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Decarbonised Coal Energy System Advancement through CO_2 Utilisation and Polygeneration

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Abstract

Development of clean coal technology is highly envisaged to mitigate the CO_2 emission level while meeting the rising global energy demands which require highly efficient and economically compelling technology. Integrated gasification combined cycle (IGCC) with carbon capture and storage (CCS) system is highly efficient and cleaner compared to the conventional coal-fired power plant. In this study, an alternative process scheme for IGCC system has been proposed, which encompasses the reuse of CO_2 from the flue gas of gas turbine into syngas generation, followed by methanol synthesis. The system modification requires extensive mass and energy integration strategies to ensure that the efficiency and economics of the system are achieved to a considerably high level. The thermodynamic and economic feasibilities are found to attain significant improvement through the realisation of a suitably balanced polygeneration scheme. The economic potential can be enhanced from negative impact to 317 M€/y (3.6 €/GJ). The results have demonstrated promising prospects of employing CO_2 reuse technology into IGCC system, as an alternative to CCS system.

Keywords: CO_2 reuse; Carbon capture and storage; IGCC; tri-reforming; polygeneration; clean coal technology

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1. Introduction

One of the priority areas identified in Cancún Agreement at the 2010 United Nations Climate Change Conference (COP16) is the low carbon emission future for energy systems (UNFCCC, 2010a). The economic growth has also been recognised as closely associated with energy security. Coal plays an important role in terms of indigenous energy resource for many countries including fastest growing major economy nations, such as China and India. Currently, 40% of the global electricity is supplied from coal and it is expected to increase over the next few decades (World Coal Association, 2011). Coal-fired power plant is the predominant technology for generating electricity from coal. However, the biggest problem is the CO_2 emission, e.g. approximately 2.9 Mt CO_2 per year to the atmosphere from 500 MW_e plant (IPCC, 2005). The energy and industrial sectors, including power station, manufacturing and transportation contribute to 77.9% (2005) of the global CO_2 emission, and 54.8% (2008) of the CO_2 emission in the UK (Prime et al., 2009; World Resource Institute, 2011). The CO_2 emission from coal and other solid fuels shares 25.6% of the total CO_2 emission by fuel in the UK, i.e. 531.8 Mt CO_2 in year 2008 (Prime et al., 2009). The deployment of CCS is highly essential in lowering the greenhouse gas (GHG) emission from the coal-fired power plant. Life cycle assessment has shown that those plants with CCS can achieve 75-84% GHG reduction with reference to a sub-critical pulverised coal power plant at 90% CO_2 capture efficiency. Furthermore, IGCC with CCS can reach 81% reduction in GHG level compared to IGCC without CCS, attaining a low GHG emission level at less than 160 g CO_2e/kWh (Odeh and Cockerill, 2008a, 2008b). In fact, IGCC has higher efficiency than conventional coal-fired power plant through the application of cogeneration concept. It is reported that IGCC without any carbon capture achieves an efficiency of 39-42.1% based on coal HHV (NETL, 2010). IGCC is also cleaner and has high potential in capturing CO_2. The efficiency reduces to 31-33.6% due to CCS (NETL, 2010). It is thus highly imperative to initiate the research activities within the scope of decarbonised polygeneration from fossil fuels as well as system enhancement through process integration (Klemeš et al., 2007; Bulatov and Klemeš, 2009; Hetland, 2009; Klemeš and Friedler, 2010; Adams and Barton, 2011).

Carbon capture technologies such as pre-combustion, post-combustion and oxy-fuel combustion are prominent (Kanniche et al., 2010). Other emerging technologies such as chemical looping and oxygen transport membrane are undergoing rapid development. The inclusion of carbon capture facilities normally increases the overall capital investment and lowers the energy efficiency of a plant (Harkin et al., 2009; Ng et al., 2010). Captured CO_2 is transported through pipelines and ships, and subsequently stored in ocean for geological formation or mineral carbonates. The series of processes of capturing, transporting and storing CO_2 is collectively known as carbon capture and storage (CCS). Other options of mitigating CO_2 emission is through CO_2 reuse. Such options include utilising CO_2 into Enhanced Oil Recovery (EOR) in oil extraction process; microalgae production; chemicals and fuels production (Li et al., 2006; Abidin et al., 2011). The question of which CO_2 mitigation options, i.e. whether to capture and store CO_2 or reuse CO_2 without capturing, is more advantageous than the others remains uncertain. Furthermore, CO_2 reuse process as well as the associated integration to the existing energy system is under explored.

