and none of the gas mixtures varies significantly from the ideal gas law, even at high pressures. The models show that the ideal gas law is a reasonable approximation for estimating volumetric density at varying pressure for methane, SNG and syngas.

Above Ground Compressed Gas Storage. Conventional methods of above-ground compressed gas storage range from small high-pressure gas cylinders to large, low-pressure spherical gas containers [3, 8]. Compressed gas pressure vessels are commercially available at pressures of 1200-8000 psi, typically holding 6000-9000 scf per vessel. Low-pressure spherical tanks can hold roughly 13,000 Nm^3 of gas at 1.2-1.6 MPa (1,700-2,300 psig) [2]. High pressure tube storage is available for larger gas volumes, typically around 500,000 scf (14,000 Nm^3) [1]. Because of the relatively small storage capacity, industrial facilities typically use above ground compressed gas storage in pressure tanks for gas storage on the order of a few million scf or less [5]. Pressure vessels are physically configured in rows or in stacks of tanks; such storage is modular, with little economy of scale [2].

Capital costs for above ground pressure vessel storage range from approximately \$22-\$214/Nm^3 (\$0.62-\$6.02/scf), as shown in Table S5.

Table S5. Above ground high pressure vessel capital costs [1, 2, 9]

Size (Nm^3)	Capital cost ($)	Cost/Nm^3 ($/$Nm^3$)	Source
2,800	187,373	67	Taylor[g]
14,000	874,405	62	Taylor
12,071	840,000	70	Amos
2,414	180,000	75	Amos
44	3,560	80	Amos
4,433	540,350	122	Amos
n/a	n/a	38.4 - 64	Amos
n/a	n/a	21.76 -115.2	Padró
n/a	n/a	51.2 - 213.76	Newson, Huston, Ledjeff, Carlson, reported in Padró
n/a	n/a	64.6 - 214	Capretis reported in Amos
n/a	n/a	98.1 -144	Oy, reported in Amos

Sizes and other physical parameters for the smallest and largest reported cost per storage volume in the range are not reported, making it difficult to explain why they vary significantly from the average costs.

Gasometer Storage. Gasometers are above ground vessels designed for storing large amounts of gas, typically at low pressure. Gasometers typically have a variable volume, through the use of a weighted movable cap, which provides gas output at a constant pressure. Gasometers operate at low pressure, with typical pressures in the range of 200-300mm water (0.28-0.43psig); maximum operating pressures are 1000mm water (1.4psig) [10]. Typical volumes for large gasometers are about 50,000-70,000m^3, with approximately 60 m diameter structures; although the largest gasholder installed by one manufacturer was 340,000m^3 [10]. Gasometers have long operating lifetimes; the structure itself can operate for over 100 years [10], while the diaphragm that seals the gasometer has a lifetime of 200,000 strokes or approximately 10 years [11].

[g] Taylor figures converted from $1986 Canadian to $2004 US. Using $1C(1986)=$0.83US(1986) and a deflator of $1986 to $2004=1.505 from http://www1.jsc.nasa.gov/bu2/inflateGDP.html

Table S6. Above ground low pressure vessel (gasometer) capital costs

Size (Nm^3)	Capital cost ($)	Cost/Nm^3 ($/$Nm^3$)	Source
65,000	22,080,000[h]	340	Clayton Walker [10]

Pipeline Storage. Syngas can also be stored, or packed, in piping systems. Pipelines are usually several miles long, and in some cases may be hundreds of miles long. Because of the large volume of piping systems, a slight change in the operating pressure of a pipeline system can result in a large change in the amount of gas contained within the piping network. By making small changes in operating pressure, the pipeline can effectively used as a storage vessel [2]. Storing gas in an existing pipeline system by increasing the operating pressure requires no additional capital expense as long as the pressure rating of the pipe and the capacity of the compressors are not exceeded [2]. Existing hydrogen pipelines are generally constructed of 0.25-0.30 m (10-12 in) commercial steel and operate at 1-3 MPa (145-435 psig); natural gas mains for comparison are constructed of pipe as large as 2.5 m (5 ft) in diameter and have working pressures of 7.5 MPa (1,100 psig) [12]. A 30 km, 3 inch diameter hydrogen distribution pipeline could carry a flow of 5 MMscf of hydrogen per day. Assuming that the pipeline operated at 1000 psi, the storage volume available in the pipeline would be 340,000 scf, or about 7 percent of the total daily flow rate [5].

Underground Compressed Gas Storage. Underground storage is a special case of compressed gas storage where the vessel is located underground and generally has a lower cost [2]. Because of their large capacities and low cost, underground compressed gas systems are generally most suitable for large quantities and/or long storage times [9]. There are four underground formations in which gas can be stored under pressure: (a) depleted oil or gas field; (b) aquifers; (c) excavated rock caverns; and (d) salt caverns [1].

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[h] Converted from reported cost of £12 million (UK 2006) using £1(UK) = $1.84 US. Single lift, Wiggins, dry seal gasometer.

There is significant industrial experience in underground gas storage: natural gas has been stored underground since 1916 [1]; the city of Kiel, Germany has been storing town gas (60-65 percent hydrogen) in a gas cavern since 1971 [1]; Gaz de France has stored town gas containing 50 percent hydrogen in a 330 million cubic meter aquifer structure near Beynes, France; Imperial Chemical Industries stores hydrogen at 50 atm ($5x10^6$ Pa) pressure in three brine compensated salt caverns at 1200 ft (366 m) near Teeside, UK; and in Texas, helium is stored in rock strata beneath an aquifer whereby water seals the rock fissures above the helium reservoir, sealing in the helium atoms [4].

Underground storage volumes in depleted oil and gas fields can be extremely large; volumes of gas stored exceed 10^9 m^3 and pressures can be up to 40 atm. Salt caverns, large underground voids that are formed by solution mining of salt as brine, tend to be smaller, typically around 10^6-10^7 m^3. Although smaller, salt caverns offer faster discharge rates and tend to be tighter than other underground formations, reducing leakage. Hydrogen, a small molecule with high leakage rates, has been stored in salt caverns [13]. Rock caverns are usually smaller cavities, typically on the order of 1 million to10 million m^3.

Underground gas storage requires the use of a cushion gas that occupies the underground storage volume at the end of the discharge cycle. Cushion gas is non-recoverable base gas necessary to pressurize the storage reservoir. Cushion gas can be as much as 50 percent of the working volume, or several hundred thousand kilograms of gas [2] and the cost of the cushion gas is a significant part of the capital costs for large storage reservoirs [1].

Capital costs for underground storage are reported in the literature. Underground storage is reported to be the most inexpensive means of storage for large quantities of gas, up to two orders of magnitude less expensive than other methods [2, 8]. The only case where underground storage would not be the least cost option is with small quantities of gas in large caverns where the amount of working capital invested in the cushion gas is large compared to the amount of gas stored [2]. Capital costs vary depending on whether there is a suitable natural cavern or rock formation, or whether a cavern must be mined. An abandoned natural gas well was reported to be the least expensive, however the likelihood of a gasification facility being near such a formation (and choosing to use it to store syngas rather than to sequester CO_2), seems small, so it is not further considered in this paper. Solution mining, excavating a salt formation with a brine solution, capital costs were estimated at \$19-\$23/m^3 (\$0.54\$0.66/ft^3) [8]; hard rock mining costs were estimated at \$34-\$84/m^3 (\$1.00-\$2.50/ft^3) depending on the depth [2]. Additionally, construction times for underground storage facilities can be long and may contribute to their costs. One estimate for solution mining a salt formation to create a 160 million cubic foot cavern was 2.5 years [14]. Table S7 shows reported ranges of underground storage capital costs for salt and excavated rock caverns.

Table S7. Underground storage capital cost estimates [1, 2, 8]

Salt caverns	Excavated rock caverns	Source
\$19-\$23/m^3 (\$0.54-\$0.66/ft^3)		Carpetis
	\$34-\$84/m^3 (\$1.00-\$2.50/ft^3)	Amos
\$19.50/m^3 (\$0.55/ft^3)		Taylor[a]

Underground compressed gas storage has been successfully used for compressed air energy storage (CAES) systems. There are currently two operating CAES systems in the world, both of which use salt caverns for air storage. The 290 MW Huntorf project in Germany uses a 62 MW compressor train to charge an 11 million ft^3 cavern to 1015 psi. The 110 MW McIntosh project in the US uses a 53 MW compressor train to charge a 19.8 million ft^3 cavern to 1100 psi [14].

As with all storage technologies, the overall cost of storage depends on throughput and storage time [9]. The longer the gas is to be stored, the more favorable underground storage becomes because of lower capital costs. If gas is stored for a long time, the operating cost can be a small factor compared to the capital costs of storage [2]. Operating costs for underground storage are primarily for compression power and limited to the energy and maintenance costs related to compressing the gas into underground storage and possibly boosting the pressure coming back out [9, 15]. The cost of the electricity requirements to compress the gas is independent of storage volume, which means the cost of underground storage is very insensitive to changes in storage time [2]. If the gasification facility is not geographically located near an area with suitable underground storage, transport costs would also need to be considered in the engineering economic analysis.

