CO-PRODUCTION OF HYDROGEN AND ELECTRICITY BY COAL GASIFICATION WITH CO_2 CAPTURE

Technical Study

Report Number: 2007/13

Date: September 2007

INTERNATIONAL ENERGY AGENCY

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ACKNOWLEDGEMENTS AND CITATIONS

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Comments were also received from reviewers at a company with expertise in IGCC who wished to remain anonymous.

Foster Wheeler asked the suppliers of the three gasifier technologies considered in this report to review the sections concerning their own technologies.

The report should be cited in literature as follows:

IEA Greenhouse Gas R&D Programme (IEA GHG), "Co-Production of Hydrogen and Electricity by Coal Gasification with CO_2 Capture", 2007/13, September 2007.

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Background

Hydrogen may become widely used in future as a low-CO_2 energy carrier for vehicles and distributed heat and power generation using fuel cells. The long term goal for the 'hydrogen economy' is generally recognised to be production of hydrogen from sustainable renewable energy sources but in the near term the cheapest way to produce hydrogen with low CO_2 emissions is expected to be by use of fossil fuels with CO_2 capture and storage.

Hydrogen can be produced from fossil fuels in stand-alone plants with CO_2 capture but it may be advantageous to co-produce hydrogen and electricity. Co-production would provide synergies within the production plant and would also help to cope with the variability in the demands for the two products. Flexible co-production plants could become increasingly attractive in future when electricity grids include large proportions of variable renewable energy generation.

Hydrogen is currently used on a large scale in ammonia plants and modern petroleum refineries. The main fuels used for hydrogen production are currently natural gas and petroleum residues. However, if the use of hydrogen as an energy carrier becomes widespread coal may become the most important fuel for hydrogen production. This study therefore focuses on estimating the costs of producing hydrogen and electricity by coal gasification with CO_2 capture and the advantages of flexible co-production.

The study was undertaken for IEA GHG by Foster Wheeler Italiana.

Study description

Scope of the study

The scope of the study is as follows:

- Screening assessment of the performance and costs of hydrogen and electricity co-production plants with CO_2 capture, based on three coal gasifiers: Shell, GE Energy (formerly Texaco) and Siemens (formerly Future Energy), and two acid gas removal processes (Selexol and Rectisol), leading to the selection of preferred technologies for later detailed assessments.

- Assessment of the performance and costs of the following coal gasification plants:
  1. Production of electricity, without CO_2 capture
  2. Production of electricity, with CO_2 capture
  3. Production of hydrogen and sufficient electricity for internal plant consumption, with CO_2 capture
  4. Co-production of hydrogen and electricity (fixed ratio), with CO_2 capture
  5. Co-production of hydrogen and electricity (flexible ratio), with CO_2 capture

- Assessment of current markets for electricity, natural gas and road vehicle fuels in The Netherlands and USA, including the variability in consumptions throughout the year.

- An outline projection of the potential future market for hydrogen as an energy carrier, assuming it is used to substitute for current actual consumptions of natural gas for small scale energy users and petrol and diesel fuel for road vehicles.

- Review of published information on large scale underground storage of hydrogen.

- Modelling of scenarios in which different types of plants are used to meet demands for hydrogen and electricity:
  1. Electricity-only and hydrogen-only plants without hydrogen storage
  2. Non-flexible co-production plants without hydrogen storage
  3. Non-flexible co-production plants with hydrogen storage
  4. Flexible co-production plants without hydrogen storage
  5. Flexible co-production plants with hydrogen storage

  The scenarios are based on the variations in energy demands in the Netherlands and USA. In the scenarios with co-production plants, electricity-only and hydrogen-only plants are also used where necessary.

Study basis
The study is based on the standard technical and economic criteria used in IEA GHG's other studies on large scale plants with CO_2 capture. The plants are assumed to be located at a coastal site in The Netherlands and the coal feed is an Australian bituminous coal with a sulphur content of 1.1% (dry ash free basis). In the plants with CO_2 capture, approximately 85% of the CO_2 is captured. For consistency with other IEA GHG studies, the production costs are based on a 10% annual discount rate, zero inflation, 25 year plant life and a coal cost of US$1.5/GJ (€1.2/GJ). Sensitivities to the fuel price and discount rate are calculated.

The electricity-only plants are conventional IGCC plants which include two 9FA gas turbines. In the co-production plants, part of the hydrogen-rich fuel gas is fed to a pressure swing adsorption (PSA) plant which separates high purity hydrogen. The rest of the fuel gas is fed to a single 9FA gas turbine. The off-gas from the PSA unit is re-compressed and fed to the gas turbine and/or used for supplementary firing of the gas turbine heat recovery steam generator. In the flexible plants, the ratio of hydrogen to electricity is varied by varying the proportion of the fuel gas which is fed to the PSA unit and the proportion of the PSA off-gas which is re-compressed and fed to the gas turbine. The hydrogen-only plant generates only sufficient electricity for internal consumption. To this end, a single smaller 6FA gas turbine is used.

The scenarios on based on average monthly energy demands in the Netherlands and USA and also average daily demands in the Netherlands. In the scenarios which include co-production plants, electricity-only and hydrogen-only plants are also used where necessary to enable the overall electricity and hydrogen demands to be satisfied.

Results and Discussion

Technology selection
The costs of electricity are very similar for the three gasification processes. The plant based on Shell gasification has the highest thermal efficiency, the lowest production of CO_2 and will therefore have the lowest cost for CO_2 transport and storage (CO_2 transport and storage is outside the scope of this study). The Shell gasification process was selected for the more detailed case studies.

Two acid removal processes were evaluated for separation of CO_2 and sulphur compounds for the GE and Shell gasification processes. The Rectisol plants has lower variable operating costs than the Selexol plants, mainly due to lower overall energy consumptions, but the capital costs are higher. The payback time for the additional capital cost of the Rectisol plants is greater than the 6 year target, so the Selexol process was selected for the detailed plant studies.

Performance and costs of hydrogen and electricity production plants
The performance and costs of hydrogen and electricity production plants are summarised in table 1. The capital costs include miscellaneous owners' costs but exclude interest during construction, although this is taken into account in the calculation of the costs of electricity and hydrogen. The operating load factor is assumed to be 85%.

Table 1 Plant performance and costs

	Without CO_2 capture	With CO_2 capture				
	Electricity	Electricity	Hydrogen	Electricity and hydrogen (fixed ratio)	Electricity and hydrogen (variable, low H_2)	Electricity and hydrogen (variable, high H_2)
Performance						
Coal feed, MW (LHV)	1800.8	1962.5	1962.5	1962.5	1962.5	1962.5
Electricity gross output, MW	891.9	875.0	208.6	518.1	565.0	443.4
Electricity net output, MW	762.3	655.8	0.1	317.1	363.1	236.6
Hydrogen net output, MW	-	-	1110.7	599.0	484.0	734.1
Hydrogen: electricity ratio				1.89	1.33	3.10
Efficiency to electricity, %	42.3	33.4	-	16.2	18.5	12.1
Efficiency to hydrogen, %	-	-	56.6	30.5	24.7	37.4
CO_2 emitted and stored						
CO_2 to storage, g/kWh$_e$		836		1729	1510	2317
CO_2 emitted, g/kWh$_e$	776	147		303	265	406
Costs						
Capital cost, M€	1266	1560	1196	1337	1350	1350
Capital cost, €/kW$_e$	1661	2379		4216	3718	5706
Cost of hydrogen, €/GJ[1]	-	-	9.45	8.8	8.8	8.8
Cost of electricity, €/kWh	0.052	0.072	-	0.071	0.073	0.078
Cost of CO_2 avoided, €/tonne		31.3				

For electricity-only plants, adding CO_2 capture decreases the efficiency by 8.9 percentage points and increases the coal consumption per kWh of electricity by 27%, increases the capital cost per kW by 43% and increases the cost of electricity by 38%. The cost of CO_2 emissions avoided is €31/tonne CO_2. The costs of the electricity-only plants are higher than in IEA GHG's previous study on bituminous coal IGCC plants with CO_2 capture[2] because costs of process plants have recently increased substantially, particularly due to increases in materials costs.

The electricity and hydrogen costs of the fixed-ratio co-production plant are lower than those of the electricity-only and hydrogen-only plants, which demonstrates the synergies of co-production.

The flexible plant can vary the hydrogen: electricity net output ratio between 1.3:1 and 3.1:1, while continuing to operate the coal gasifiers and gas turbine at full load. Including this flexibility slightly increases the capital and operating costs. However, as described later in the scenario analyses, flexibility has the advantage of enabling plants to meet the varying market demands more effectively and at lower costs.

Doubling the coal price to $3/GJ increases the cost of electricity from the electricity-only plant with capture to €0.085 /kWh and increases the cost of CO_2 avoided to €35/t. Doubling the coal price increases the cost of hydrogen from the hydrogen-only plant to €11.57/GJ. In the same cases, decreasing the discount rate to 5% decreases the electricity cost to €0.056/kWh and decreases the hydrogen cost to €7.4/GJ

Electricity and hydrogen demands

Scenarios involving varying demands for hydrogen and electricity energy carriers were assessed to illustrate the benefits of flexible co-production. Monthly consumptions of electricity, natural gas and motor vehicle fuels in The Netherlands and USA for the years 2004 and 2005 were obtained from published sources. A more detailed analysis on an hourly basis for the Netherlands was also obtained. For the purposes of this study it was assumed that hydrogen replaces 60% of the natural gas currently used for residential, commercial and other non-industrial consumers and all of the gasoline and diesel

[1] For the co-production plants an arbitrary assumption has to be made about the split between the revenues associated with hydrogen and electricity. The hydrogen value of €8.8/GJ (LHV) assumed for the co-production plants gives similar electricity costs for the fixed-ratio co-production plant and the electricity-only plant.

[2] IEA GHG report PH4/19, May 2003

fuel used for motor vehicles. The higher efficiencies of fuel cells compared to internal combustion engines in vehicles are taken into account when calculating the hydrogen requirements. The electricity demand to be met from coal-based plants in the scenarios was assumed to be the current demand minus the current production from nuclear and renewable sources.

The extent to which hydrogen and electricity may substitute for natural gas and vehicle fuels in future is highly uncertain. Development of lower cost fuel cells is a critical issue for large scale use of hydrogen, although hydrogen can be used in internal combustion engines if necessary. Improvements in battery technologies may result in electricity being preferred instead of hydrogen for some types of vehicles. Similarly, electricity may also be used to replace some of the natural gas currently used by small stationary consumers, e.g. through greater use of heat pumps. This would decrease the ratio of hydrogen: electricity required from coal-based plants. On the other hand, some of the hydrogen used by small stationary consumers may be used in fuel cells which would co-generate electricity, which would increase the ratio of hydrogen: electricity required from coal-based plants. Detailed prediction of future energy demand scenarios is beyond the scope of this study and no judgement is implied regarding the extent to which hydrogen will be used in future as an energy carrier. The scenarios in this study are intended simply to illustrate possible future benefits of hydrogen and electricity co-production and the sensitivity to different demand profiles. An Excel tool was developed to enable different hydrogen and electricity demand scenarios to be evaluated by others if required and the tool is distributed with this report.