In this study, a conventional coal IGCC with CCS system, generating electricity as the sole product through cogeneration concept has been used as the base case. This system can be modified into a polygeneration system, wherein CO_2 from the flue gas of gas turbine is reused into syngas generation through tri-reforming process. The syngas is subsequently converted into methanol. This system does not involve pre-capturing CO_2. Such modified system can be regarded as an integrated dual syngas production system, comprising of a coal to power and a natural gas to liquid fuel process. These two systems with different CO_2 mitigation options are compared in terms of thermodynamic, economic and environmental performances.

Additionally, heuristic-based heat integration methodology (Smith, 2005; Ng et al., 2010) has been adopted for achieving maximum energy savings from the system and thus ensuring maximum economic benefit.

2. Methodology

Process flowsheet simulation in ASPEN Plus is undertaken for the modelling of IGCC systems. Heat integration (section 2.1) and economic analysis (section 2.2) are performed in Excel spreadsheet, using data extracted from the mass and energy balances obtained from the simulation.

2.1 Heat Integration Strategies

Important thermodynamic data such as temperature and heat duties across heat exchangers and process units are extracted from the flowsheet simulation. Screening and classification of these data are performed to ensure appropriate utilisation of heat at various levels. The heat supply and demand within the system are categorised into high and low levels based on temperature and heat duties. In other words, high temperature and / or high heat duty process units are utilised for high level tasks, i.e. steam generation, whilst low temperature and / or low heat duties process units are utilised for low level tasks, i.e. process-to-process heating or hot water generation. The composite curve analysis and energy balance are carried out to estimate the amount of steam that can be generated and the amount of steam requirement for heating. If a high level task is found to be inappropriate after performing the analysis, e.g. negligible amount of steam is generated or too much steam has to be used for heating, screening and classification procedures are repeated and extraction of data is revised. Process stream matching and energy balance are adopted for analysing low level tasks. The proposed strategy considers a high to low level approach, since any excess heat can be used into hot water generation and this is normally less likely to violate the minimum approach temperature rule. The final step is the design of combined heat and power network (steam generation and distribution), based on the information obtained from the composite curve and energy balance analyses. Steam is generated and collected at various steam mains, e.g. VHP, HP, MP and LP. Steam is distributed from the steam mains to process units / heat exchangers within a process site. Remaining steam from each steam main level can be expanded through steam turbine into power and low level steam generation.

2.2 Economic Analysis

Capital cost of the system is evaluated by taking the direct and indirect capital costs into account. The cost of equipment is estimated using the cost and size correlation given in equation (1). The cost estimation parameters (Hamelinck and Faaij, 2002; Denton, 2003; IPCC, 2005; Larson et al., 2005) are given in Appendix A. The costs associated with the equipments are levelised to the current cost by the inclusion of Chemical Engineering Plant Cost Index (CEPCI), i.e. CEPCI = 555.2 (April 2010) using equation (2). The discounted cash flow method is applied for determining the annual charge for the capital investment.

$$\frac{\mathrm{COST}_{\mathrm{size2}}}{\mathrm{COST}_{\mathrm{size1}}} = \left(\frac{\mathrm{SIZE}_2}{\mathrm{SIZE}_1}\right)^{\theta} \qquad (1)$$

SIZE_1 and $\mathrm{COST}_{\mathrm{size1}}$ represent the capacity and the cost of a base unit, whilst SIZE_2 and $\mathrm{COST}_{\mathrm{size2}}$ represent the capacity and the cost of the unit after scaling up/down, respectively. θ is the scale factor.

$$\text{Present cost} = \text{Original cost} \times \left(\frac{\text{Index at present}}{\text{Index when original cost was obtained}}\right) \qquad (2)$$

The operating cost is evaluated in terms of fixed and variable costs. The parameters for estimating the operating costs (Tijmensen et al., 2002; Sinnott, 2006; DECC, 2010) are given in Appendix A.