Syngas Storage Process Block Description

The design of the syngas storage scenario used in this analysis is conceptual and is provided to outline the potential benefits of such a system and to open a line enquiry as to whether syngas storage should be fully considered in the design of an IGCC facility. The purpose of the compression and storage component is to compress the syngas coming out of the gasifier to increase its density and reduce its storage volume.

The process description used in the analysis is illustrated in Figure Sx.

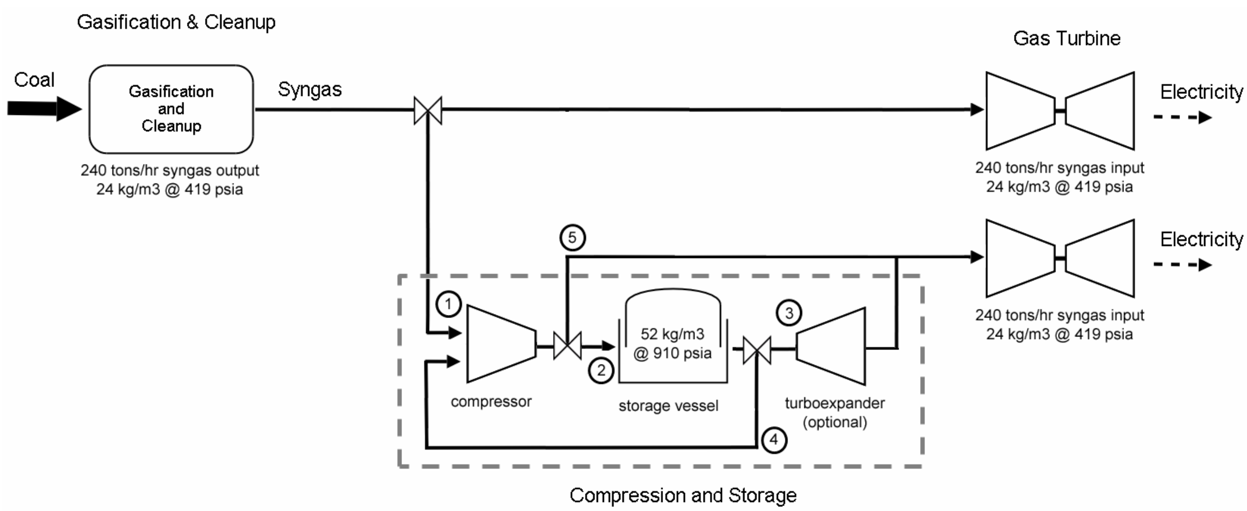

Figure S4. Conceptual design of syngas storage process block used in the analysis

The syngas storage process for this analysis is:

1. syngas from the gasification and cleanup block is pressurized to 910 psia

2. the high pressure syngas is stored in a vessel

3. high pressure syngas is released out of the storage vessel at a controlled rate (although not considered here, energy may be recovered through a turboexpander) and used in the peaking turbine

4. as the pressure in the storage vessel is reduced, the syngas is routed through the compressor to maintain an input pressure required by the peaking turbine

5. syngas at 419 psia is routed to the peaking turbine

The particular arrangement and operating parameters will depend on site specific details such as the type of coal, gasifier and gas turbine and it is possible that there will be areas where energy losses can be reduced and efficiencies increased through smart engineering design.

For the gasifier and turbine used in the analysis, a 5.6 MW compressor is required to increase syngas pressure to approximately 910 psia (63 bar) for storage. At that pressure, a 20 meter diameter storage sphere will hold enough syngas for approximately 1 hour of turbine operation. (A larger storage vessel would require reduced storage pressure and a smaller compressor). The worst case operating scenario , is that 5 MW are required to compress the syngas, and then 5MW are required to pull syngas out of storage and into the turbine (10 MW loss). This assumes that there is no turboexpander and that the compressor must be operated for the entire discharge cycle (an overestimate).

Historical Accuracy of Energy Information Administration (EIA) Price Forecasts

The economic results of the analysis depend, in part, on the price at which the facility can purchase coal. The analysis examined coal price data from different sources and timeframes in order to analyze the scenarios within an envelope of prices incorporating the recent past as well as future forecasts. The coal prices used include: historical FOB prices for Illinois #6 coal, with a higher heating value of 11,350 Btu/lb and a sulfur content of 3.2 percent by weight [16]; Energy Information Administration Annual Energy Outlook (AEO) forecasts for year 2007 coal prices[i] [17, 18]; 2007 NYMEX futures for central application coal [19]; and EIA forecasts for year 2007 coal prices with a factor that includes EIA's historical error in forecasting price data [20]. This last price distribution incorporates uncertainty in the price due to error in EIA forecasts.

EIA price forecasts do not include much data on the relative uncertainty in the estimate. We the uncertainty in EIA Annual Energy Outlook (AEO) price forecasts by examining historical deviation of actual prices from EIA forecasted prices following the methods from Rode and Fischbeck [20].

Using recent historical AEO forecast data from 1994 to 2005, we model a EIA forecast error as a normal distribution with a mean of 2.5 percent and a standard deviation of 5.0 percent. The EIA forecast error was applied to the 2007 EIA forecast from the Annual Energy Outlook. Figure S4 shows the 2007 EIA forecast from the Annual Energy Outlook compared to the same forecast with the EIA historical accuracy factor for the error included.

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[i] The mean estimate is taken from the 2007 Annual Energy Outlook (early release), Table 15, delivered prices for electric power; the standard deviation is derived from data in the December 2006 Short-Term Energy Outlook.

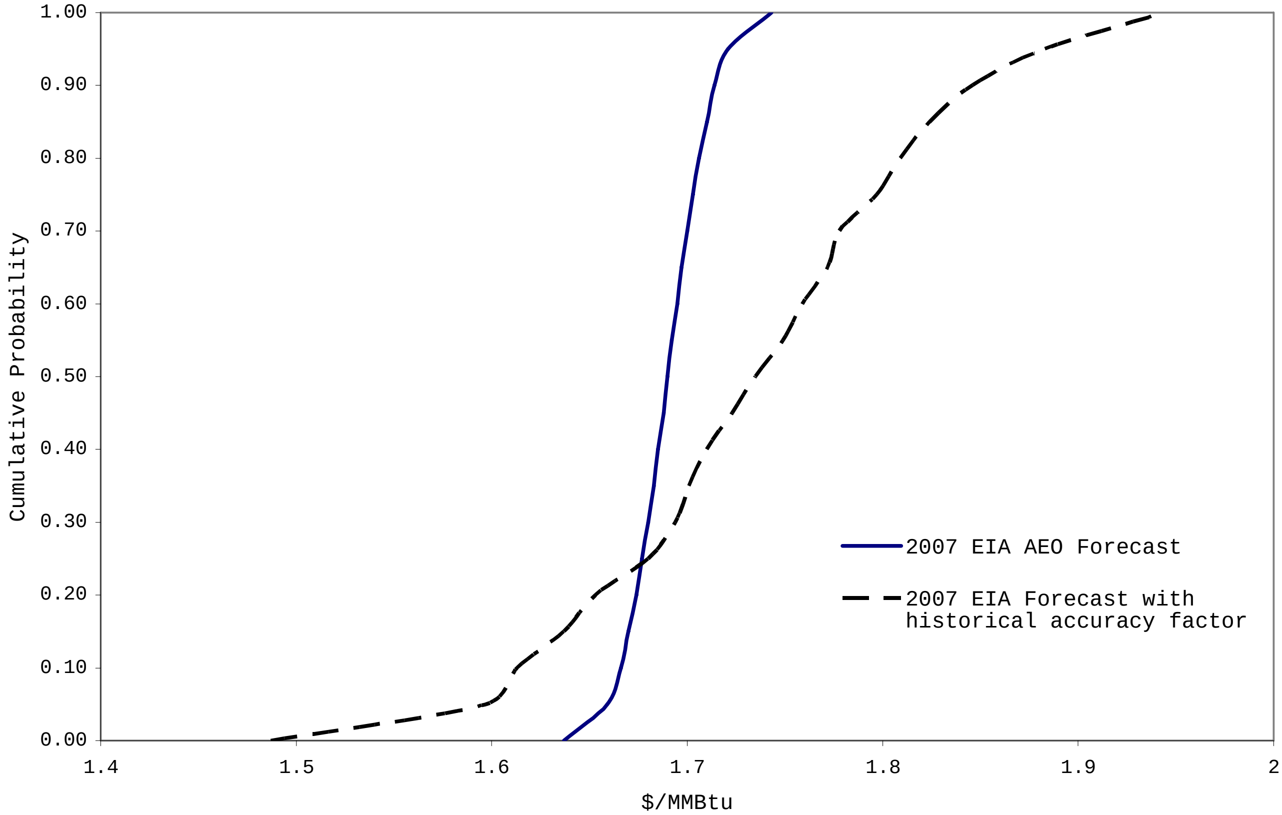

Figure S5. CDF of 2007 EIA AEO coal price forecasts with and without the historical accuracy factor

As the figure shows, including a factor which incorporates the historical error in EIA forecasts significantly widens the cdf for coal prices. It is this broader price distribution, reflecting greater uncertainty in the future price for coal that is used in the analysis.

Figure S5 illustrates the cumulative distribution functions of the coal price distributions examined in the analysis including the EIA forecast with the historical accuracy factor.