In the Netherlands the peak electricity demand is in the winter, although the peak monthly demand is only about 15% higher than the lowest monthly demand in summer. Similarly there is no large trend in motor fuel demand across the year, the main fluctuation is between adjacent months. There is a much greater variation in natural gas demand, which is about 2.4 times higher in the peak winter month compared to the minimum demand in summer.

In the USA there is a similar although less pronounced peak winter demand for natural gas (about 1.8 times minimum monthly demand). There is a modest peak demand for vehicle fuels in the summer and a more pronounced peak for electricity (up to 40% higher than in winter).

The electricity, natural gas and vehicle fuel demands were used to predict the relative monthly demands for electricity and hydrogen for the scenarios in this report. These are shown in Figure 1.

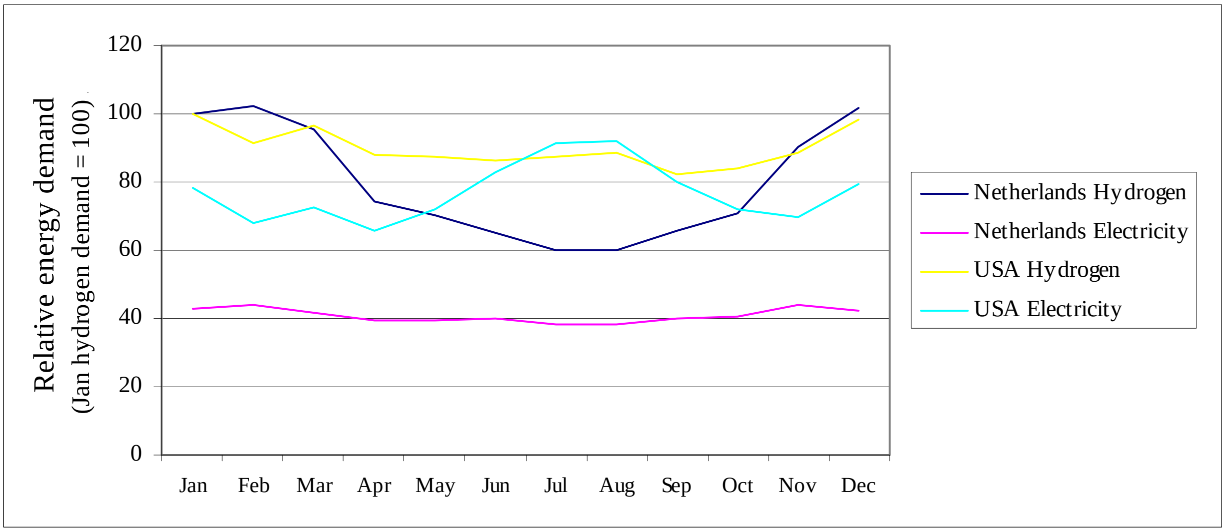

Figure 1 Monthly consumptions of hydrogen and electricity in the Netherlands and USA (2005)

Co-production scenarios

Five scenarios were assessed, in which the different types of plants shown in table 1 were used to satisfy projected electricity and hydrogen demands of the Netherlands and USA. In the scenarios it is assumed that only coal gasification plants are used to meet the electricity and hydrogen demands. Including other technology options could be advantageous but the analysis would have become much more complex. Other researchers could use the plant performance and cost data provided by this study to carry out more complex scenario analyses if required.

Hydrogen storage was included in some of the scenarios to help to smooth out fluctuations in hydrogen demand. Hydrogen can be stored above ground as a refrigerated liquid, in metal hydrides or as a high pressure gas or it can be stored underground in salt caverns, aquifers etc. For storage of large quantities of hydrogen, underground storage has substantially lower costs and has therefore been used in the scenarios in this study. Underground storage of natural gas is widely used and there are some places where hydrogen is commercially stored underground, e.g. in the UK and Texas. Published costs of underground hydrogen storage vary greatly, between 1 and 40 US$/kg of storage capacity. A cost of €1.5/kg was assumed for this study but storage was shown to have advantages up to a storage cost of €35/kg. Potential issues to be assessed on a site specific basis include potential loss of hydrogen by seepage from the store and contamination of the hydrogen product. The additional costs of purification of stored hydrogen were taken into account in this study.

The results of the scenarios for the Netherlands, based on monthly demands, are shown in table 2.

Table 2 Co-production scenarios for the Netherlands

Scenario	1	2	3	4	5
	Electricity and hydrogen-only plants	Non-flexible co-production w/o H_2 storage	Non-flexible co-production with H_2 storage	Flexible co-production w/o H_2 storage	Flexible co-production with H_2 storage
Numbers of plants					
Electricity-only	21	7	4	7	
Hydrogen-only	29	13	5	9	
Non-flexible co-production		29	36		
Flexible co-production				33	41
Total number of plants	50	49	45	49	41
Weighted average % plant utilisation[3]	81	81	87	82	99
Performance and costs relative to scenario 1					
Total capital cost (inc storage)	100	97	90	98	83
Coal consumption and CO_2 emissions	100	98	96	100	100
Electricity cost[4]	100	95	87	97	78

Including hydrogen storage improves the costs for non-flexible and flexible co-production. The lowest costs are for scenario 5, based on flexible co-production and hydrogen storage. The cost of electricity is 22% lower than in scenario 1 (electricity-only and hydrogen-only plants without storage) and the average plant utilisation is 99% compared to 81%.

Costs of electricity for the scenarios in the Netherlands and the USA are compared in figure 2. Costs in the US scenarios are lower than in the Netherlands scenarios. There is less overall variability in demand in the US scenarios and the peak demands for hydrogen and electricity are at different times of year whereas in the Netherlands they are at similar times of year. This results in higher plant utilisation and less need for hydrogen storage in the US scenarios.

[3] The utilisation rate is a percentage of the plant availability, which is assumed to be 85%.
[4] Hydrogen is assumed to have a constant value of €8.8/GJ. The variation in overall operating costs is therefore less than the variation in the electricity cost.

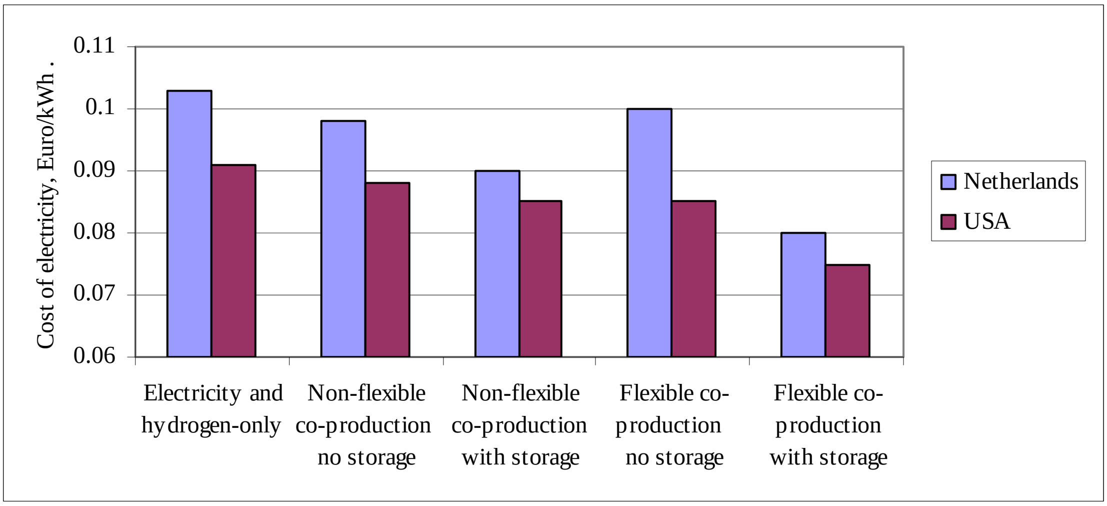

Figure 2   Comparison of electricity costs in Netherlands and US scenarios

Electricity grids in future are expected to include greater amounts of variable renewable energy sources (wind, solar, marine energy etc.). Flexible hydrogen and electricity co-production plants with hydrogen storage may be a relatively low cost way of accommodating large proportions of renewable energy in electricity grids but assessment of this was beyond the scope of this study.

Expert Reviewers' Comments

The draft study report was reviewed by various external experts. IEA GHG is very grateful to those who contributed to this review. The report was generally well received by the reviewers. Most of the comments were concerned with improving the presentation of the large amount of information in the report. A small number of specific technical issues were also raised. Where possible, modifications were made to the report to address the reviewers' comments.

Major Conclusions

The costs of energy production and conversion plants in general have recently increased substantially, due in particular to large increases in materials and equipment prices.

The cost of generating electricity from coal in a base load IGCC plant with CO_2 capture is estimated to be €0.072/kWh and the cost of avoiding CO_2 emissions, compared to an IGCC plant without CO_2 capture, is €31/t CO_2. The cost of producing hydrogen by coal gasification with CO_2 capture in a base load plant is estimated to be €9.4/GJ.

Hydrogen and electricity and be readily co-produced in gasification plants. Simple modifications to the plant design enable the hydrogen: electricity ratio to be varied between 1.3:1 and 3.1:1 on an energy basis, while continuing to operate the coal gasifiers at full load.

The least cost way of meeting hydrogen and electricity demands is to use flexible co-production plants, in combination with underground buffer storage of hydrogen. Assuming a constant hydrogen value, the cost of electricity generation in scenarios based on flexible co-production plants and hydrogen storage is around 20% lower than in scenarios based on electricity-only and hydrogen-only plants without storage.

Recommendations

Costs of abating CO_2 emissions from small stationary sources by CO_2 capture and storage should be compared to the costs of using low-CO_2 energy carriers produced by large plants with CO_2 capture and storage. This study provides costs of producing hydrogen and electricity energy carriers and other recent IEA GHG studies provide information on CO_2 capture from medium scale combustion installations and pipeline collection of CO_2. Information on costs of hydrogen and electricity distribution will need to be obtained to complete the comparison. Use of biomass to reduce net CO_2 emissions could also be included in the comparison. Development needs for each of the options should be assessed.

Researchers who are carrying out detailed modelling of electricity systems and national energy scenarios are recommended to use the plant performance and cost data from this study as input data for their models. A significant issue to be addressed in such modelling would be the possibility that flexible co-production plants with hydrogen storage could reduce the electricity system costs associated with high levels of variable renewable electricity generation.