An annualised charge of 11% is determined by using the following assumptions:

- Discount rate: 10%
- Plant life: 15 years
- Start-up period: 3 years (20%, 45%, 35%)

The economic potential (*EP*) of the system is then determined using equation (3).

$$EP = H\sum_{i=1}^{i=NP} r_i p_i - CC - OC \tag{3}$$

H is the total number of operating hours per year (8000 hours is assumed); r_i and p_i are the production rate and unit price of product *i*, respectively; *NP* is the total number of products; *CC* and *OC* are annual capital cost and annual operating cost, respectively.

The current market prices / estimated costs of production are identified for evaluating the total value of the products, i.e. electricity (74.14 Euro/MWh (DECC, 2010)) and methanol (255 Euro/t (Methanex, 2010)).

3. Existing and Alternative IGCC Process Schemes

3.1 Process Description

Scheme A - Coal IGCC with CCS

The IGCC system under consideration has a capacity of 648 MW, with a coal throughput of 2000 t/d. The ASPEN Plus simulation model for the IGCC with CCS system is illustrated in Figure 1. The types of models and specifications are summarised in Table 1. This is a conventional process scheme where coal slurry is gasified (GASIFIER) using oxygen as a gasifying medium to generate syngas for power generation. The intermediate processes involve gas cooling (SYNGCOOL) as well as a series of gas cleaning and conditioning processes, such as ash removal from CYCLONE, high and low water gas shift reactors (HTWGS and

LTWGS), H_2S and CO_2 capture through H2SREM and CO2SEP. 99% of CO_2 is assumed to be captured and subsequently transported via pipelines to a storage site after compressing to 80 bar. Finally, the clean syngas is sent to gas turbine for power generation. Heat from the exhaust gas of gas turbine is recovered into VHP steam in heat recovery steam generator (HRSG) and the exhaust gas is eventually released into the atmosphere. The heat integration strategies as proposed in section 2.1 are employed to ensure appropriate levels of generation and utilisation of heat and power, aimed at achieving maximum heat recovery from the site.

Figure 1

Table 1

Scheme B - Coal IGCC with tri-reforming and methanol synthesis

An alternative polygeneration scheme B using the same basis as in the IGCC Scheme A (e.g. coal throughput of 2000 t/d), has been proposed, depicted in Figure 2. Parameters used for modelling this process scheme in ASPEN Plus are given in Table 2. The proposed scheme requires a major modification from an original heat and power cogeneration system (Figure 1) into polygeneration system with methanol as an additional product. This involves the utilisation of CO_2 from the exhaust gas of gas turbine into tri-reforming process (equations (4)-(6)) (Song, 2001; Song and Pan, 2004), to further generate syngas. The exhaust gas from the gas turbine contains 64 mol% CO_2, 34 mol% H_2O and 2 mol% inert gases. This modified system can be visualised to have dual syngas processing routes, the first route is aimed at electricity generation, whilst the second route is targeted into methanol production.

$$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2 \qquad \Delta H_R^\circ = +247.3\ \text{kJ/mol} \qquad (4)$$

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 \qquad \Delta H_R^\circ = +206.3\ \text{kJ/mol} \qquad (5)$$

$$CH_4 + 0.5O_2 \leftrightarrow CO + 2H_2 \qquad \Delta H_R^\circ = -35.6\ \text{kJ/mol} \qquad (6)$$

Figure 2

Table 2

The methane tri-reforming process was first implemented by Song in 2001 as a potential method to utilise CO_2 into the production of valuable syngas at a desired ratio and to reduce or eliminate carbon formation on catalyst (Song, 2001; Song and Pan, 2004). Tri-reforming process fed with CH_4, CO_2, H_2O and O_2 at a ratio of 1: 0.475: 0.475: 0.1 is operated at 1 bar and 850°C (Song and Pan, 2004). The syngas produced from the tri-reforming process comprises of 59 mol% H_2, 3 mol% H_2O, 36 mol% CO and 2 mol% CO_2, thus providing a H_2/CO molar ratio of 1.6.