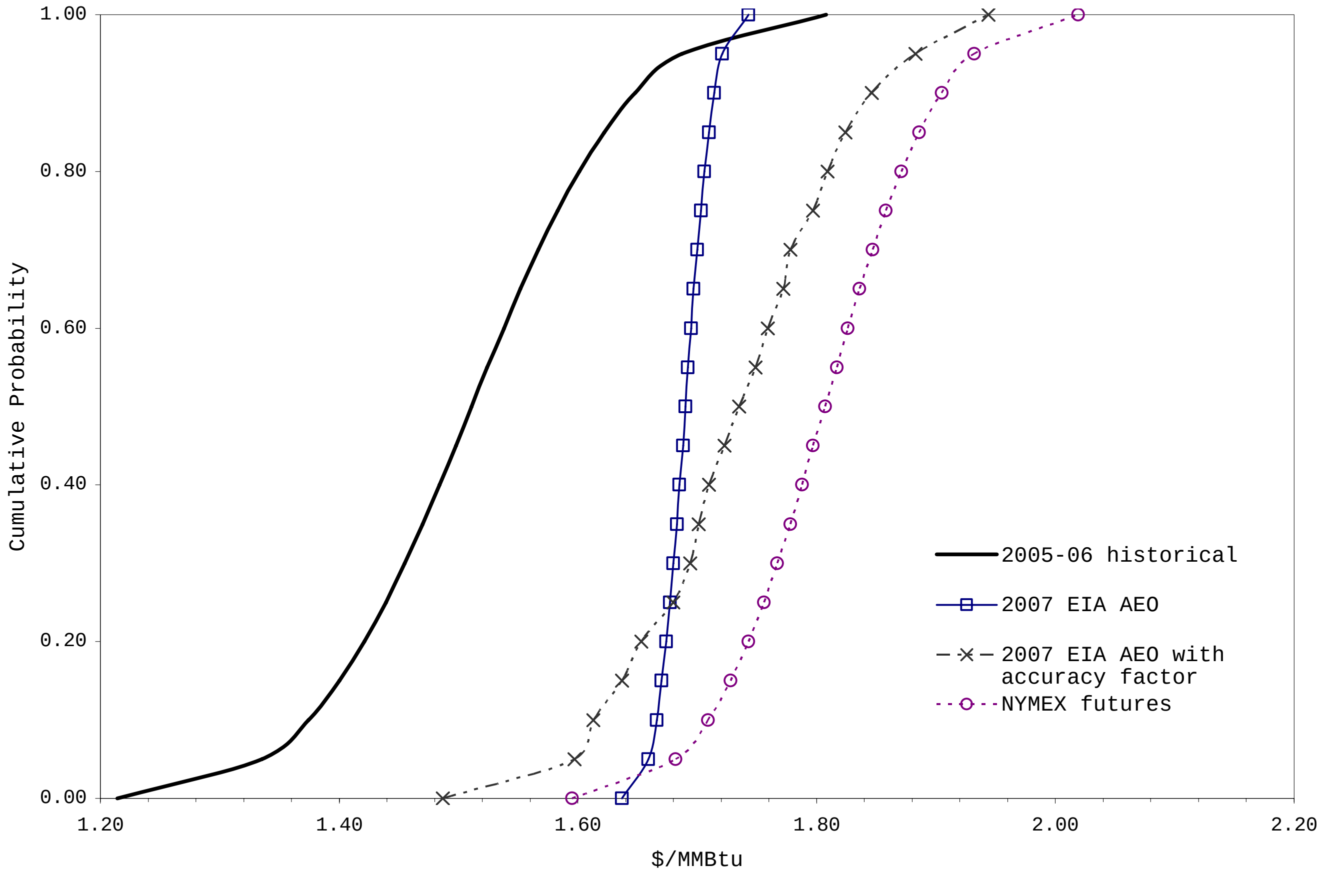

Figure S6. Coal price distributions. CDF of historical and future FOB coal prices [16-18, 20]

The historical 2005-06 prices have a mean of $1.51/MMBtu and standard deviation of 0.13. The 2007 EIA forecast shown in the figure has a mean value of $1.69/MMBtu and a standard deviation of 0.02. The 2007 EIA forecast that included the historical accuracy factor has a mean value of $1.73/MMBtu and a standard deviation of 0.10. The NYMEX futures price for Central Appalachian coal is higher than the EIA and historical prices for Illinois #6 coal, with a mean value of $1.81/MMBtu and a standard deviation of 0.09. Although futures prices vary as the contract settlement date approaches, and although Appalachian coal has a lower sulfur content than the Illinois coal, the NYMEX futures price serves as a useful upper bound for the Illinois coal price distribution. The forecasted future prices for coal represent an approximate 15 percent increase over the historical 2005-06 prices.

Constructing Cost Distributions from Cost Data

Distributions of costs were used in the analysis to capture the uncertainty in the cost parameter. Cost distributions were constructed directly from the cost data. The cost data were plotted on the y-axis against the relevant parameter (size, output, etc) on the x-axis and a mean regression line was calculated using an ordinary least squares method shown in equation 2.

$$\text{mean regression line: } \hat{y} = \beta_0 + \beta_1 x_0 \tag{2}$$

where β_0 and β_1 are calculated using the usual method of ordinary least squares. At any point x_0, the prediction interval for the value of y is given by

$$\text{prediction interval: } \hat{y} \pm t_{1-\alpha/2} \cdot se(\hat{y}_0) \tag{3}$$

$$= \hat{y} \pm t_{1-\alpha/2} \cdot \sqrt{\sigma^2 \cdot \left(1 + \frac{1}{n} + \frac{(x_0 - \bar{x})^2}{\sum (x_i - \bar{x})^2}\right)} \tag{4}$$

where $t_{1-\alpha/2}$ is the student's t distribution evaluated at the α significance level, *se* is the standard error, $\bar{x}$ is the average and σ^2 is the mean square error. Figure S6 illustrates the prediction interval for the value of y at any given x value, in relation to the underlying data.

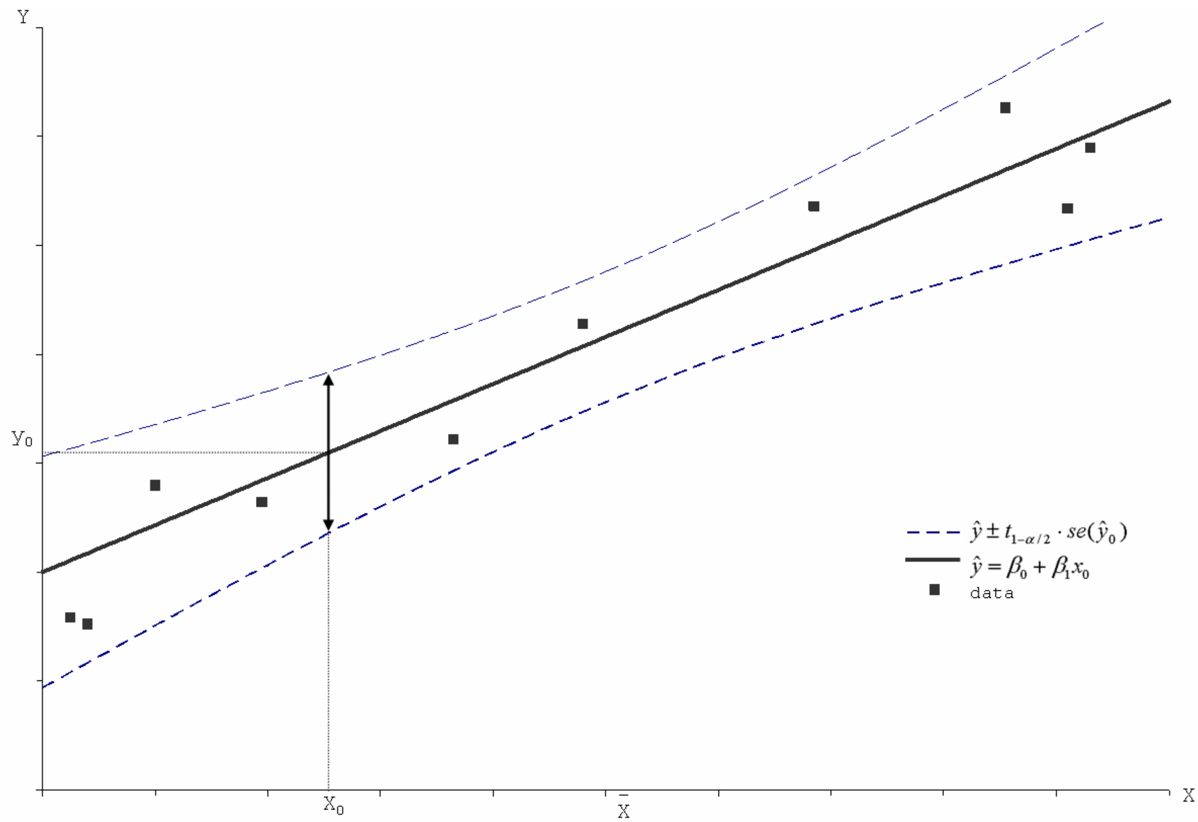

Figure S7. Regression analysis illustration with underlying data points, mean regression line, and upper and lower prediction intervals plotted. The mean and prediction interval for the value of y at point x_0 is shown.