Co-production of hydrogen and electricity by coal gasification with CO_2 capture

Final Report

September 2007

CO-PRODUCTION OF HYDROGEN AND ELECTRICITY BY COAL GASIFICATION WITH CO_2 CAPTURE

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SECTION J ATTACHEMENTS

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : EXECUTIVE SUMMARY

ISSUED BY : L. VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev. 1	L. Valota	P. Cotone	S. Arienti

SECTION A

EXECUTIVE SUMMARY

I N D E X

1.0 Purpose of the Study

IEA Greenhouse Gas R&D Programme (IEA GHG) retained Foster Wheeler to investigate alternative power and hydrogen generation plant designs, based on high rank coal gasification, in order to assess the potential advantage of flexible co-production of hydrogen and electricity with capture of CO_2.
The primary purpose of this study is, therefore, the evaluation of the technologies and the process alternatives that can be used in these complex power and hydrogen generation schemes to optimise efficiency and capital cost and reduce, at the same time, emissions to the atmosphere.
Use of hydrogen storage is considered to match the hydrogen demand. Different storage options are analysed in Attachment C to the study report.

The study is based on the hydrogen and electricity demands of The Netherlands and the USA, in a future scenario with the standard fossil fuel systems replaced as much as possible by hydrogen systems. The Netherlands and the USA represent, on a regional scale, two significantly different consumption scenarios. The future demands are evaluated in Attachment A to this report.

The plant of the study has a nominal capacity of 750 MWe and is fed with a typical bituminous coal having a low heating value (LHV) equal to 25870 kJ/kg and a sulphur content equal to 1.1 % wt.
The study is based on the current state-of-the-art technologies, evaluating costs and performances of plants which can be presently engineered and built.
The study reviews and compares three available gasification technologies and two available solvents for acid gas removal from the syngas.
After the selection of the technologies (Shell gasification and Selexol solvent), the study develops five possible production plant schemes, described in paragraph 5.1 of this Executive Summary.

Finally five co-production scenarios, obtained as combinations of different types of production plants and of different hydrogen storage volumes (see paragraph 5.2 of this Executive Summary) are evaluated and compared to find the most promising combination of plants and storages.
A software model has been prepared to provide automatically the data relevant to each scenario on the basis of different energy consumption values.

For the preparation of the study FWI based part of the work on the two following studies performed by FWI for IEA GHG:

- Gasification Power Generation Study – March 2003;
- CO_2 Capture in Low-Rank Coal Power Plants – November 2005.

These previous studies were supported by several companies (Dow, General Electric, Shell, Synetix, Sud-Chemie, Texaco, UOP, Future Energy, Siemens and Johnson Matthey Catalysts).

For the present study FW would like to acknowledge the following companies for their fruitful support:

- General Electric, Shell and Siemens for the review of the sections concerning gasification;
- Linde for the data provided on the Rectisol solvent;
- UOP for the data provided on the hydrogen production system.

2.0 Bases of Design

2.1 Process design basis

The IGCC Complex is designed to process, in an environmentally acceptable manner, a coal from eastern Australia (see Section B, paragraph 2.1) and produce electric energy and hydrogen.

The Gasification Island design capacity is determined to produce the syngas that matches the appetite of two gas turbines GE 9 FA. In the different alternative IGCC schemes considered in this study, one or two gas turbines have been selected, depending on the configuration analysed.

The Power Island inside the IGCC Complex is also able to process Natural Gas as back-up fuel for start-up and emergency situations; use of back-up fuel was not taken into account in the economic assessment.

The IGCC Complex main products are electric energy and hydrogen.
By-products are:

- Sulphur (liquid or solid);
- Carbon Dioxide for the alternatives recovering CO_2;
- Solid by-products: slag, fly ash and filter cake, depending on the gasification technology.

The overall gaseous emissions from the IGCC Complex referred to dry flue gas with 15% volume O_2 shall not exceed the following limits:

NO_x (as NO_2)	:	≤ 80 mg/Nm^3
SOx (as SO_2)	:	≤ 10 mg/Nm^3
Particulate	:	≤ 10 mg/Nm^3
CO	:	≤ 50 mg/Nm^3

Characteristics of wastewater discharged from power plants shall comply with the limits stated by the current EU directives.

The bases of design of the IGCC Complex, such as capacity, required availability, location, climatic data etc, are defined in Section B of the Study.

2.2 Consumption of hydrogen and electric power

As part of the present study (reported in attachment A) a section aimed at determining the hydrogen and electricity demand in The Netherlands and in USA is presented.

The first part of this section is a collection and description of the energy consumption data such as electricity, natural gas, gasoline and diesel oil, of the two above-mentioned regions. These regions have been chosen because they represent, at a regional scale, two possible different world consumption scenarios; indeed The Netherlands presents a peak winter demand for electricity mostly due to electrical heaters while in the United States the electricity peak is during summertime for the massive use of electrical air conditioner.

Table A.2.1 and A.2.2 and Figure A.2.1 and A.2.2 show the trend of the actual energies consumption for The Netherlands and USA for 2004-2005.

FOSTER WHEELER

MONTHLY ENERGY CONSUMPTION - TABLE A.2.1

Date: July 2007
Rev: Rev. 1
Made by: FWI

NL	EE consumption	EE Without Nucl/Ren	EE Without Nucl/Ren	NG consumption	NG Without Power Gen & Ind.	Actualized NG	Gasoline	Actualized Gasoline	Gasoline FC	Diesel	Actualized Diesel	Diesel FC	H2	H2	H2/EE
	TJ	*TJ*	*GWh*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*GWh*	
jan-04	28,237	26,232	7,287	212,472	89,238	53,543	13,833	13,833	4,940	20,095	20,095	11,483	69,966	19,435	2.667
feb-04	27,700	25,733	7,148	199,850	83,937	50,362	14,353	14,353	5,126	20,644	20,644	11,797	67,285	18,690	2.615
mar-04	28,920	26,867	7,463	177,005	74,342	44,605	15,994	15,994	5,712	24,042	24,042	13,738	64,055	17,793	2.384
apr-04	28,377	26,362	7,323	122,671	51,522	30,913	16,877	16,877	6,027	22,398	22,398	12,799	49,739	13,816	1.887
may-04	27,404	25,458	7,072	114,782	48,208	28,925	14,511	14,511	5,182	21,171	21,171	12,098	46,205	12,835	1.815
june-04	28,474	26,453	7,348	97,690	41,030	24,618	15,773	15,773	5,633	22,919	22,919	13,097	43,347	12,041	1.639
july-04	27,810	25,835	7,176	87,171	36,612	21,967	14,353	14,353	5,126	21,350	21,350	12,200	39,294	10,915	1.521
ago-04	28,060	26,068	7,241	85,725	36,004	21,603	14,038	14,038	5,013	20,095	20,095	11,483	38,099	10,583	1.462
sept-04	29,070	27,006	7,502	102,817	43,183	25,910	15,457	15,457	5,520	23,093	23,093	13,196	44,626	12,396	1.652
oct-04	29,188	27,116	7,532	131,973	55,429	33,257	14,748	14,748	5,267	23,145	23,145	13,225	51,750	14,375	1.908
nov-04	31,099	28,891	8,025	163,956	68,861	41,317	15,931	15,931	5,689	23,266	23,266	13,295	60,301	16,750	2.087
dec-04	31,796	29,538	8,205	200,244	84,102	50,461	15,899	15,899	5,678	23,324	23,324	13,328	69,468	19,297	2.352
jan-05	30,676	28,498	7,916	197,220	82,832	49,699	14,117	14,117	5,042	20,274	20,274	11,585	66,326	18,424	2.327
feb-05	31,290	29,068	8,074	202,085	84,876	50,925	14,890	14,890	5,318	20,581	20,581	11,760	68,003	18,890	2.339
mar-05	29,767	27,653	7,681	174,375	73,238	43,943	15,552	15,552	5,554	23,862	23,862	13,636	63,132	17,537	2.283
apr-05	28,148	26,149	7,264	121,652	51,094	30,656	15,410	15,410	5,504	22,919	22,919	13,097	49,256	13,682	1.884
may-05	28,086	26,092	7,248	111,429	46,800	28,080	15,047	15,047	5,374	22,786	22,786	13,020	46,475	12,910	1.781
june-05	28,444	26,424	7,340	96,046	40,339	24,204	15,284	15,284	5,459	23,961	23,961	13,692	43,354	12,043	1.641
july-05	27,528	25,573	7,104	90,031	37,813	22,688	13,801	13,801	4,929	20,992	20,992	11,995	39,612	11,003	1.549
ago-05	27,360	25,417	7,060	90,064	37,827	22,696	14,511	14,511	5,182	20,812	20,812	11,893	39,771	11,048	1.565
sept-05	28,732	26,692	7,414	97,492	40,947	24,568	15,221	15,221	5,436	23,787	23,787	13,593	43,597	12,110	1.633
oct-05	29,106	27,040	7,511	113,993	47,877	28,726	14,416	14,416	5,149	22,786	22,786	13,020	46,895	13,027	1.734
nov-05	31,333	29,108	8,086	160,471	67,398	40,439	15,631	15,631	5,582	24,134	24,134	13,791	59,812	16,615	2.055
dec-05	30,353	28,198	7,833	192,191	80,720	48,432	15,615	15,615	5,577	23,503	23,503	13,431	67,439	18,733	2.392

Nuclear and Renewable Energy % of Total Electric Power Production	7.1%
Power Generation and Industrial Natural Gas % of Total consumed Gas	58.0%
Natural Gas actualization factor	0.60
Gasoline actualization factor	1.00
Diesel actualization factor	1.00
Gasoline Motor Efficiency	25.0%
Diesel Motor Efficiency	40.0%
Fuel Cell Efficiency	70.0%

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MONTHLY ENERGY CONSUMPTION - TABLE A.2.2