This scheme demonstrates a system with the CO_2 reuse from the flue gas without pre-capture. H_2 is instead separated from CO_2 via a pressure swing adsorption process. 98% by mole of H_2 is assumed to be separated from the product gas stream (from gasification) and combined with the product gas from the tri-reforming process. The remaining syngas from gasification after separating H_2 contains significant amount of CO, which is then used into power generation via gas turbine. A small amount of natural gas is needed to manipulate the Wobbe Index of the gas turbine, since only small amount of H_2 is present in the inlet gas to the gas turbine combustor. Oxygen instead of air is used in the gas turbine combustor for avoiding further dilution of the fuel gas by nitrogen, and thus to avoid accumulation of nitrogen in the downstream process (tri-reforming and methanol synthesis) incurring additional capital cost. This is similar to the oxy-fuel combustion concept and it has advantages such as concentrating the CO_2 in the exhaust gas stream and reducing NO_x emission (Figueroa et al., 2008). Methanol reactions require a feed with $(H_2-CO_2) / (CO+CO_2)$ of 2. 95% by volume take place at 100 bar and 250°C. The unreacted offgas from methanol synthesis reactor after 5% purged is recycled to enhance the production of methanol. The optimal methanol synthesis reactor operating conditions are based on our previous study (Ng and Sadhukhan, 2011). The liquid methanol is sent to distillation units, where 99.5% by weight of methanol can be recovered. Introduction of natural gas to the tri-reforming process enhances the overall capacity of Scheme B by ~5 times compared to Scheme A.

3.2 Performance Analysis

The performances of IGCC and polygeneration process schemes, with respect to thermodynamic efficiency and economic potential are evaluated and compared, in Tables 3 and 4, respectively.

Table 3

Table 4

44.7 t/h of CO_2 is emitted and 141.9 t/h of CO_2 is captured in Scheme A, while 52.1 t/h of CO_2 is emitted and 216.8 t/h of CO_2 is reused in Scheme B. Both schemes can achieve CO_2 reduction of 76-80%.

4. Discussion

Section 3.2 demonstrates promising outcome by the transformation of cogeneration system into polygeneration system. The efficiency can be improved from 36% (Scheme A) to 86% (Scheme B) (Table 3). The modification also involves an expansion into a secondary syngas processing route from natural gas feedstock to tri-reforming process. The capacity is increased from 648 MW to 3450 MW. The advantage of Scheme B is that a substantial amount of methanol is produced increasing the overall value of products that can offset the increased capital and operating costs of natural gas utilisation in tri-reforming process. Nevertheless, the economic potential can be significantly improved from −13 M€/y (Scheme A) to 317 M€/y (Scheme B) (Table 4). Furthermore, the CO_2 emission per unit product from Scheme B, 16.9 t CO_2/GWh, is lower than that from Scheme A, 127.8 t CO_2/GWh. These imply 86% reduction in the plant greenhouse gas emissions and that Scheme B is thermodynamically and economically more promising compared to an equivalent coal IGCC system with CCS.

The CO_2 reuse scheme however, does not save the total emission across life cycle, because eventually the products are consumed. Therefore, there is no clear-cut decision on which CO_2 mitigation option is more superior to the others. The key consideration is the economic and environmental policy and acceptance of these technologies. CO_2 can be stored underground for up to hundreds and thousands of years using CCS options. In particular, CCS has the benefits of terminating the CO_2 life cycle. However, the leakage of CO_2 from the storage reservoir can also be severe and the effect to the ecological system can be devastating. The risks are not fully understood yet. The CO_2 reuse, on the other hand, has the advantage of delaying the CO_2 emission to the atmosphere as well as prolonging the CO_2 life time depending on the consumption route of the final product of CO_2. Converting CO_2 into a polymer may be a better option than a fuel since the carbon can be retained in polymers for a long period of time. Flexibility in product generation and system modification are amongst the desired criteria for future energy systems. Scheme B provides an indirect CO_2 utilisation platform (CO_2→syngas→product), has the advantage of generating different products according to different market needs. This is because syngas is versatile in various applications, such as Fischer-Tropsch liquid, dimethyl ether and other chemical production.