The figure shows the individual data, the mean regression line and the prediction interval. The mean regression line represents the point estimate for the value of y given a value of x. The prediction interval represents the distribution at the α confidence level for the value of y given a value of x. As the figure illustrates, as x_0 moves away from the mean value of x, the prediction interval spreads out indicating more uncertainty in the value of y at the point x_0. At any point x0, the distribution of y can be plotted using equations 3 and 4. Figure S7 shows the cumulative distribution function of the value of y at a point x_0.

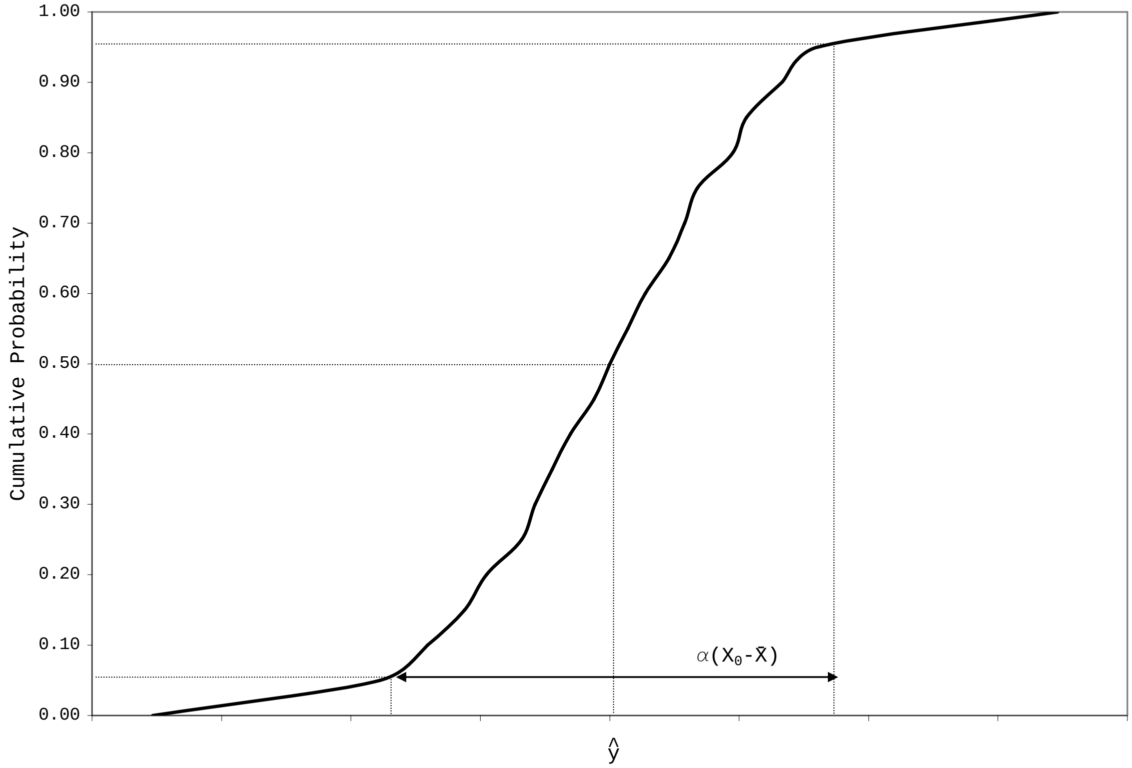

Figure S8. Cumulative distribution function of the value of Y at point x_0

As the figure shows, the standard error of y increases as x_0 moves away from $\bar{x}$, resulting in a wider cumulative distribution function.

Hydrogen Embrittlement

There is significant research on embrittlement and other metallurgical issues associated with hydrogen and hydrogen-rich gases. The oil and gas industry has been troubled by internal and external hydrogen attack on steel pipelines, described variously as hydrogen-induced cracking (or corrosion) (HIC), hydrogen corrosion cracking (HCC), stress corrosion cracking (SCC), hydrogen embrittlement (HE), and delayed failure [4]. These issues are serious; corrosion damages cause most of the failures and emergencies of trunk gas pipelines, and stress corrosion defects of pipelines are extremely severe. Corrosion defects, such as general corrosion, pitting corrosion and SCC, make up the major number of detected effects in pipelines [21].

Hydrogen can cause corrosion, hydrogen induced cracking or hydrogen embrittlement if there is a mechanism that produces atomic hydrogen (H^+) [6]. Atomic hydrogen diffuses into a metal and reforms as microscopic pockets of molecular hydrogen gas, causing cracking, embrittlement and corrosion which can ultimately lead to failure. The hardness of a metal correlates to the degree of embrittlement; if a material has a Vickers Hardness Number (VHN) greater than 300, the tendency for the material to fail due to plastic straining when there is significant absorption of atomic hydrogen is greater than with a softer material [21].

Molecular hydrogen (H_2) alone does not cause embrittlement of steel; however problems can arise if there is a mechanism that produces atomic hydrogen. The two primary mechanisms leading to hydrogen induced cracking are HIC due to wet conditions and HIC due to elevated temperatures [21]. Temperatures greater than 220°C can cause dissociation of molecular hydrogen into atomic hydrogen. Studies show that molecular hydrogen should be water dry, or below 60 percent relative humidity, to provide a sufficient margin for avoidance of moisture and water dropout [6]. Molecular hydrogen, then, may be handled without problems with standard low-alloy carbon steel irrespective of the gas pressure, provided that the conditions are dry (to prevent HIC due to wet conditions) and under 220°C (to prevent HIC due to elevated temperatures) [6].

Because of the metallurgical issues associated with hydrogen, care must be taken when choosing metals for hydrogen pipelines and storage. Surveys of existing hydrogen pipelines show that a variety of steels, but primarily mild steel, is in use [22, 23]. Options for steel pipe for 100 percent hydrogen service include Al-Fe (aluminum-iron) alloy; and variable-hardness pipe, with the harder material in the interior and softer material toward the exterior, so that any hydrogen which diffuses into the interior steel diffuses rapidly outward and escapes [4].

Existing natural gas pipelines can be used for less than 15 to 20 percent hydrogen, by volume, without danger of hydrogen attack on the line pipe steel, however further hydrogen enrichment will risk hydrogen embrittlement [4]. Existing pipelines originally designed for sour service can provide additional protection against HIC and hydrogen embrittlement due to their specific metallurgy [6]. If hydrogen embrittlement is found to be a potential problem for an unusual situation, costs for any materials will be relatively low. Steel used for hydrogen transport and storage are low carbon steel and low in alloy content. These steels may have a restriction of some alloy elements (those that attract and stabilize H and a structure called austenite), however the cost should not be affect by these restrictions [24]. For large diameter pipelines and vessels, options include low carbon steel plate, such as type X52, which is easy to make, readily available, easy to weld, and easy to fabricate. Smaller pipes can be constructed from either seamless or welded pipe. The main failure of the material is by hydrogen embrittlement in the zone near the weld. This area is affected by the heating and cooling during welding and has more internal stress. Because of the care required for welding, the most costly component is likely welding by certified welders [24].

Sensitivity of ROI to Facility Size

Larger IGCC facilities may have a smaller fraction of capital devoted to spare equipment, and may also benefit from economies of scale, thereby increasing the ROI. The ROI was examined for a larger baseline IGCC facility with 3 operating gasifiers and one spare (3+1) producing 814 net MW, with no storage. With the addition of syngas storage and turbines to produce peaking power, the facility produces 1.6 net GW of electricity. Table S8 shows the capital, operating and maintenance costs for a smaller, 1+1, and larger, 3+1, IGCC facilities.

Table S8. Baseline facility capital cost comparison (gasifier + spare)

	Capital costs (mean)		O&M costs			
	$2005 million		Fixed ($M/y)		Variable	
Component	(1+1)	(3+1)	(1+1)	(3+1)	(1+1)	(3+1)
Gasifier	196.6	448.3	10.8	20.8	609 ($/hr)	1,671 ($/hr)
Air Separation Unit	90.5	255.6	4.7	8.1	1,530 ($/hr)	3,684 ($/hr)
Cold-gas Cleanup	32.4	72.9	3.3	4.3	-22 ($/hr)	-367 ($/hr)
Power block	154.6	459.3	4.7	9.7	-6.2 ($/MWh)	-5.5 ($/MWh)
Total ($ million)	474	1,236	23.5	42.9		
($/kW)	1,760	1,520				

As the table shows, the capital cost per kW is lower for the larger sized facility, because of the increases in economies of scale. Although the 3+1 facility requires a larger volume of storage capacity, the increases in profit due to selling electricity during peak periods offsets any additional capital costs. Figure S8 shows the ROI for the 3+1 facility with and without storage.