Date: July 2007
Rev: Rev. 1
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USA	EE consumption	EE Without Nucl/Ren	EE Without Nucl/Ren	NG consumption	NG Without Power Gen & Ind.	Actualized NG	Gasoline	Actualized Gasoline	Gasoline FC	Diesel	Actualized Diesel	Diesel FC	H2	H2	H2/EE
	TJ	*TJ*	*GWh*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*GWh*	
jan-04	1,247,566	1,089,125	302,535	2,604,930	1,015,923	609,554	1,373,022	1,373,022	490,365	484,422	484,422	276,813	1,376,731	382,425	1.264
feb-04	1,131,408	987,719	274,366	2,420,824	944,121	566,473	1,303,993	1,303,993	465,712	484,422	484,422	276,813	1,308,997	363,610	1.325
mar-04	1,111,723	970,534	269,593	2,124,855	828,693	497,216	1,423,343	1,423,343	508,337	522,620	522,620	298,640	1,304,193	362,276	1.344
apr-04	1,046,016	913,172	253,659	1,672,841	652,408	391,445	1,392,977	1,392,977	497,492	531,302	531,302	303,601	1,192,538	331,260	1.306
may-04	1,178,568	1,028,890	285,803	1,463,684	570,837	342,502	1,447,851	1,447,851	517,090	520,884	520,884	297,648	1,157,240	321,456	1.125
june-04	1,242,306	1,084,533	301,259	1,307,188	509,803	305,882	1,422,880	1,422,880	508,171	550,401	550,401	314,515	1,128,568	313,491	1.041
july-04	1,358,395	1,185,879	329,411	1,500,118	585,046	351,028	1,475,932	1,475,932	527,119	526,093	526,093	300,624	1,178,771	327,436	0.994
ago-04	1,326,380	1,157,930	321,647	1,515,791	591,158	354,695	1,471,068	1,471,068	525,381	531,302	531,302	303,601	1,183,677	328,799	1.022
sept-04	1,208,239	1,054,793	292,998	1,397,736	545,117	327,070	1,376,121	1,376,121	491,472	555,610	555,610	317,491	1,136,033	315,565	1.077
oct-04	1,124,820	981,968	272,769	1,477,558	576,248	345,749	1,434,826	1,434,826	512,438	559,082	559,082	319,476	1,177,662	327,128	1.199
nov-04	1,087,564	949,443	263,734	1,714,534	668,668	401,201	1,382,176	1,382,176	493,634	524,357	524,357	299,632	1,194,467	331,797	1.258
dec-04	1,231,013	1,074,674	298,521	2,240,004	873,602	524,161	1,451,973	1,451,973	518,562	515,675	515,675	294,672	1,337,394	371,498	1.244
jan-05	1,235,624	1,078,700	299,639	2,526,700	985,413	591,248	1,389,991	1,389,991	496,425	501,785	501,785	286,734	1,374,407	381,780	1.274
feb-05	1,072,584	936,366	260,102	2,218,690	865,289	519,173	1,262,435	1,262,435	450,869	494,840	494,840	282,766	1,252,809	348,002	1.338
mar-05	1,140,408	995,576	276,549	2,175,786	848,557	509,134	1,418,539	1,418,539	506,621	539,983	539,983	308,562	1,324,317	367,866	1.330
apr-05	1,038,838	906,906	251,918	1,716,022	669,249	401,549	1,393,232	1,393,232	497,583	543,456	543,456	310,546	1,209,678	336,022	1.334
may-05	1,129,583	986,126	273,924	1,592,050	620,900	372,540	1,463,451	1,463,451	522,661	534,774	534,774	305,585	1,200,786	333,552	1.218
june-05	1,301,299	1,136,034	315,565	1,483,648	578,623	347,174	1,430,634	1,430,634	510,941	567,764	567,764	324,436	1,182,551	328,486	1.041
july-05	1,437,307	1,254,769	348,547	1,555,305	606,569	363,941	1,503,748	1,503,748	537,053	529,565	529,565	302,609	1,203,603	334,334	0.959
ago-05	1,447,121	1,263,337	350,927	1,587,067	618,956	371,374	1,504,272	1,504,272	537,240	541,719	541,719	309,554	1,218,168	338,380	0.964
sept-05	1,255,723	1,096,246	304,513	1,399,674	545,873	327,524	1,360,806	1,360,806	486,002	555,610	555,610	317,491	1,131,017	314,171	1.032
oct-05	1,134,122	990,089	275,025	1,450,397	565,655	339,393	1,425,256	1,425,256	509,020	538,247	538,247	307,570	1,155,982	321,106	1.168
nov-05	1,097,636	958,236	266,177	1,746,539	681,150	408,690	1,391,324	1,391,324	496,901	543,456	543,456	310,546	1,216,138	337,816	1.269
dec-05	1,246,514	1,088,207	302,280	2,274,362	887,001	532,201	1,466,163	1,466,163	523,630	522,620	522,620	298,640	1,354,471	376,242	1.245

Nuclear and Renewable Energy % of Total Electric Power Production	12.7%
Power Generation and Industrial Natural Gas % of Total consumed Gas	61.0%
Natural Gas actualization factor	0.60
Gasoline actualization factor	1.00
Diesel actualization factor	1.00
Gasoline Motor Efficiency	25.0%
Diesel Motor Efficiency	40.0%
Fuel Cell Efficiency	70.0%

Figure A.2.1: Actual Energy consumption for the Netherlands (2005-2005)

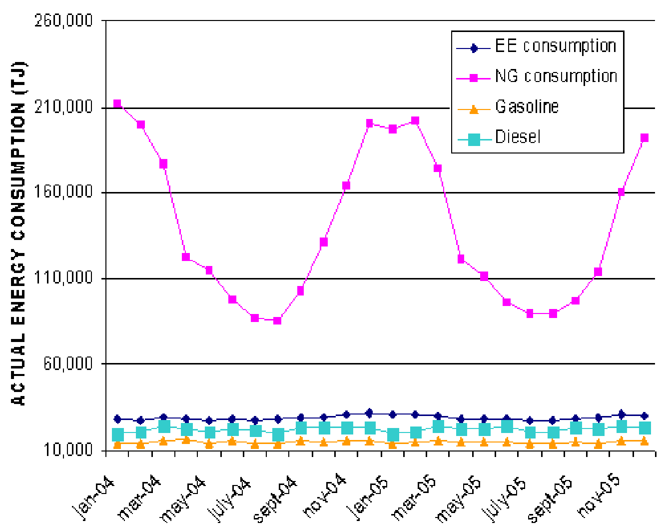
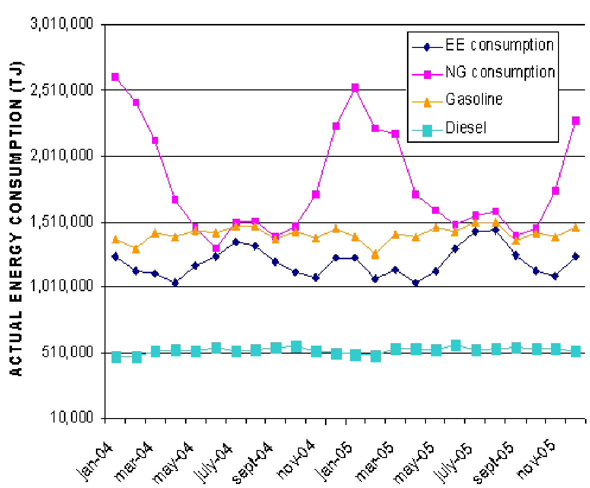


Figure A.2.2: Actual Energy consumption for the USA (2005-2005)



The second part of this section performs an estimate of the required quantity of hydrogen and electricity needed in such areas with the standard fossil fuel systems replaced as much as possible by hydrogen systems. Several criteria are followed for the conversion:

- The production of electricity coming from renewable energy sources and nuclear is not converted in electricity consumption. This because in a hypothetical hydrogen energy scenario, nuclear and renewable may still be used for power production.
- The natural gas consumed by industry and power generation plants is not converted in to H_2 consumption. That is, natural gas will continue to be consumed by power plants.
- 60% of the remaining part of the natural gas consumption is converted to hydrogen. The remaining 40% is kept as gas consumption.
- The diesel and gasoline consumption is converted into hydrogen consumption considering the state-of-the-art fuel cell efficiency.

The final outputs are the absolute demand values of hydrogen and electricity in The Netherlands and USA, derived from 2004-2005 energy consumptions (Tables A.2.3 and A.2.4). Refer to Section I for detailed description of the conversion criterion.

Figure A.2.3: Electricity and hydrogen consumption for the Netherlands (2004-2005)

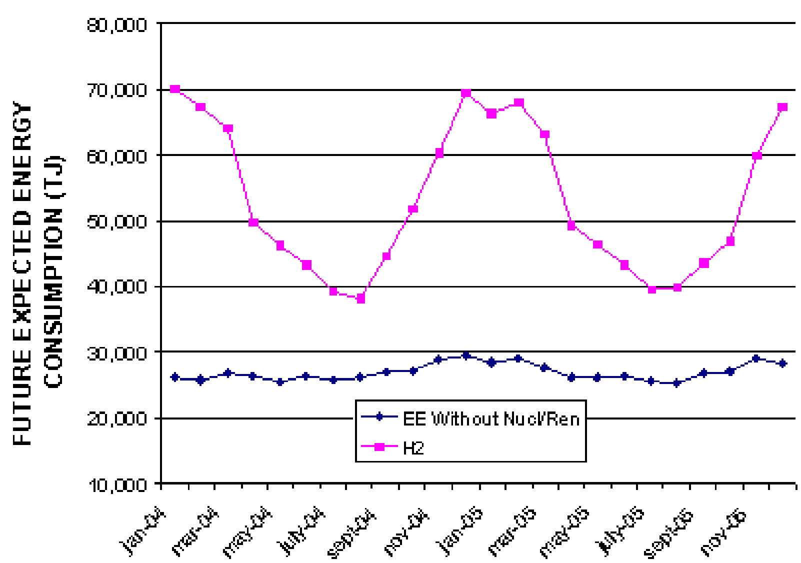
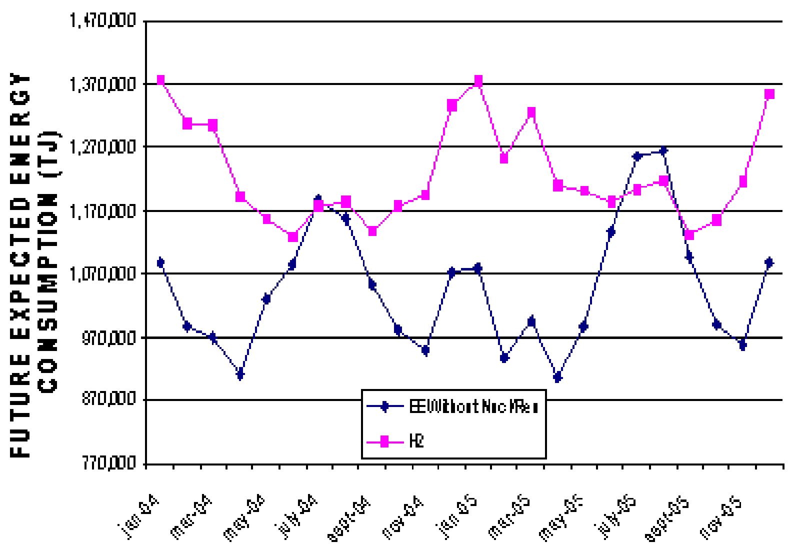


Figure A.2.4: Electricity and hydrogen consumption for the USA (2004-2005)



3.0 Storage of hydrogen

In order to constantly match the hydrogen demand with different plant configurations and keep the number of plants as low as possible, hydrogen storage is considered.

In attachment C, different storage technologies are described, focusing on advantages and disadvantages for the requirement to store large amounts of hydrogen. Estimation of the costs is also provided for different storage options. Finally relevant data from state-of-the-art hydrogen storage experiences are provided and an explanation of the criteria of choice is presented.

The main options for storing hydrogen are as a compressed gas (above ground or underground), as a liquid or in metal hydrides.

The following general conclusions can be made:

- The metal hydride option is not suitable to large quantities of hydrogen;
- Underground storage is convenient for large quantities of gas and long-term storage;
- Aboveground compressed gas storage is suitable only for small quantities of gas and short periods due to its very high costs;
- Liquid hydrogen has specific applications related to high storage energy density but requires very expensive cryogenic facilities.

For the scope of this study underground storage is the best solution in relation to the very large volumes of hydrogen to be stored for long periods.