It is recognised that the coal cogeneration has a significant contribution to the security of electricity supply in the UK (ScottishPower, 2008). The ScottishPower has engaged into a demonstration project that uses Scottish coal and biomass co-firing technology integrated with advanced CCS options. The large scale plant yet to be exploited is an example of moving forward step towards flexible generation needed to support the UK's growth goals in renewable energy and at the same time ensures security of supply. It is also assessed that to follow a low carbon energy trajectory in order to restrict the temperature rise up to $2^{o}C$ over this century, the

Annex I countries must switch to non-fossil resources or only use fossil resources with CCS and with other renewable resources for complete decarbonisation of the sector, within years. At the same time, the Non-Annex I countries have also committed towards adaptation to alternative technologies to combat against climate change (UNFCCC, 2010b). Thus, the techno-economic feasibility of the polygeneration scheme considered here offers low carbon technology solution in the interim period (e.g. before 2020). Thus, converting CO_2 into fuel is also a promising route to resolve the issues associated with the rising energy demand at present.

5. Conclusions

Alternative route to mitigating CO_2 emission by CO_2 reuse is analysed in terms of thermodynamic efficiency, economic and environmental impacts, and compared with a cogeneration IGCC route using CCS. Simultaneous process modification, mass and energy integration as well as economic and environmental analyses of an overall system are imperative for the synthesis of efficient and economically appealing system. Reusing CO_2 can be beneficial in reducing the amount of CO_2 that needs to be captured. Also the process economics can be enhanced through the generation of additional product under the current economic drives. There is no clear-cut answer to which options should be adopted. It is suggested that other factors such as global energy demand, economic aspects as well as government policies should be taken into consideration.

Nomenclatures

CC	Annual capital cost
EP	Economic potential
H	Total number of operating hours per year
NP	Total number of products
OC	Annual operating cost

p_i	Unit price of product *i*
r_i	Production rate of product *i*
θ	Scale factor, equation (1)

Appendix A

The economic parameters required for evaluating capital and operating costs are presented in Tables A.1 and A.2, respectively.

Table A.1

Table A.2

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Table 1: Coal IGCC with CCS process specification in ASPEN Plus simulation.

'Compr' = Compressor / turbine; 'Sep' = Component separator; 'RGibbs' = Gibbs reactor; 'REquil' = Equilibrium reactor; 'Heater' = Heater; 'Mixer' = Stream mixer; 'SSplit' = Substream splitter; 'Pump' = Pump.

Unit	ASPEN Plus model	Outlet Temperature (°C)	Pressure (bar)	Other Specification
AIRCOMP	Compr		14	Isentropic efficiency = 0.9
CO2COMP	Compr		80	Isentropic efficiency = 0.9
CO2SEP	Sep			CO_2 split fraction = 0.99
CYCLONE	SSplit			Ash split fraction = 1.0
GASIFIER	RGibbs	1371.1	75	
GASTURB	Compr		2	Isentropic efficiency = 0.9
GTCOMB	REquil	1200	14	
H2SREM	Sep			H_2S, Cl_2, COS split fraction = 1.0
HE1	Heater	83.3	47	
HE2	Heater	121.1	42.4	
HE3	Heater	370	15	
HE4	Heater	200	25	
HE5	Heater	35	80	
HE6	Heater	480	25	
HRSG	Heater	100	1	
HTWGS	Requil	370	15	
LTWGS	Requil	200	15	
SLURMIX	Mixer		1	
SLURPUMP	Pump		42.4	Pump efficiency = 0.9
SYNGCOOL	Heater	200	75	
SYNGEXP	Compr		15	Isentropic efficiency = 0.9

Table 2: Additional data for simulating coal IGCC with tri-reforming and methanol synthesis process in ASPEN Plus.