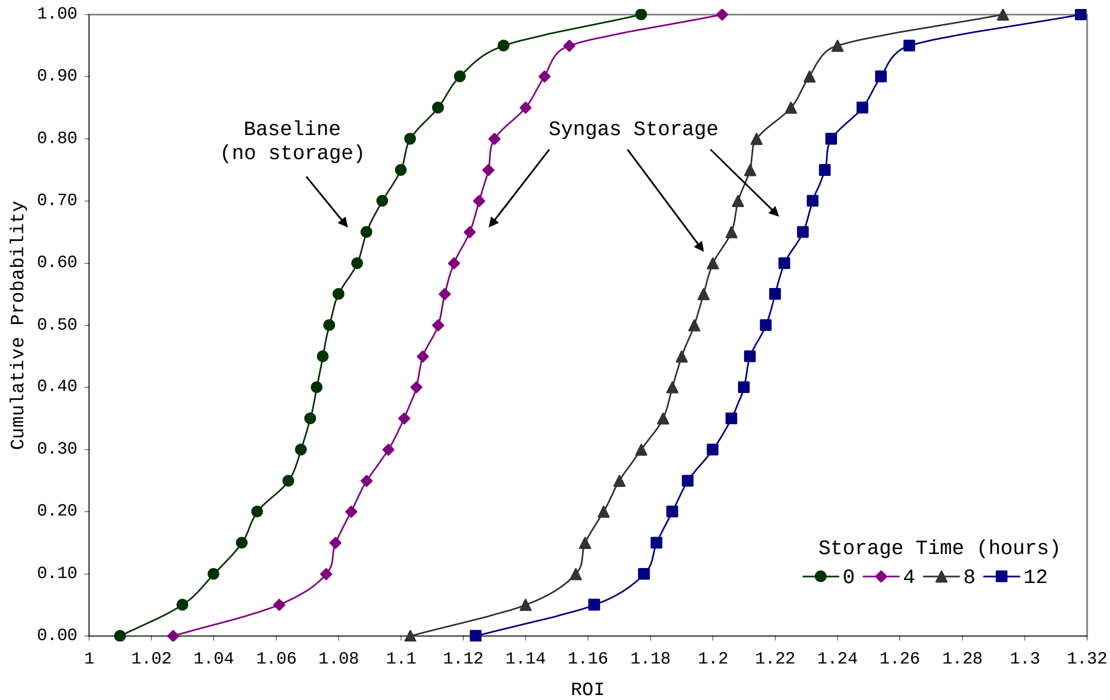

Figure S9. ROI for syngas storage scenario using a 3+1 IGCC facility

As the figure illustrates, the mean ROI for the baseline facility with no storage is 1.08. This is significantly larger than the mean ROI for the smaller IGCC facility with one operating gasifier and one spare gasifier, suggesting that the larger facility would be more profitable, and more likely to be built. Additionally, the figure shows that adding syngas storage and producing peak power increases the ROI for the facility. The addition of 4, 8 and 12 hours of syngas storage increases the mean ROI by 3, 11 and 14 percentage points, respectively.

Results of other facility configurations

IGCC facilities of different sizes and configurations were investigated. Similar results to the 1+0 scenario presented in the paper were found.

Table S9. 1+0 Baseline Facility with CCS Capital and Operating and Maintenance Costs [25]

Component	Capital Cost (million \$2005) Triangular (min, mode, max)			O&M Costs Fixed (\$M/y)	O&M Costs Variable
Gasifier	136.0	143.2	150.4	8.1	376[j] (\$/hr)
Air Separation Unit	91.7	96.5	101.3	4.8	
Cold-gas Cleanup	35.4	37.2	39.2	3.4	-235 (\$/hr)
Power block	143.1	150.6	158.1	4.7	
CO_2 capture and transport	132.9	139.9	146.9	5.6	1,839 (\$/hr)
Total (\$ million)	539.1	567.4	595.9	26.6	
(\$/kW)	2,265	2,380	2,500		

Table S10 summarizes the results of the analysis and shows the ROI for the baseline and syngas storage scenarios at the 0.05, 0.5 and 0.95 percent probability levels, along with the net present value at a 30 year economic and loan life for the assumed operating and economic parameters.

Table S10. Baseline and syngas storage scenarios (80% availability, 100% debt financing at 8% interest rate, economic and plant life of 30 years, Cinergy node)

	ROI 90% confidence (min, mid, max)	NPV (\$ million)
Baseline		
no storage	(0.89, 0.92, 0.95)	-89
Syngas Storage		
4 hours	(0.92, 0.94, 0.97)	-70
8 hours	(1.00, 1.02, 1.06)	30
12 hours	(1.05, 1.07, 1.11)	92

[j] Excluding coal costs

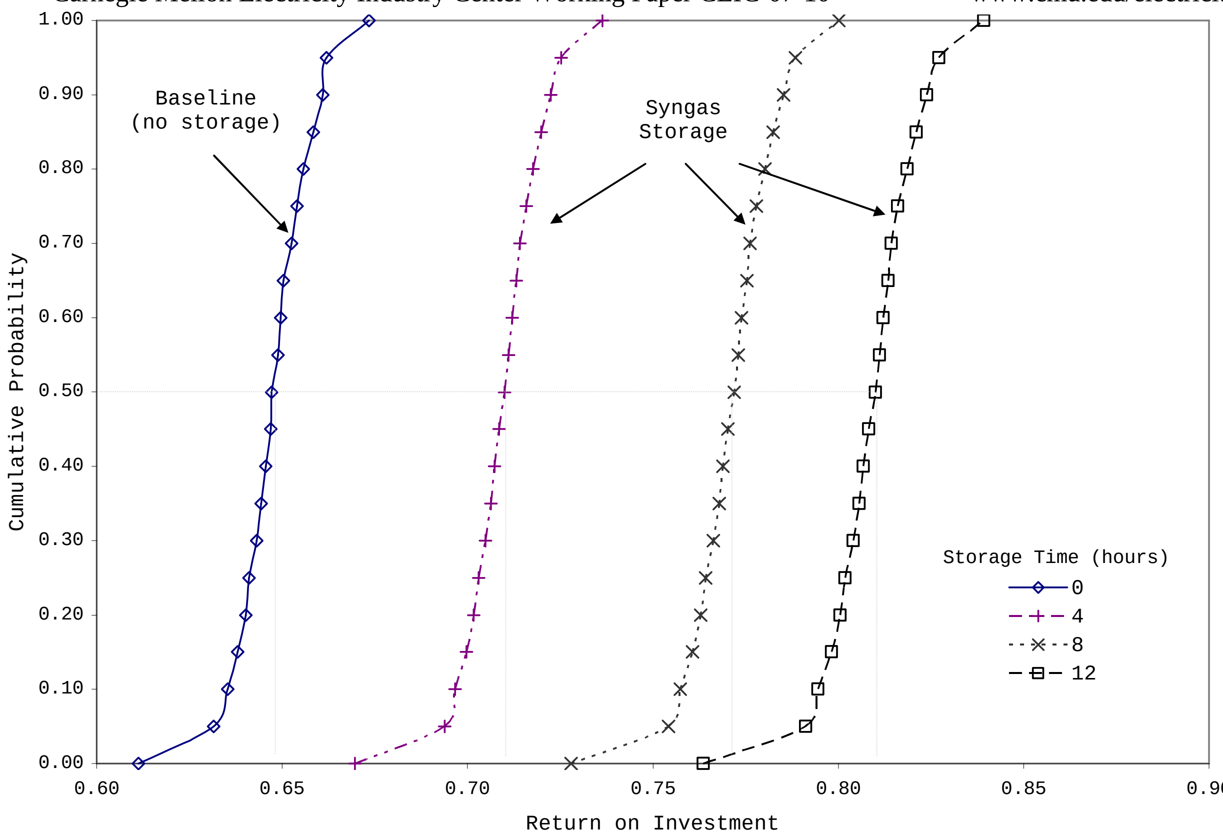

Figure S10. 1+1+ccs. EIA 2007 coal price, 80% availability, 100% financing, 8% interest rate, 35 bar storage pressure, Cinergy node