4.0 Comparison of technologies

Three technologies for gasification and two technologies for Acid Gas Removal are investigated and compared. The comparison is made considering the effect of each technology on the co-production plant described in para 5.0 of this executive summary as case 4 (CO_2 capture and fixed H_2/electricity production ratio).

4.1 Gasification technologies

Three technologies are evaluated:

- General Electric Energy (GEE)
- Shell
- Siemens

The most important performance and economic data of the co-production plants based on the three gasification technologies are summarized in the following Table A.4.1.

Table A.4.1 – Performance data.

		GEE Gasifier	Shell Gasifier	Siemens Gasifier
ACID GAS REMOVAL TECHNOLOGY		Selexol	Selexol	Selexol
CO_2 Capture Efficiency	**%**	84.8	85.1	84.9
OVERALL PERFORMANCES				
Coal Flow Rate A.R.	**t/h**	323.1	273.1	295.3
Coal LHV	**kJ/kg**	25,869.5	25,869.5	25869.5
Thermal Energy of Feedstock	**MWth**	2321.8	1962.5	2122.0
Actual Gross Electric power output	**MWe**	625.1	518.1	538.5
H_2 produced	**MWth**	598	599	591.8
Auxiliary Consumption	**MWe**	234.3	201	211.7
Actual Net Electric power output	**MWe**	390.8	317.1	326.8
Net Equivalent Electric Power Output	**MWe**	725.7	652.5	658.2
Gross Equivalent Electrical Efficiency	**%**	41.3	43.5	41.0
Hydrogen Equivalent electric power	**MWe**	334.9	335.4	331.4
Gross Equivalent Electric Power Output	**MWe**	960	853.5	869.9
Net Equivalent Electrical Efficiency	**%**	31.3	33.3	31.0
(H_2/effective EE) ratio	**MWth/MWe**	1.5	1.9	1.8
INVESTMENT COST DATA				
Total Investment	**10^6 €**	1476.8	1336.9	1312.2
Equivalent Specific Net Investment Cost	**€/kW**	2035	2049	1994
O&M Costs	**MM€**	136.2	116.5	123.4
PRODUCTION COST DATA				
C.O.E (DCF=10%)	**c€/kWh**	0.071	0.071	0.071

Shell gasification allows the best efficiency of the plant.
These parameters contribute to the evaluation of the cost of electricity (COE), the figure taken to compare the three alternatives, at fixed H_2 selling price (9.5 €/Nm^3) and 10% discount rate.
The calculated COE for Shell, GEE and Siemens are the same (0.071€/kWh).

For the prosecution of the study Shell technology is used for four reasons:

- Shell technology appears (like Siemens) the most suitable to match the H_2/electricity ratio of Netherlands, taken as a reference parameter for the fixed co-production plant (see paragraph 5.1 of this executive summary).
- Shell gasification technologies (like GEE) have more operating plants than Siemens.
- Shell gasification presents higher efficiency (and as consequence lower CO_2 production and lower CO_2 storage costs).
- Better accuracy of Shell investment cost, which is based on the most recent data of the year 2005 study, while GEE figures are taken from the year 2003 study and Siemens costs have been derived by FWI from on data provided by Siemens in previous studies.

4.2 Acid Gas Removal solvent

Two Acid Gas Removal solvents are evaluated:

- Option 1 - Selexol
- Option 2 - Rectisol

For both solvents, the comparison has been performed on the following gasification technologies:

- GEE HP gasification with separate H_2S and CO_2 capture;
- Shell LP gasification with separate H_2S and CO_2 capture;

For both gasification technologies, the CAPEX comparison is in favour of Option 1 – Selexol (saving respectively 58.0 MM€ and 92.8 MM€ in the GEE and Shell cases).

For both gasification technologies, the OPEX comparison is in favour of Option 2 – Rectisol (saving respectively 9.5 MM€/y and 3.6 MM€/y in the GEE and Shell cases).

From the comparison of OPEX and CAPEX, the pay back time for Rectisol in the GEE case is approx 6 years, while for the Shell case it is more than 20 years. The Selexol based AGR is preferred both for GEE and for Shell gasification technology.

5.0 Co-production alternatives

5.1 Plant alternatives review

The following five design alternatives of the IGCC complex are developed in the Study:

- Case 1: production of electric energy only, without CO_2 capture; this is taken as a reference case
- Case 2: production of electric energy only, with CO_2 capture
- Case 3: co-production of the maximum quantity of hydrogen and of the minimum electric energy to satisfy the internal electrical consumption, with CO_2 capture
- Case 4: co-production of hydrogen and electric energy at a fixed specific ratio and with CO_2 capture; the ratio corresponds to the future H_2/electric energy consumptions evaluated for the Netherlands.
- Case 5: co-production of hydrogen and electric energy at flexible ratios and with CO_2 capture.

The following table 5.1 summarizes the performances, O&M costs and investment costs of the five cases.
For the case 5 the performances are given at the minimum and at the maximum required H_2/electric energy ratio for the Netherlands.
The data contained in this table are used for the evaluation of the different co-production scenarios presented in para. 5.2 of this Executive Summary.

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OVERALL ECONOMICS
PERFORMANCE and COST SUMMARY - TABLE A.5.1

Date: July 2007
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			Case #1 plant	Case #2 plant	Case #3 plant	Case #4 plant	Case #5 plant-R low	Case #5 plant-R high
			w/o CO_2 capture, w/o H_2 production	CO_2 capture; No H_2 production	CO_2 capture; maximum H_2 production	CO_2 capture; H_2 production; optimum fixed H_2/EE ratio;	CO_2 capture; H_2 production; flexible H_2/EE ratio; R low	CO_2 capture; H_2 production; flexible H_2/EE ratio; R high
Gasification	Coal consumption	t/h	**250.6**	**273.1**	**273.1**	**273.1**	**273.1**	**273.1**
PSA	Hydrogen production (99.5% purity)	Nm^3/h	**n/a**	**n/a**	**372,400.0**	**200,858.0**	**162,240.0**	**246,160.0**
	Hydrogen Thermal Power (E)	MWt	**n/a**	**n/a**	**1,110.7**	**599.0**	**484.0**	**734.1**
Consumption	Electric power consumption of IGCC complex	MWe	**129.6**	**219.2**	**208.5**	**201**	**201.9**	**206.8**
Power Island	Gas turbines total power output	MWe	**553.6**	**572**	**87.6**	**286**	**286**	**286**
	Steam turbine power output	MWe	**338.3**	**303**	**121**	**232.1**	**279**	**157.4**
	Actual gross electric power output	MWe	**891.9**	**875**	**208.6**	**518.1**	**565**	**443.4**
	Net electric power output (B)	MWe	**762.3**	**655.8**	**0.10**	**317.1**	**363.1**	**236.6**
CO2 capture	Net Carbon flowing to process unit	kmol/h	**n/a**	**14640**	**14640**	**14640**	**14640**	**14640**
	CO_2 to Storage	kmol/h	**n/a**	**12458**	**12458**	**12458**	**12458**	**12458**
	CO_2 Emissions	kmol/h	**n/a**	**2183**	**2183**	**2183**	**2183**	**2183**
Sold Sulphur	Sulphur	t/h	**2.15**	**2.35**	**2.35**	**2.35**	**2.35**	**2.35**
Emissions	NOx	kg/h	**453.6**	**371.2**	**83.6**	**233.6**	**245**	**184.3**
	SOx	kg/h	**28.3**	**5**	**5**	**5**	**5**	**5**
	CO	kg/h	**176**	**155.5**	**36**	**99**	**104**	**78**
	Particulate	kg/h	**28**	**25.1**	**6.3**	**16**	**16**	**10.1**
Cost	Capital cost	EUR	**1,266,055,000**	**1,560,120,000**	**1,196,050,000**	**1,336,860,000**	**1,350,140,000**	**1,350,140,000**
	O&M fixed cost	EUR/y	**39,560,000**	**54,930,000**	**40,670,000**	**46,290,000**	**46,780,000**	**46,780,000**
	O&M variable cost	EUR/y	**62,455,000**	**70,270,000**	**70,250,000**	**70,260,000**	**70,270,000**	**70,270,000**
Avaibility	Availability Factor		**0.85**	**0.85**	**0.85**	**0.85**	**0.85**	**0.85**

For electricity only plants, adding CO_2 capture decreases the efficiency (more coal consumed, less net electric power produced) and consequently increases the capital cost per kWh produced and the cost of electricity.

Considering the co-production plants, the flexible plant (case 5) allows a wide variation of hydrogen/electricity production with a modest increase of the capital cost (1%) with respect to the fixed ratio plant (case 4). The maximum possible hydrogen production in the plants with CO_2 capture (case 3) is 372,400 Nm^3/h vs. 200,800 Nm^3/h of the fixed ratio co-production plant (case 4).

5.2 Co-production scenarios comparison

To satisfy the future H_2 and electric energy needs of the Netherlands, five scenarios are evaluated, consisting of combinations of different types of IGCC plants and H_2 storage.

The five scenarios are:

- Scenario 1: electricity-only and H_2-only production plants, without H_2 storage
- Scenario 2: non-flexible co-production plants, without H_2 storage
- Scenario 3: non-flexible co-production plants, with H_2 storage
- Scenario 4: flexible co-production plants, without H_2 storage
- Scenario 5: flexible co-production plants, with H_2 storage

Refer to section I for detailed description of the criteria behind each scenario.

The economics and overall performances of the five scenarios are summarized in the following table 5.2 for The Netherlands and table 5.3 for the USA, showing also the number of plants and the hydrogen storage volumes necessary in each scenario, as well as the gaseous emission quantities.
The scenarios are compared on the basis of the electricity production cost, at fixed hydrogen price (9.5 € cent/Nm^3), and considering an underground hydrogen storage capital cost of 1.5 €/kg.
The most attractive scenario is Scenario 5, consisting of 41 flexible co-production plants and 6,822 million Nm^3 of hydrogen storage.

The electricity production cost of Scenario 5 is 0.080 €/kWh. This figure is based on the monthly energy consumption of the Netherlands for years 2004-2005 (see table A.2.1 and A.2.2).
A more detailed analysis based on hourly consumptions for the same years shows that the electricity production cost for the same scenario is even lower (0.075 €/kWh). In fact, through the hourly consumption it is possible to take into account the contribution of the hydrogen storage to satisfy the hourly demand variation day by day. This allows running the plants at an optimised hydrogen to electricity ratio.

The advantage of Scenario 5 is confirmed also considering the energy consumption of the USA.

In both cases, the Netherlands and USA, the economics of flexible co-production and fixed-ratio co-production are similar in the case hydrogen storage is not used (see Scenarios 2 and 4).