'Sep' = Component separator; 'REquil' = Equilibrium reactor; 'Flash2' = Two-outlet flash.

Process unit	ASPEN Plus model	Outlet Temperature (°C)	Pressure (bar)	Other Specification
H_2O removal column	Flash2	30	15	
Methanol synthesis reactor	REquil	250	100	
Pressure swing adsorption	Sep			H_2 split fraction = 0.98
Tri-reformer	REquil	850	1	
Water-gas shift reactor	REquil	200	15	Steam inlet condition: 14 bar, 250°C

Table 3: Efficiency analysis.

Process Scheme	Scheme A	Scheme B
Product	LHV (MW)	
1. Electricity	237.0	123.4
2. Methanol	-	2852.8
Total LHV of products	**237.0**	**2976.2**
Feed	LHV (MW)	
Main feedstock	Coal	Coal
LHV of main feedstock	648.0	648.0
Additional feedstock	-	Natural gas
LHV of additional feedstock	-	2802.6
Total LHV of feedstock	**648.0**	**3450.7**
Thermal efficiency based on LHV of feedstock (%)	**36.6**	**86.3**

Table 4: Economic analysis.

Process Scheme	Scheme A	Scheme B
Capital cost (M€/y)	86.2	142.9
Operating cost (M€/y)	67.4	655.4
Value of products (M€/y)	140.6	1115.5
1. Electricity	140.6	73.2
2. Methanol	-	1042.3
Economic Potential (M€/y)	**−13.1**	**317.2**
Economic Potential (€/GJ)	**−1.9**	**3.6**

Note: Unit price of electricity = 74.14 Euro/MWh; methanol = 255 Euro/t.

Table A.1: Capital cost parameters.

ISBL					
No.	Process unit	Base Cost (million USD)	Scale factor, θ	Base scale	Scale unit
1	Coal handling [a]	29.58	0.67	2367	t/d coal input
2	Gasifier (GE type) [a]	62.92	0.67	716	MW coal input
3	Cyclone [a]	0.91	0.7	68.7	m^3/s gas feed
4	Water-gas shift reactor [a]	12.24	0.67	1377	MW LHV coal input
5	Rectisol [b, i]	54.1	0.7	9909	kmol CO_2/h
6	CO_2 transport and storage [c]	5.6 Euro/t CO_2			
7	Methanol reactor [b]	7	0.6	87.5	t MeOH/h
8	Methanol separation [b]	15.1	0.7	87.5	t MeOH/h
9	PSA [b]	28	0.7	9600	kmol/h feed
10	Gas turbine [a]	56	0.75	266	MW
11	Steam turbine (inc. condenser) [a]	45.5	0.67	136	MW
12	HRSG [a]	41.2	1	355	MW heat duty
13	SYNGCOOL [a]	25.4	0.6	77	MW heat duty
14	ASU [a]	35.6	0.5	76.6	t O_2/h
15	Compressor [a]	4.83	0.67	10	MW
16	Expander [a]	2.41	0.67	10	MW
17	Tri-reformer [b, ii]	9.4	0.6	1390	kmol/h feed
OSBL [b]					
No.	Specification	Cost estimation (% of ISBL)			
18	Instrumentation and control	5			
19	Buildings	1.5			
20	Grid connections	5			
21	Site preparation	0.5			
22	Civil works	10			
23	Electronics	7			
24	Piping	4			
	Total Direct Capital (TDC)	**ISBL + OSBL**			
Indirect Capital Cost [b]					
No.	Specification	Cost estimation (% of TDC)			
25	Engineering	15			
26	Contingency	10			
27	Fees/overheads/profits	10			
28	Start-up	5			
	Total Indirect Capital (TIC)				
	Total Capital Cost	**TDC+TIC**			

Note:

[a] Larson et al., 2005. Economic parameters taken from year 2003. Assume 1USD = 0.9 Euro (2003).
[b] Hamelinck and Faaij, 2002. Economic parameters taken from year 2001. Assume 1 USD = 1.1 Euro (2001).
[c] IPCC, 2005. Cost of CO_2 transport: 0-5 USD/t CO_2; Cost of CO_2 storage: 0.6-8.3 USD/t CO_2. Average values of CO_2 transport and storage are taken. Assume 1 USD = 0.8 Euro (2010).