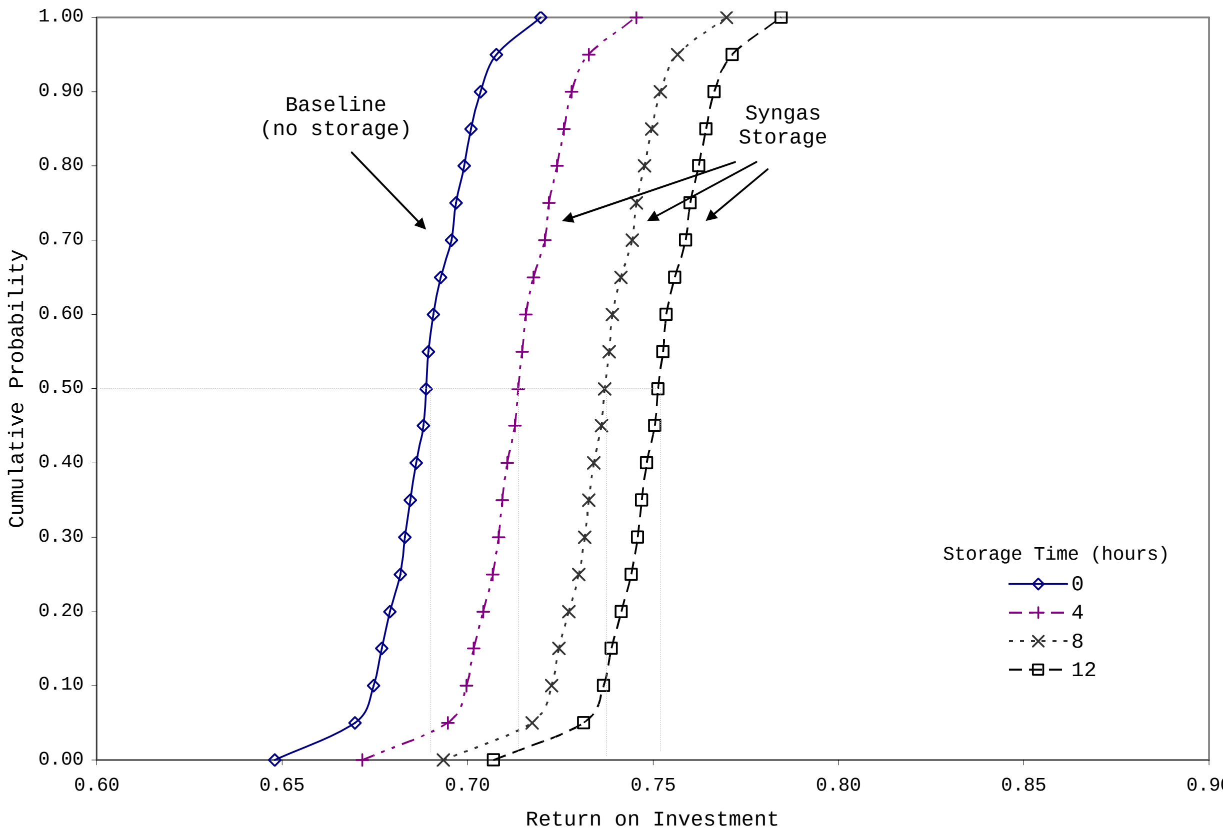

Figure S11. 3+1+ccs EIA 2007 coal price, 80% availability, 100% financing, 8% interest rate, 35 bar storage pressure, Cinergy node

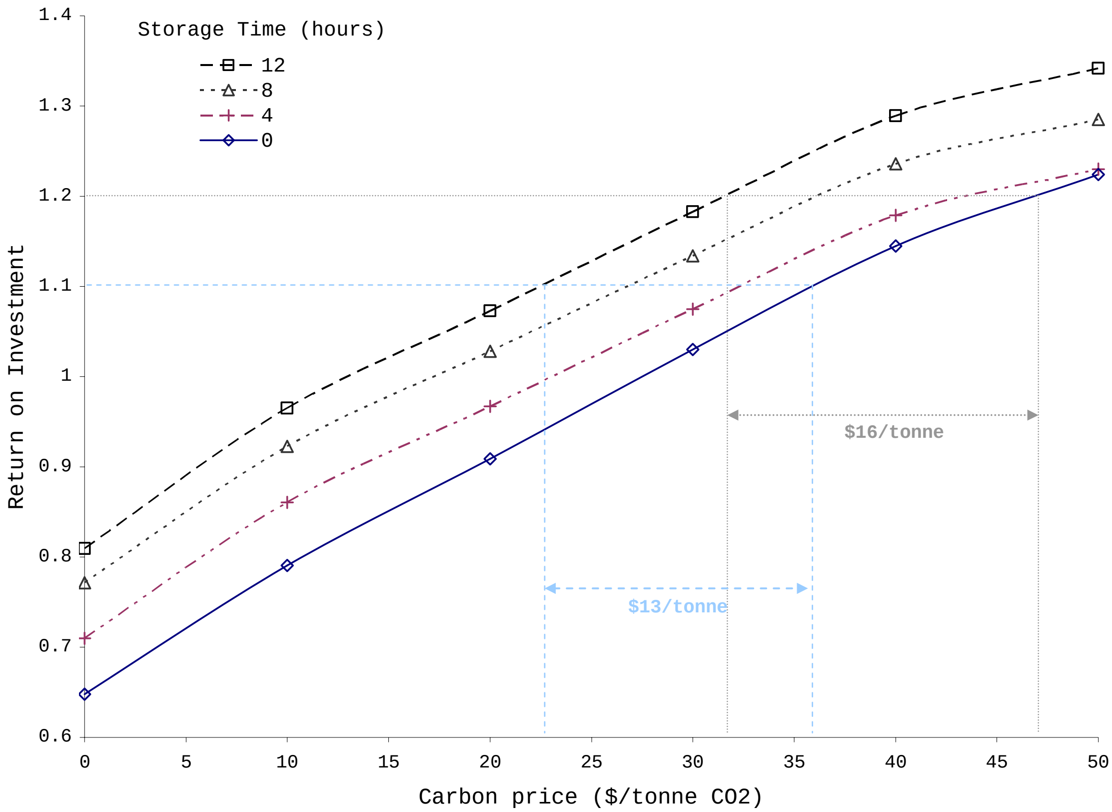

Figure S12. 1+1+ccs EIA 2007 coal price, 80% availability, 100% financing, 8% interest rate, 35 bar storage pressure, Cinergy node

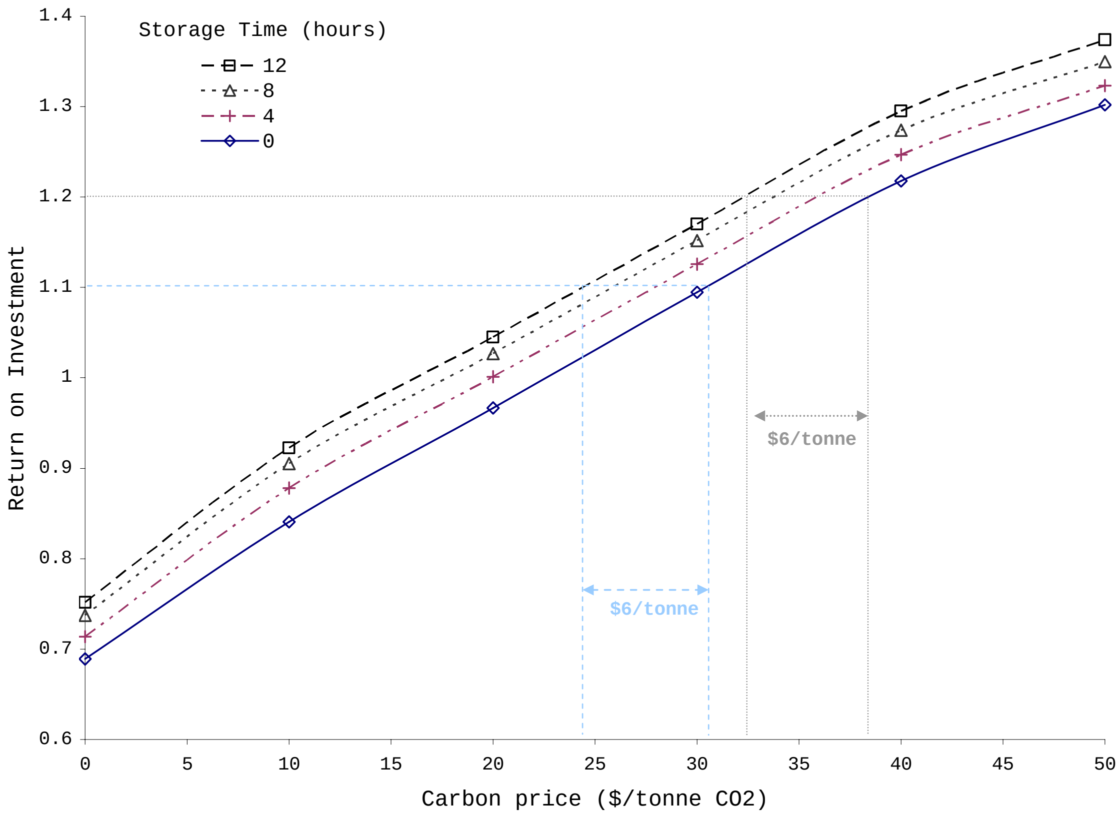

Figure S13. 3+1+ccs EIA 2007 coal price, 80% availability, 100% financing, 8% interest rate, 35 bar storage pressure, Cinergy node