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OVERALL ECONOMICS AND ADVANTAGES OF COPRODUCTION - NL - TABLE A.5.2

Date: July 2007
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	SCENARIO 1	SCENARIO 2	SCENARIO 3	SCENARIO 4	SCENARIO 5
	EE PLANT AND H_2 PLANT ONLY	NON FLEX COPROD PLANT W/O H_2 STORAGE	NON FLEX COPROD PLANT WITH H_2 STORAGE	FLEXIBLE COPROD PLANT W/O STORAGE	FLEXIBLE COPROD PLANT WITH STORAGE - monthly
Quantity Plants #1	0	0	0	0	0
Quantity Plants #2	21	7	4	7	0
Quantity Plants #3	29	13	5	9	0
Quantity Plants #4	0	29	36	0	0
Quantity Plants #5	0	0	0	33	41
Total quantity of plant	50	49	45	49	41
Monthly average installed plants #1 load factor					
Monthly average installed plants #2 load factor	89.1%	66.5%	35.1%	45.9%	
Monthly average installed plants #3 load factor	75.0%	47.1%	32.5%	45.6%	
Monthly average installed plants #4 load factor		100.0%	100.0%		
Monthly average installed plants #5 load factor				100.0%	99.1%
Max quantity hydrogen in storage (million Nm3)	n/a	n/a	2,389	n/a	6,822
Max quantity hydrogen in storage per plant with storage (million Nm3)	n/a	n/a	66	n/a	166
Overall coal consumption (t/h)	9392	9234	9060	9358	9432
CO_2 capture (kg/h)	18,855,935	18,537,750	18,189,524	18,787,583	18,935,567
CO_2 emission (kg/h)	3,304,102	3,248,347	3,187,328	3,292,125	3,318,056
Plants Capital Cost (excluding storage) (milions EUR)	67,448	65,238	60,348	66,240	55,356
Underground Storage Capital Cost (including extra PSA unit) (milions EUR)	n/a	n/a	390	n/a	962
Total Capital Cost (underground)(milions EUR)	67,448	65,238	60,738	66,240	56,318
Total O&M Cost million EUR/y (underground) (base on monthly average)	5,176	5,050	4,843	5,127	4,786
Electricity Prod Cost [Euro/kWh]	0.103	0.098	0.090	0.100	0.080
NOx EMISSION (kg/h) (including availability, month average	7,447	2,481	1,274	7,591	7,741
SOx EMISSION (kg/h) (including availability, month average)	172	169	166	171	173
CO EMISSION (kg/h) (including availability, month average)	3,138	3,243	3,265	3,216	3,283
PART EMISSION (kg/h) (including availability, month average)	516	563	528	482	483

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OVERALL ECONOMICS AND ADVANTAGES OF COPRODUCTION - USA- TABLE A.5.3

Date: July 2007
Rev: Rev. 1
Made by: FWI

	SCENARIO 1	SCENARIO 2	SCENARIO 3	SCENARIO 4	SCENARIO 5
	EE PLANT AND H_2 PLANT ONLY	NON FLEX COPROD PLANT W/O H_2 STORAGE	NON FLEX COPROD PLANT WITH H_2 STORAGE	FLEXIBLE COPROD PLANT W/O STORAGE	FLEXIBLE COPROD PLANT WITH STORAGE monthly
Quantity Plants #1	0	0	0	0	0
Quantity Plants #2	875	461	425	170	0
Quantity Plants #3	563	101	0	97	0
Quantity Plants #4	0	856	932	0	0
Quantity Plants #5	0	0	0	1132	1253
Total quantity of plant	1438	1418	1357	1399	1253
Monthly average installed plants #1 load factor					
Monthly average installed plants #2 load factor	83.0%	67.7%	65.3%	40.0%	
Monthly average installed plants #3 load factor	89.2%	40.3%		46.7%	
Monthly average installed plants #4 load factor		100.0%	100.0%		
Monthly average installed plants #5 load factor				100.0%	99.99%
Max quantity hydrogen in storage (million Nm3)	n/a	n/a	37,830	n/a	85,016
Max quantity hydrogen in storage per plant with storage (million Nm3)	n/a	n/a	41	n/a	68
Overall coal consumption (t/h)	285091	280559	280730	289074	290832
CO_2 capture (kg/h)	572,350,688	563,251,509	563,593,831	580,345,618	583,874,557
CO_2 emission (kg/h)	100,292,306	98,697,868	98,757,853	101,693,248	102,311,620
Plants Capital Cost (excluding storage) (milions EUR)	2,038,481	1,984,369	1,909,005	1,909,596	1,691,725
Underground Storage Capital Cost (including extra PSA unit) (milions EUR)	n/a	n/a	5,717	n/a	13,718
Total Capital Cost (underground)(milions EUR)	2,038,481	1,984,369	1,914,721	1,909,596	1,705,444
Total O&M Cost million EUR/y (underground) (base on monthly average)	157,251	153,974	151,641	153,743	147,089
Electricity Prod Cost [Euro/kWh]	0.091	0.088	0.085	0.085	0.075
NOx EMISSION (kg/h) (including availability, month average	264,707	118,335	106,043	265,393	266,275
SOx EMISSION (kg/h) (including availability, month average)	5,220	5,137	5,140	5,292	5,325
CO EMISSION (kg/h) (including availability, month average)	111,307	114,506	115,085	112,576	113,052
PART EMISSION (kg/h) (including availability, month average)	18,175	18,837	18,592	17,573	17,561

6.0 Conclusions

The primary scope of the study is the evaluation of plant scenarios to satisfy the future demands of hydrogen and electricity for the Netherlands and for the USA, based on the monthly consumptions of years 2004 and 2005.
The scenarios are compared on the basis of the electricity production cost, at fixed hydrogen price (9.5 €cent/Nm^3) and considering the underground hydrogen storage capital cost of 1.5 €/kg.

The most important conclusions of the study are:

- The preferred scenario is by far Scenario 5, including flexible co-production plants with gaseous hydrogen underground storage. In this scenario for the Netherlands the electricity production cost is 0.080 €/kWh vs. 0.090 €/kWh of the scenario including non-flexible co-production plants and hydrogen storages, and even higher costs for the other scenarios without storage. The same conclusion applies also to the USA case.
- Making reference to more detailed data of energy consumption on an hourly basis, the number of required co-production plants decreases and the electricity production cost for the Netherlands in Scenario 5 becomes 0.075 €/kWh.
- The capital cost of gaseous underground storage varies widely between 1 €/kg and 40 €/kg, depending on the geological configuration of the area, based on available studies on the subject. The comparison among different plant scenarios depends on this cost. For this reason a sensitivity analysis has been performed evaluating the electricity production cost and the underground storage capital cost for each scenario (see graph I.7.1 of this report). Scenario 5 (flexible co-production plants + hydrogen storage) remains the winning choice for a hydrogen storage cost lower than approximately 20 €/kg; for higher costs the impact of the storage on investment cost becomes too high and both alternatives with hydrogen storage appear uncompetitive.
- One concern of gaseous underground storage is the possible contamination of hydrogen with other gases such as H_2S and CH_4. For this reason a cost allowance for a hydrogen purification unit has been considered in the scenarios including storage. Another concern is the possibility of leakage of hydrogen through the storage walls, which is strongly dependent on the type of storage environment (for example the leaks in underground caverns are evaluated to be 1-3% of the total volume per year).
- Other types of hydrogen storage have been evaluated; liquefied storage and aboveground storage have been excluded because of their huge cost; storage in metal hydride form is not suitable for large quantities;

storage of hydrogen in pipelines poses significant challenges and costs for pipelines design, due to the issues related to hydrogen leakage and metal embrittlement.

7.0 Glossary of terms

AGR	Acid gas removal
ASU	Air separation unit
BFW	Boiler feed water
CAPEX	Capital cost
COE	Cost of Energy
EE	Electric energy
EPC	Engineering, procurement & construction
EPRI	Electric Power Research Institute
FWI	Foster Wheeler Italiana
HHV	High heating value
HP	High pressure
HRSG	Heat recovery steam generator
IEA GHG	International Energy Agency - Greenhouse Gas R&D Programme
IGCC	Integrated gasification combined cycle
LHV	Low Heating Value
LP	Low pressure
MHP	Medium high pressure
MP	Medium pressure
NG	Natural gas
O&M	Operation and Maintenance
OPEX	Operative cost
PSA	Pressure swing adsorption
SCGP	Shell coal gasification process
SRU	Sulphur Recovery unit
TGP	Texaco gasification process
TGT	Tail gas treatment
VLP	Very low pressure

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : GENERAL INFORMATION

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Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev. 1	L. Valota	P. Cotone	S. Arienti

GENERAL INFORMATION

I N D E X

GENERAL INFORMATION

1.0 Purpose of the Study

IEA Greenhouse Gas R&D Programme (IEA GHG) retained Foster Wheeler to investigate alternative power and hydrogen generation plant designs, based on high rank coal gasification, aimed at assessing the potential advantage of flexible co-production of hydrogen and electricity with capture of CO_2.

The primary purpose of this study is, therefore, the evaluation of the technologies and the process alternatives that can be used in these complex power and hydrogen generation schemes to optimize efficiency and capital cost and reduce, at the same time, emissions to the atmosphere.

The plant of the study has a nominal capacity of 750 MWe and is fed with a typical coal having a low heating value (LHV) equal to 25870 kJ/kg and a sulphur content equal to 1.1 % wt.

The study is based on the current state-of-the-art technologies, evaluating costs and performances of plants which can be presently engineered and built.

2.0 Project Design Bases

The IGCC Complex is designed to process, in an environmentally acceptable manner, an open-cut coal from eastern Australia and produce electric energy (750 MWe nominal capacity) to be delivered to the local grid.

The Power Island inside the IGCC Complex is also able to process Natural Gas as back-up fuel.

2.1 Feedstock Specification

The feedstock characteristics are listed hereinafter.

2.1.1 Design Feedstock

	Eastern Australian Coal Proximate Analysis, wt%
Inherent moisture	9.50
Ash	12.20
Coal (dry, ash free)	78.30
Total	100.00

	Ultimate Analysis, wt% (dry, ash free)
Carbon	82.50
Hydrogen	5.60
Nitrogen	1.77
Oxygen	9.00
Sulphur	1.10
Chlorine	0.03
Total	100.00
Ash Fluid Temperature at reduced atm., °C	1350
HHV (Air Dried Basis), MJ/kg (*)	27.06
LHV (Air Dried Basis), MJ/kg (*)	25.87
Grindability, Hardgrove Index	45

(*) based on Ultimate Analysis, but including inherent moisture and ash.

2.1.2 Back-up Fuel

	Natural Gas Composition, vol%
- Nitrogen	0.4
- Methane	83.9
- Ethane	9.2
- Propane	3.3
- Butane and C5	1.4
- CO_2	1.8
Total	100.0
- Sulphur content (as H_2S), mg/Nm3	4
LHV, MJ/Nm^3	40.6
Molecular weight	19.4

The gas specification is based on a pipeline quality gas from the southern part of the Norwegian off-shore reverses.

2.2 Products and by-products

The main products and by-products of the IGCC Complex are listed here below with their specifications.