[i] Cost of Rectisol is assumed to be 2 times of Selexol, as suggested by Denton, 2003.
[ii] Cost of tri-reformer is assumed to be the same as the cost of steam reformer.

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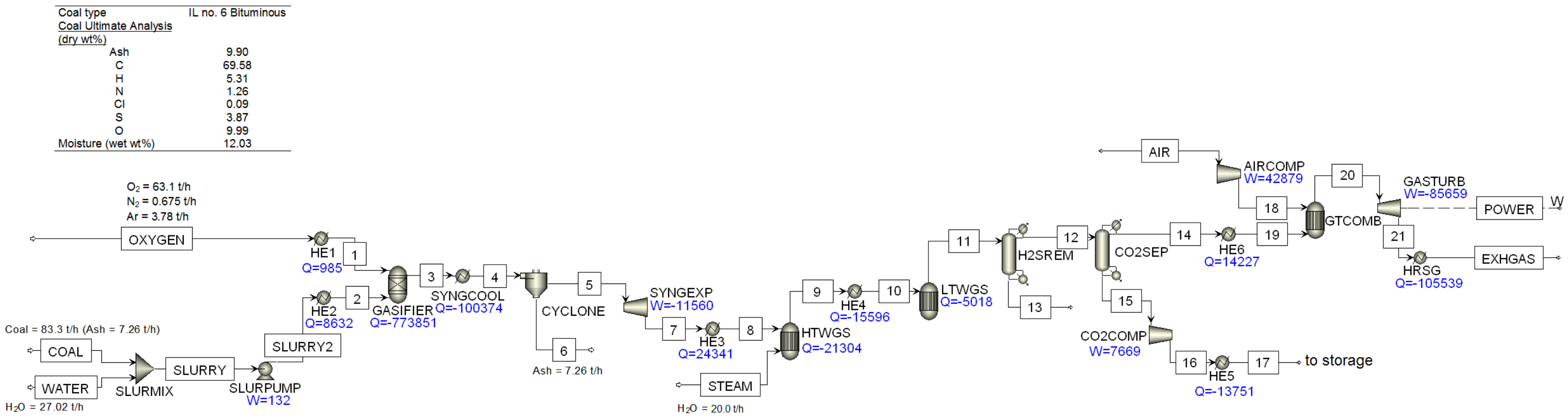
2001= 394.3; 2003=402.0; 2010 (April)=555.2

Table A.2: Operating cost parameters.

Fixed Operating Cost [a]		
No.	Specification	Cost Estimation
1	Maintenance	10 % of indirect capital cost
2	Personnel	0.595 million Euro/100 MWth LHV
3	Laboratory costs	20% of (2)
4	Supervision	20% of (2)
5	Plant overheads	50% of (2)
6	Capital Charges	10% of indirect capital cost
7	Insurance	1% of indirect capital cost
8	Local taxes	2% of indirect capital cost
9	Royalties	1% of indirect capital cost
Variable Operating Cost [b]		
No.	Specification	Cost estimation
10	Natural Gas	20 Euro/MWh
11	Coal	2.4 Euro/GJ
12	Electricity	74.14 Euro/MWh
	Direct Production Cost (DPC)	**Variable + Fixed Operating Costs**
Miscellaneous [a]		
No.	Specification	Cost estimation
13	Sales expense, general overheads, research and development	30% of DPC
	Total OPEX per year	**DPC + Miscellaneous**