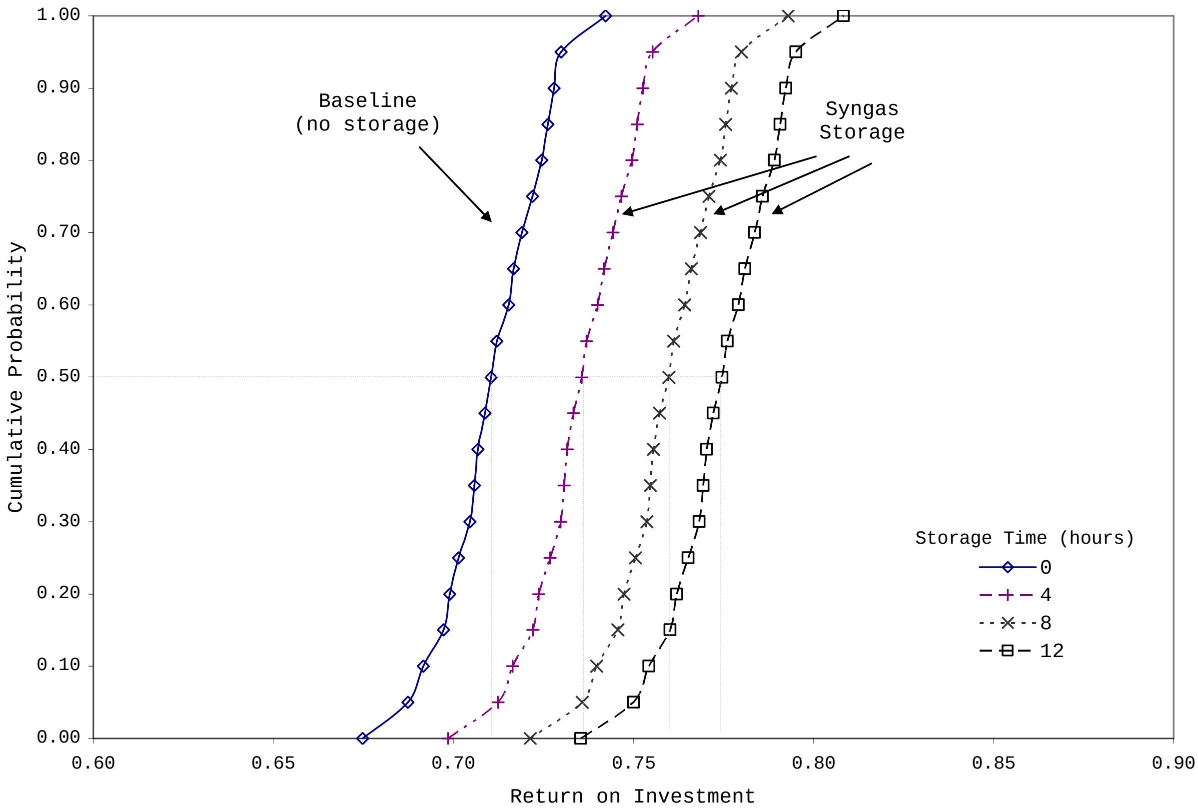

Figure S14. 3+0+ccs EIA 2007 coal price, 80% availability, 100% financing, 8% interest rate, 35 bar storage pressure, Cinergy node

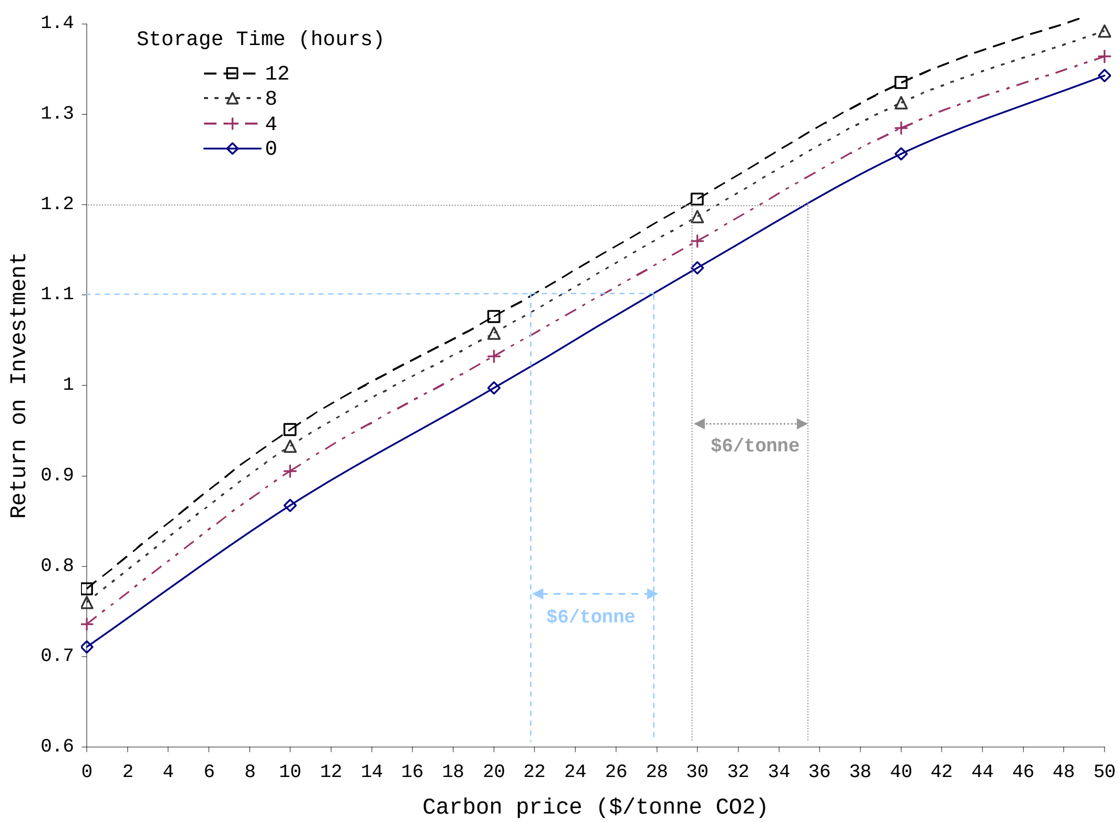

Figure S15. 3+0+ccs EIA 2007 coal price, 80% availability, 100% financing, 8% interest rate, 35 bar storage pressure, Cinergy node

Table S11. Reported syngas compositions

Facility	Wabash [26]	Wabash [26]	Dow [27] Plasquemine	Elcogas [27] Puetrollano	Nuon [27] Power	Polk [28]	El [28] Dorado	Schwarze [28] Pumpe	Exxon [28] Singapore	Eskom [29, 30]
Feedstock	Coal	Petcoke	Coal	Coal/ Petcoke	Coal/ Biomass	Coal	Petcoke	Lignite/ Waste	Fuel Oil	Coal
Gasifier	E-Gas	E-Gas	Dow	Shell	Shell	GE/Texaco	GE/Texaco	BG/Lurgi	GE/Texaco	ErgoExergy
Composition (% vol)										
Carbon Monoxide	45.3	48.6	38.5	29.2	24.8	46.6	45.0	26.2	35.4	8.3
Hydrogen	34.4	33.2	41.4	10.7	12.3	37.2	35.4	61.9	44.5	6.7
Carbon Dioxide	15.8	15.4	18.5	1.9	0.8	13.3	17.1	2.8	17.9	9.5
Methane	1.9	0.5	0.1	0.01	--	0.1	0.0	6.9	0.5	1.0
Argon	0.6	0.6	--	0.6	0.6	2.5	2.1	1.8	1.4	n/r
Nitrogen	1.9	1.9	1.5	53.1	42.0					
Sulfur, ppmv	68	69	n/r	n/r	n/r	n/r	n/r	n/r	n/r	n/r
Water				4.2	19.1	0.3	0.4	--	0.44	17.0
HHV, Btu/scf	277	268	n/r	n/r	n/r					
LHV, Btu/scf						253	242	317	241	150

n/r, not reported

Sensitivity of ROI to Other Factors

The sensitivity of the analysis to variations in the parameters was analyzed.. Figure 5 shows the change in ROI for the 12 hour storage scenario due to a $\pm$10 percent change in the value of the parameters.

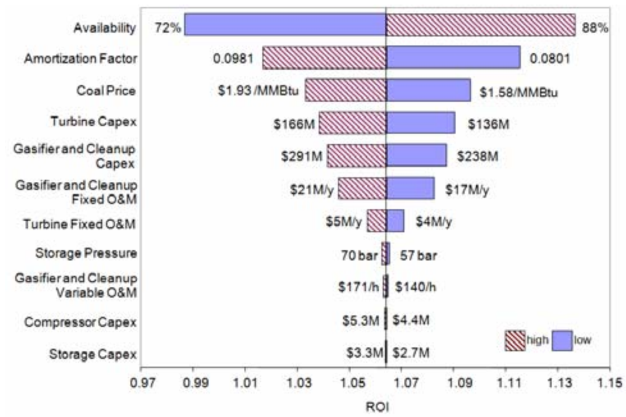

Figure S16. Sensitivity analysis for the 1+0 scenario with 12 hour storage, 10% variation in the parameters. Other parameters as in figure 3.

The ROI for the 12 hour storage scenario is sensitive to the gasifier availability, structure of the financing, price of coal, and capital costs of the turbines, gasifier, air separation unit, and cleanup processes. The gasifier availability and the financing are the most important parameters over which the facility developer or operator has control.

The syngas storage scenario is economically attractive if the revenues received for selling electricity at a high price exceed the additional capital and operating expenses incurred to install and operate the storage equipment (extra turbine, storage vessel and compressor).