2.2.1 Electric Power

Net Power Output	:	750	MWe	nominal capacity
Voltage	:	380	kV	
Frequency	:	50	Hz	
Fault duty	:	50	kA	

2.2.2 Hydrogen

The Hydrogen characteristics at IGCC B.L. are the following:

Composition	:	
Hydrogen (% vol)	:	>99.5
$CO + CO_2$ (ppm)	:	10 max
CO	:	10 max
H_2S, HCL, COS, HCN, NH_3	:	free
$N_2 + Ar$	:	balance
Pressure	:	20-25 barg (to be confirmed based on gasification pressure and both syngas treatment and PSA Unit pressure losses)

2.2.3 Carbon Dioxide

The Carbon Dioxide characteristics at IGCC B.L. are the following:

Status	:	supercritical
Pressure	:	110 bar g
Temperature	:	30 °C
Purity	:	(1)
H_2S content	:	0.1 % wt (max)
Moisture	:	<0.1 ppmvd
Non-CO_2 content	:	4% max

(1) Depending on the process alternative considered

Minimum Capture level	:	80%
Preferred Capture level	:	85%

2.2.4 Sulphur

Sulphur is a by-product of the IGCC Complex for all the process alternatives considered.

Status	:	solid/liquid
Color	:	bright yellow
Purity	:	99.9 % wt. S (min)
H_2S content	:	10 ppm (max)
Ash content	:	0.05 % wt (max)
Carbonaceous material	:	0.05 % wt (max)

2.2.5 Solid By-products

The IGCC Complex produces solid by-products that are saleable, in particular:

- flyash
- slag
- filter cake

Type and water content in slag and filter cake, depending on gasification technology.

2.3 Environmental Limits

The environmental limits set up for the IGCC Complex are outlined hereinafter.

2.3.1 Gaseous Emissions

The overall gaseous emissions from the IGCC Complex referred to dry flue gas with 15% volume O_2 shall not exceed the following limits:

NOx(as NO_2)	:	≤	80	mg/Nm^3
SOx(as SO_2)	:	≤	10	mg/Nm^3
Particulate	:	≤	10	mg/Nm^3
CO	:	≤	50	mg/Nm^3

Lower emissions for NOx and CO, will be investigated based on GT performances.

2.3.2 Liquid Effluent

Characteristics of waste water discharged from the IGCC Complex shall comply with the limits stated by the following EU directives:

- 1991/271/EU
- 2000/60/EU

The effluent from the Waste Water Treatment shall be generally recovered and recycled back to the Gasification Island as process water.

The only continuous liquid effluent from the IGCC Complex is the seawater return stream. Main characteristics of the water are listed in the following:

• Temperature	:	19	°C
• Cl_2	:	<0.05	ppm

2.3.3 Solid Wastes

The process does not produce any solid waste, except for typical industrial plant waste e.g. (sludge from WasteWater Treatment etc.). However even the wastewater sludge is recovered and recycled back to the Gasification Island to be processed by the Gasifiers.

2.3.4 Noise

All the equipment of the IGCC Complex will be designed to obtain a sound pressure level of 85 dB(A) at 1 meter from the equipment.

2.4 IGCC Operation

2.4.1 Capacity

For the base case, the design capacity is fixed to match the appetite of 2x400 MWe combined cycles.

A minimum equivalent availability of 85% corresponding to 7446 hours of syngas operation in one year at 100% capacity is expected for all the alternatives starting from the second year of commercial operation.

The whole gasification train from the Gasification Unit to the Power Island is designed to operate at 100% of nominal design capacity, even though the single Units may have a design capacity selected on the basis of specific criteria.

The Air Separation Unit capacity is defined by oxygen requirements of the IGCC Complex (mainly the gasifiers requirement plus the marginal consumption of Sulphur Recovery Unit). ASU is also requested to produce nitrogen at different levels of pressure to be supplied to the IGCC complex. Nitrogen production is dependent on oxygen production, consequently nitrogen flowrate available for syngas dilution may be different case by case, based on the other requirements of the IGCC Complex. The ASU is partially integrated with the Gas Turbines: the air needed by the ASU is partly supplied by the gas turbine and partly by a separate air compressor. The integration between two major components of the IGCC, i.e. the gas turbine and the Air Separation Unit represents an important potential benefit.

The Sulphur Recovery Unit consists of two trains at 100% capacity due to the low reliability of these units. The Tail Gas Treatment consists in a Hydrogenation step plus gas scrubbing sections and a dedicated compressor to recycle the stream back to the AGR Unit. This Unit is designed for 100% of the max tail gas production of the SRU.

2.4.2 Unit Arrangement

The IGCC Complex is in part a twin or multiple train facility due to constraints on equipment size and/or reliability reasons. The arrangement of the process units is as follows:

Process Units		Trains
1000	Gasification	1 x 100%
	gasifiers	(*)
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	(*)
2300	AGR	(*)
2400	SRU	2 x 100%
	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50%
Power Island (Unit 3000)		
	Gas Turbine	(*)
	HRSG	(*)
	Steam Turbine	(*)

(*)Depending on the process alternative and the technology considered.

2.4.3 Turndown

The IGCC Complex is designed to operate with a large degree of flexibility in terms of turndown capacity and feedstock characteristics.

The Gasification Unit will be composed of multiple gasifiers, at least two, thus allowing to operate at low loads with respect to the IGCC design capacity, the turndown of the single gasifier being 50%.

Most other Units are based on twin trains (50% capacity each) thus limiting the events causing the shutdown of the entire IGCC Complex or of the entire Gasification Island. This ensures a large availability of syngas production, at least at reduced load, which ensures a high power production by co firing syngas and natural gas in the gas turbines and a high hydrogen production.

The minimum turndown of each Gas Turbine on syngas is 20% as far as electrical generation is concerned. The minimum turndown of the Power Island when all the machines are in operation (two Gas Turbines and one Steam Turbine) is about 25% of the IGCC capacity. This figure should be verified with GT emissions at reduced load.

The Hydrogen production plant turndown is 35% per train. Based on the flowrate of Hydrogen produced, the Unit could have multiple trains configuration, further reducing the minimum turndown.

In conclusion, even if the IGCC complex operation at 25% load is a necessary step of the start-up procedure, its duration has to be limited. In fact, during the prolonged continuous operation, the load is expected to be 35%.

2.5 Location

The site is a green field located on the NE coast of The Netherlands.
No special civil works implications are assumed. The plant area is assumed to be close to a deep sea, thus limiting the length of the sea water lines (both the submarine line and the sea water pumps discharge line). The site is also close to an existing harbor equipped with a suitable pier and coal bay to allow coal transport by large ships and a quick coal handling.

2.6 Climatic and Meteorological Information

The conditions marked (*) shall be considered reference conditions for plant performance evaluation.

- atmospheric pressure

	: 1013	mbar	(*)

- relative humidity

average	: 60	%	(*)
maximum	: 95	%	
minimum	: 40	%	

- ambient temperatures

minimum air temperature	: -10	°C	
maximum air temperature	: 30	°C	
average air temperature	: 9	°C	(*)

2.7 Economic/Financial Factors

2.7.1 Design and Construction Period

IGCC design and construction will be completed in 34 months starting from issue of Notice to Proceed to the EPC contractor. Overnight construction will be applied.
The curve of capital expenditure during construction is assumed to be:

Year	Investment Cost %
1	20
2	45
3	35

2.7.2 Capital Charges

Discounted cash flow calculations will be expressed at a discount rate of 10% and to illustrate sensitivity at 5%.

2.7.3 Cost of Debt

All capital requirements will be treated as debt at the same discount rate used to derive capital charges. This is equivalent to assuming 100% equity. No interest during construction is applied but the timing of capital expenditure is taken into account in the discounted cash flow analysis.

2.7.4 Inflation

No inflation shall be applied to the economical analysis.

2.7.5 Commissioning

IGCC commissioning will take a 6 month period during the last two months of the third year of construction and the first four months of first year of IGCC operation.

Note: The commissioning duration has been modified, with respect to the three months proposed, as agreed in previous study made by Foster Wheeler Italiana for IEA GHG (Gasification Power Generation Study – 2003).

2.7.6 Working Capital

Sufficient storage for 30 days operation at rated capacity will be allowed for raw materials, products, and consumables. No allowance will be made for receipts from sales in this period.

2.7.7 Land purchase, surveys, general site preparation

5% of the installed plant cost is assumed.

2.7.8 Fees

2% of the installed plant cost is assumed to cover process/patent fees, consultant services other than EPC Contractor's services, fees for agents, legal and planning costs. This item is part of the capital cost.

2.7.9 Operation and Maintenance

Labour and maintenance data used for the economical evaluation are summarized in Section E, paragraph 4.0.

2.7.10 Taxation and Insurance

1% of the installed plant cost per year is assumed to cover local taxation. Taxation on profits is not included. The same percentage of the installed plant cost per year is assumed for insurance.

2.7.11 Fuel Costs

Cost of coal delivered to site is 1.5 $/GJ.
Cost of natural gas delivered by a pipeline to site is 3 $/GJ.

2.7.12 Hydrogen Price

Hydrogen price is 8.799 €/GJ (0.095 €/Nm^3)

2.7.13 By-Products Price

Sulphur Price is 103.3 €/t.

2.7.14 Currency exchange rate

The currency exchange rate used is 1.25 $/Euro.

2.7.15 CO_2 price

No selling price is attributed to CO_2.

2.8 Software Codes

For the development of the Study, two software codes have been mainly used:

- HYSYS v3.2 (by Hyprotech Ltd.): Process Simulator used for syngas treatment and conditioning line simulation of the Process Units downstream the Gasification Island.
- Gate Cycle v5.51.0 (by General Electric): Simulator of Power Island used for Combined Cycle Unit simulation.

3.0 Basic Engineering Design Data

Scope of the Basic Engineering Design Data is the definition of the common bases for the design of all the units included in the Integrated Gasification Combined Cycle (IGCC) Complex to be built on the east coast area of Netherlands.

The IGCC Plant is constituted by the following groups of units:

Process Units (Unit 900 to 2500) including:
- Coal Handling and Storage (Unit 900);
- Gasification Island (Unit 1000);
- Air Separation Unit (Unit 2100);
- Syngas Treatment and Conditioning Line (Unit 2200);
- Acid Gas Removal Unit (Unit 2300);
- Sulphur Recovery and Tail Gas Treatment (Unit 2400);
- CO_2 Compression and Drying (Unit 2500);
- H_2 production (Unit 2600).

Power Island including:
- Gas Turbines (Unit 3100);
- Heat Recovery Steam Generators (Unit 3200);
- Steam Turbine (Unit 3300);
- Electrical Power Generation (Unit 3400).

Utility and Offsite Units providing services and utility fluids to all the units of the plant; including:
- Sea Cooling Water/Machinery Cooling Water Systems (Unit 4100);
- Demineralized, Condensate Recovery, Plant and Potable Water Systems (Unit 4200);
- Natural Gas System (Unit 4300);
- Plant/Instrument Air Systems (Unit 4400);
- Waste Water Treatment (Unit 4600);
- Fire fighting System (Unit 4700);
- Flare (Unit 4800);
- Chemicals (Unit 4900);
- Solid (Slag & Flyash or Filtercake) Handling (Unit 5000);
- Sulphur Storage and Handling (Unit 5100);
- Interconnecting (instrumentation, DCS, piping, electrical, 400 kV substation) (Unit 5200).