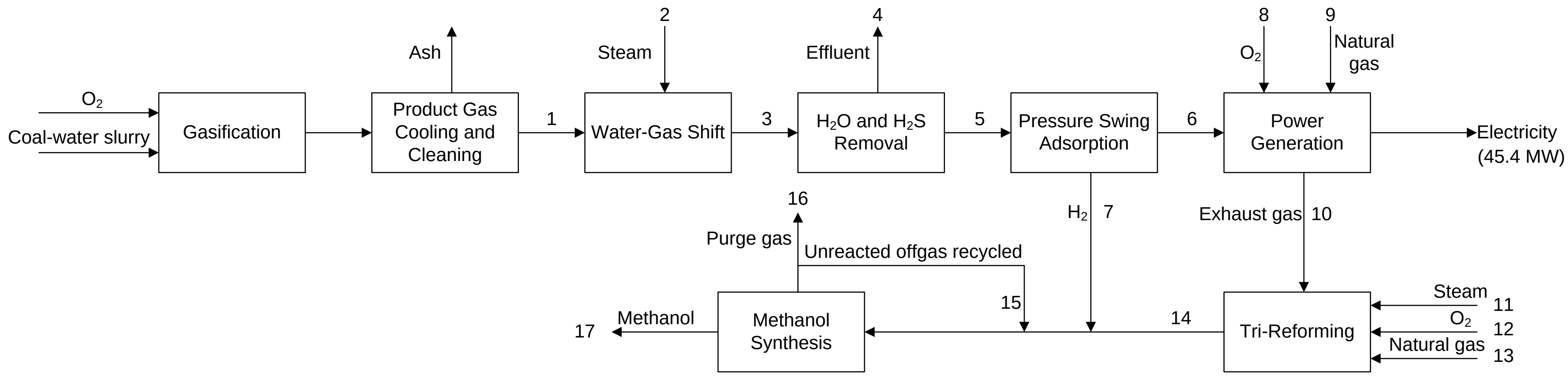
Note:
[a] The parameters except personnel are taken from Sinnott, 2006. Estimation for personnel is taken from Tijmensen et al., 2002.
[b] The variable operating costs for various feedstocks are taken from DECC, 2010.



Q: Heat duty (kW)
W: Power (kW)

Component Mole Fraction	Stream No.							
	3	9	11	13	14	17	AIR	EXHGAS
H_2	0.285	0.473	0.521		0.800			
H_2O	0.180	0.055	0.007		0.011			0.285
CO	0.421	0.151	0.103		0.159			
CO_2	0.084	0.295	0.343		0.005	1.000		0.058
O_2							0.21	0.003
N_2	0.007	0.006	0.006		0.009		0.79	0.649
Ar	0.010	0.010	0.010		0.015			0.005
H_2S	0.011	0.009	0.009	0.929				
Cl_2				0.010				
COS				0.061				
Molar flow rates (kmol/s)	2.33	2.64	2.64	0.025	1.72	0.90	4.00	4.89

Figure 1: Coal IGCC system with CCS (Scheme A) ASPEN Plus simulation model and results.



Component Mole Fraction	Stream No.																
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
H_2	0.285		0.540		0.559	0.025	1.000							0.592	0.671	0.671	0.001
H_2O	0.180	1.000	0.026	0.684	0.003	0.006				0.341	1.000			0.028			0.049
CO	0.421		0.033		0.035	0.076								0.355	0.064	0.064	0.002
CO_2	0.084		0.377	0.060	0.388	0.859				0.639				0.016	0.139	0.139	0.047
CH_4									1.000				1.000	0.005	0.065	0.065	0.004
CH_3OH															0.008	0.008	0.895
O_2								1.000		0.002		1.000					
N_2	0.007		0.005		0.006	0.013				0.006				0.001	0.020	0.020	
Ar	0.010		0.009		0.009	0.021				0.011				0.002	0.032	0.032	0.001
H_2S	0.011		0.008	0.237													
Cl_2				0.003													
COS				0.016													
Molar flow rates (kmol/s)	2.33	0.54	2.87	0.097	2.77	1.25	1.52	0.87	0.40	2.46	0.732	0.331	3.31	12.98	13.00	0.684	4.95

Figure 2: Block diagram and simulation results of coal IGCC system with CO_2 reuse through tri-reforming and methanol synthesis (Scheme B).