To be profitable, the annual increase in revenue must be larger than the levelized costs of the additional storage equipment.

$$\mathrm{Revenue}_{\mathrm{increase}} > \mathrm{Expense}_{\mathrm{increase}}$$

The increase in revenue due to syngas storage is

$$\text{Revenue}_{increase} = \text{Revenue}_{storage} - \text{Revenue}_{no\ storage}$$

Annual revenues depend on the hourly electricity prices. We use all 8760 hours in the analysis to find an exact solution, however, to get an idea of the important factors in the analysis, we can look at the average prices. For a 12 hour storage scenario, using average prices

$$\overline{LMP} = \frac{\sum_{j}^{365} \sum_{i}^{24} LMP_{ij}}{365 \cdot 24}$$

$$\overline{LMP_{high}} = \frac{\sum_{j}^{365} \sum_{i=12}^{24} LMP_{ij}}{365 \cdot 12}$$

$$\overline{LMP_{low}} = \frac{\sum_{j}^{365} \sum_{i=1}^{12} LMP_{ij}}{365 \cdot 12}$$

The annual revenues with no syngas storage and with syngas storage are then:

$$\text{Revenue}_{no\ storage} = 365 \cdot 24 \cdot F_{avail} \cdot MW_1 \cdot \overline{LMP}$$
$$\text{Revenue}_{storage} = 365 \cdot 12 \cdot F_{avail} \cdot LMP_{high} \cdot (MW_1 + MW_2)$$

where F_{avail} is the availability of the facility, MW_1 is the size of the turbine and MW_2 is the size of the peaking turbine.

The increase in revenue due to syngas storage is then

$$\text{Revenue}_{increase} = 365 \cdot 12 \cdot F_{avail} \cdot \left(MW_2 \cdot \overline{LMP_{high}} - MW_1 \cdot \overline{LMP_{low}}\right)$$

If MW_1=MW_2, such as in the analysis presented

$$\text{Revenue}_{increase} = 365 \cdot 12 \cdot F_{avail} \cdot MW \cdot \left(\overline{LMP_{high}} - \overline{LMP_{low}}\right)$$

The annual expenses for the syngas storage scenario are the same as for the non storage scenario with the addition of the annualized capital costs for the additional equipment and the annual operating costs for that equipment

$\text{Expense}_{\text{no storage}}$ = Levelized $\text{Expense}_{\text{no storage}}$

$\text{Expense}_{\text{storage}}$ = Levelized $\text{Expense}_{\text{no storage}}$ + Amortization factor × (Storage vessel capital + Peaking turbine capital + Compressor capital) + Storage vessel O&M + Compressor O&M

where the amortization factor is a function of the interest rate, *i* and debt term, *n*

$$\text{Amortization factor} = i/(1-(1+1)^{-n})$$

The increase in expenses is then

$$\text{Expense}_{\text{increase}} = \text{Expense}_{\text{storage}} - \text{Expense}_{\text{no storage}}$$

$\text{Expense}_{\text{increase}}$ = Amortized (Storage vessel capital + Peaking turbine capital + Compressor capital) + Storage vessel O&M + Compressor O&M

So, to be profitable, the annual increase in revenue must be larger than the levelized costs of the additional storage equipment.

$$\text{Revenue}_{\text{increase}} > \text{Expense}_{\text{increase}}$$

$365 \cdot 12 \cdot F_{avail} \cdot MW \cdot \left(\overline{LMP_{high}} - \overline{LMP_{low}}\right) >$ Amortized (Storage vessel capital + Peaking turbine capital + Compressor capital) + Storage vessel O&M + Compressor O&M

As the equation shows, the gains from using syngas storage depend on the differences in electricity prices at peak and off peak hours for every hour the facility is operated. The mean prices used in the closed form solution do not capture the ‘peakiness’ of the price duration curves and may be of limited use. Because of this, the actual 8760 hours of electricity price are used in the analysis.

Technical and Engineering Considerations

Implementing syngas storage efficiently and cost-effectively in an IGCC facility requires detailed engineering analysis that is beyond the scope of this paper. Engineering issues that have been identified and that should be addressed for successful operation of an IGCC facility with syngas storage follow.

• Humidification and reheating of stored syngas and the implications on thermal plant efficiency.

• Integration and optimization of potential future hot/warm syngas cleaning technologies where the syngas is maintained at a high enough temperature to keep it humid (greater than 500ºF).

• Stability of syngas for long term storage and investigation of potential deposits on the storage vessel.

• Potential effects of short term operating periods for the gas turbine. In the analysis the IGCC plant gasifier operates continuously, but the gas are both operated with potentially several short operating periods each day – as short as 1 hour in the report example. Although gas turbines are commonly used for peaking applications, (the size-weighted average capacity factor for the 884 operating gas turbines in eGRID 2004 was 0.29) such transient gas turbine operation may lead to increases plant maintenance. Data on thermal cycling limits for turbines was not available. The design of a facility using syngas storage should consider the specific turbine manufacturer's cycling limits during the design process. For syngas storage times that the analysis shows is most economically favorable (8 and 12 hours) short cycling is less of a concern. For 12 hours of storage, peak hours are generally during the day, and the turbine is operated continuously over this period.

• The degree of integration between the air separation unit and the gas turbines and the implications for NOx control in the peaking turbine. In a fully integrated IGCC facility, nitrogen from the plant air separation unit is used as a diluent to control NOx emissions. In the configuration used in the present analysis this method of NOx control would not be feasible. A site specific engineering solution would be needed for a real world application.

To envelope the costs for NOx control for the peaking turbine we consider three options: 1) a second air separation unit is constructed and operated solely for the purpose of supplying nitrogen as a dillutant to the peaking turbine; 2) NOx emissions are uncontrolled from the peaking turbine and emission allowances are purchased; and 3) steam is injected to lower the flame temperature in the second turbine and reduce NOx emissions. It may also be economically feasible at certain facilities to perform dilution with nitrogen stored from the production by the ASU when the syngas output of the gasifier is being routed to storage rather than to the turbine.

For the additional ASU scenario, a second air separation train is added to the facility and operated to provide nitrogen to the second peaking turbine. The produced oxygen is not used, or sold, rather vented to the atmosphere. We consider this approach to be an extreme worst case design scenario; it is likely that a fully engineering design analysis would lead to a more efficient and less wasteful design. Adding another train of equal size to the ASU to accommodate the second turbine adds $96.5 million in capital costs, $2.1 million per year in fixed operating costs and consumes, or reduces the net output of the facility by, 30.59 MW [31]. The return on investment is shown in Figure S17.

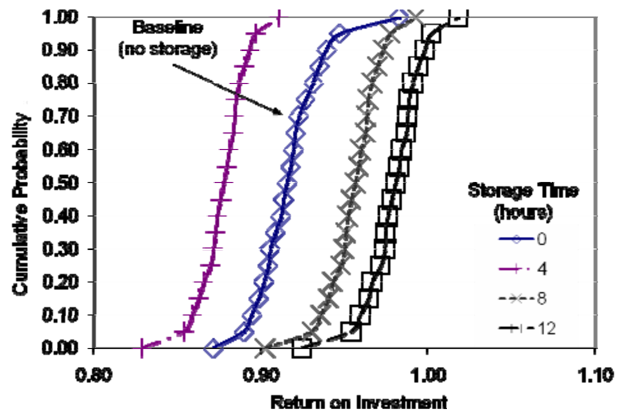

Figure S17. 1+0 with 2 trains of air separation unit for NOx control

The additional gains in ROI from adding syngas storage are reduced by the addition of a second ASU train for NOx control. However, despite the additional cost, adding 8 and 12 hours of syngas storage increases the median ROI over the no storage scenario by 4.5% and 6.5%, respectively.

A second way to envelope the costs of NOx control is to simply leave the peaking turbine uncontrolled and pay for NOx emission allowances. Uncontrolled NOx emissions from a GE 7FA turbine are 8 lbs/MWh [32]. The US EPA reports the cost of (vintage 2008) NOx permits at about $2,500 per ton [33]. The purchase of NOx emissions for the peaking turbine would cost about $2,600 per hr of peaking turbine run time. The resulting ROI is shown in figure S18.

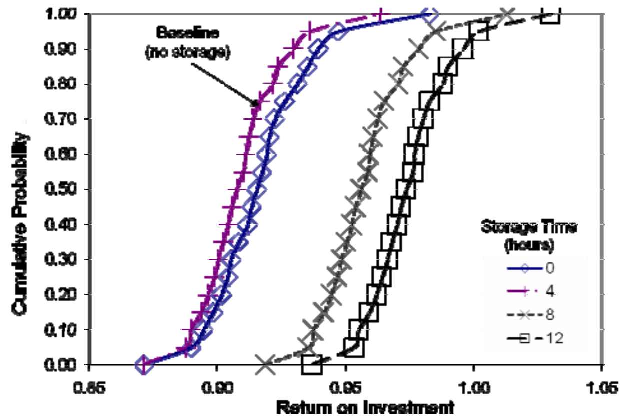

Figure S18. 1+0 with the purchase of NOx allowances for the peaking turbine

The additional gains in ROI from adding syngas storage are reduced when NOx emissions allowances are purchased. However, despite the additional cost, adding 8 and 12 hours of syngas storage increases the median ROI over the no storage scenario by 4.1% and 5.8%, respectively.

A third method of enveloping the costs of NOx control was to consider the losses associated with steam injection into the gas turbine. Directing a portion of the steam into the gas turbine results in a lower thermal efficiency; values in the literature suggest that this reduction will be approximately 5% [34, 35] when the second turbine is run. The effect of these thermal losses is to lower the output of the facility. The ROI for 4, 8 and 12 hours with the steam injection energy penalty was 0.93, 0.99 and 1.03, respectively. Despite the reduced output, adding 8 and 12 hours of syngas storage increases the median ROI over the no storage scenario by 8% and 11%, respectively.

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