3.1 Units of Measurement

All calculations are and shall be in SI units, with the exception of piping typical dimensions, which shall be in accordance with ANSI.

3.2 Site conditions

- site elevation
  IGCC complex area : 6 m above mean sea level.

- atmosphere type : coastal area with salt pollution.

3.3 Climatic and Meteorological Information

Reference is made to para. 2.6 for main data.

Other data:

- rainfall
  design : 25 mm/h
  50 mm/day

- wind
  maximum speed : 35 km/h

- snow
  : 50 kg/m^2

- winterization
  winterization is required.

- sea water supply temperature and salinity
  average (on yearly basis) : 12 °C
  maximum average (summer) : 14 °C
  minimum average (winter) : 9 °C

  salinity : 22 g/l

3.4 Soil data

- earthquake
  earthquake factor : negligible

- geology
  green field site with no special civil works implications.

3.5 Project Battery Limits design basis

3.5.1 Electric Power

High voltage grid connection:	380 kV
Frequency:	50 Hz
Fault duty:	50 kA

3.5.2 Process and Utility Fluids

The streams available at plant battery limits are the following:

- Coal;
- Natural Gas;
- Sea water supply;
- Sea water Return;
- Plant/Raw/Potable water;
- Sulphur product;
- CO_2 rich stream;
- Hydrogen stream.

3.6 Utility and Service fluids characteristics/conditions

In this paragraph are listed the utilities and the service fluids distributed inside the IGCC Complex.

3.6.1 Cooling Water

The IGCC primary cooling system is sea water in once through system.

Sea Cooling Water (primary system)

Source	: sea water in once through system
Service	: for steam turbine condenser, ASU exchangers, CO_2 compression and drying exchangers, fresh cooling water-cooling.
Type	: clear filtered and chlorinated, without suspended solids and organic matter.

Supply temperature:

- average supply temperature (on yearly basis)	: 12	°C
- max supply temperature (average summer)	: 14	°C
- min supply temperature (average winter)	: 9	°C
- max allowed sea water temperature increase	: 7	°C

Return temperature:

- average return temperature	: 19	°C
- max return temperature	: 21	°C

Operating pressure at Users inlet	: 0.9	barg
Max allowable ΔP for Users	: 0.5	barg
Design pressure for Users	: 4.0	barg
Design pressure for sea water line	: 4.0	barg
Design temperature	: 55	°C
Cleanliness Factor (for steam condenser)	: 0.9	
Fouling Factor	: 0.0002	h °C m^2/kcal

Fresh Cooling Water (secondary system)

Service	: for machinery cooling and for all IGCC users other than steam turbine condenser, ASU and CO_2 compression and drying exchangers.
Type	: demiwater stabilized and conditioned.

Supply temperature:

- max supply temperature	: 17	°C
- min supply temperature	: 13	°C
- max allowed temperature increase	: 12	°C
- design return temperature for fresh cooling water		

cooler : 29 °C

Operating pressure at Users	: 3.0 barg
Max allowable ΔP for Users	: 1.0 bar
Design pressure	: 5.0 barg
Design temperature	: 60 °C
Fouling Factor	: 0.0002 h °C m^2/kcal

3.6.2 Waters

Potable water

Source : from grid
Type : potable water

Operating pressure at grade	: 0.8 barg (min)
Operating temperature	: Ambient
Design pressure	: 5.0 barg
Design temperature	: 38 °C

Raw water

Source : from grid
Type : potable water

Operating pressure at grade	: 0.8 barg (min)
Operating temperature	: Ambient
Design pressure	: 5.0 barg
Design temperature	: Ambient

Plant water

Source : from storage tank of raw water
Type : raw water

Operating pressure at grade	: 3.5 barg
Operating temperature	: Ambient
Design pressure	: 9.0 barg
Design temperature	: 38°C

Demineralized water

Type : treated water (mixed bed demineralization)

Operating pressure at grade	:	5.0	barg
Operating temperature	:	Ambient	
Design pressure	:	9.5	barg
Design temperature	:	70	°C

Characteristics:

- pH		6.5÷7.0	
- Total dissolved solids	mg/kg	0.1	max
- Conductance at 25°C	µS	0.15	max
- Iron	mg/kg as Fe	0.01	max
- Free CO_2	mg/kg as CO_2	0.01	max
- Silica	mg/kg as SiO_2	0.015	max

3.6.3 Steam, Steam Condensate and BFW

Steam

These conditions refer to the Process Units. Inside Power Island the steam levels are different even if interconnected to the Process Units (see INTRODUCTION-List of units).

Table B.3.1 – Process Units steam conditions.

	Pressure, barg			Temperature, °C	
	Max	**Min**	**Design**	**Norm**	**Design**
High Pressure **(HP)** Nominal Pressure: 160 barg	170	160	187	353	370
Medium High Pressure **(MHP)** Nominal Pressure: 70 barg	76	70	84	288	310
Medium Pressure **(MP)** Nominal Pressure: 40 barg	43	40	47	256	270
Low Pressure **(LP)** Nominal Pressure: 6.5 barg	8.0	6.5	12	175	250
Very Low Pressure **(VLP)** Nominal Pressure: 3.2 barg	4.0	3.2	12	152	250

Note: Based on Shell gasification technology, different conditions for each case.

In the table above:

- The maximum value indicates the steam generation pressure to be adopted for steam generators in the Process Units.

- The minimum pressure indicates the steam pressure available for steam users.
- The normal Temperature indicates the *saturation T* corresponding to the Max Pressure indicated.

Cold condensate

Type: condensate from Power Island plus (demineralized water make up)

Supply:

Operating pressure at Users	: 16	barg
Operating temperature	: 21	°C
Design pressure	: 22	barg
Design temperature	: 50	°C
Fouling Factor	: 0.0001	h °C m^2/kcal

Return:

Operating pressure	: 9.9	barg
Operating temperature	: (*)	
Design pressure	: 22.8	barg
Design temperature	: 130	°C
Fouling Factor	: 0.0002	h °C m^2/kcal

(*) Depending on the process alternative and technology considered.

Steam Condensate from process, utility and off site units

Steam condensate will be flashed within process units whenever possible to recover steam and piped back to the condensate collection header.
The condensate collection header shall have the following characteristics:

Operating pressure for other Units B.L.	: 1	barg
Operating temperature	: 94	°C
Design pressure	: 12.0	barg
Design temperature	: 250	°C

Boiler Feed Water

The main characteristics of the Boiler Feed Water at Units B.L. is shown in the following table.

Table B.3.2 – Boiler Feed Water at units B.L.

	Pressure Barg	**Temperature °C**
	Normal	**Normal**
Boiler Feed Water, Very Low Pressure **(BWV)**	15	120
Boiler Feed Water, Low Pressure **(BWL)**	15	160
Boiler Feed Water, Medium Pressure **(BWM)**	60	160
Boiler Feed Water, High Pressure **(BWH)**	195	160

3.6.4 Instrument and Plant Air

Instrument air

Operating pressure			
- normal	:	7.0	barg
- minimum	:	5.0	barg
Operating temperature	:	40	°C (max)
Design pressure	:	10.0	barg
Design temperature	:	60	°C
Dew point @ 7 barg	:	-30	°C

Plant air

Operating pressure	:	7.0	barg
Operating temperature	:	40	°C (max)
Design pressure	:	10.0	barg
Design temperature	:	60	°C

3.6.5 Nitrogen

Low Pressure Nitrogen

Supply pressure : 6.5 barg

Supply temperature	:	15 °C min
Design pressure	:	11.5 barg
Design temperature	:	70 °C
Min Nitrogen content	:	99.9 % vol.

<u>Medium Pressure Nitrogen (Syngas dilution)</u>

Supply pressure	:	30 barg
Supply temperature	:	210 °C
Design pressure	:	35 barg
Design temperature	:	240 °C
Min Nitrogen content	:	98 % vol.

<u>Medium Pressure Nitrogen (GT injection)</u>

Supply pressure	:	26 barg
Supply temperature	:	213 °C
Design pressure	:	35 barg
Design temperature	:	240 °C
Min Nitrogen content	:	98 % vol.

<u>High Pressure Nitrogen</u>

Supply pressure	:	(*)
Supply temperature	:	15 °C min
Design pressure	:	(*)
Design temperature	:	(*)
Min Nitrogen content	:	99.9 % vol.

(*) Depending on the process alternative considered.

3.6.6 <u>Natural Gas</u>

Characteristics of Natural Gas are listed at para 2.1.2, Project Design Bases.

<u>High Pressure</u>

Type : natural gas.
Service : gas turbine and gasification island start-up and back-up fuel

Operating pressure at Users	:	27.0 barg
Operating temperature at Users	:	30°C above natural gas dew point
Design pressure	:	33.0 barg
Design temperature	:	70 °C

<u>Low Pressure</u>

Type : natural gas.
Service : distribution.

Operating pressure at Users	:	3.5	barg
Operating temperature at Users	:	30	°C
Design pressure	:	6.0	barg
Design temperature	:	60	°C

Characteristics: as for High Pressure Natural Gas.

3.6.7 Oxygen

The Oxygen for the gasification unit has the following characteristics:

Supply pressure	:	(*)
Supply temperature	:	(*)
Design pressure	:	(*)
Design temperature	:	(*)

(*) Depending on the process alternative considered.

Purity	:	95.0	% mol. O_2 min
		3.5	% mol Ar
		1.5	% mol N_2
H_2O content	:	1.0	ppm max
CO_2 content	:	1.0	ppm max
HC as CH_4 (number of times the content in ambient air)	:	5	max

Oxygen for Sulphur plant

Supply pressure at IGCC BL	:	5.0	barg
Supply temperature	:	15	°C min
Design pressure	:	8.0	barg
Design temperature	:	50	°C
Purity	:	95	% mol. O_2 min

3.6.8 Chemicals

Caustic Soda

A concentrated (50% by wt) NaOH storage tank is foreseen and used to unload caustic from trucks.

Concentrated NaOH is then pumped and diluted with demineralized water to produce 20% by wt NaOH accumulated in a diluted NaOH storage tank.
The NaOH solution is distributed within IGCC with the following characteristics:

Supply temperature, °C	Ambient
Design temperature, °C	70
Supply pressure (at grade) at unit BL barg	3.5
Design pressure barg	9.0
Soda concentration wt %	20

Hydrochloric Acid

Two concentrated (20% by wt) HCl storage vessels are foreseen and used to unload hydrochloric acid from trucks.
Concentrated HCl is pumped to users where is firstly diluted if necessary.

Supply temperature, °C	Ambient
Design temperature, °C	70
Supply pressure (at grade) at unit BL barg	2.5
Design pressure barg	5.0
Hydrochloric concentration wt %	20

3.6.9 Electrical System Distribution

The voltage levels foreseen inside the plant area are as follows:

	Voltage level (V)	*Electric Wire*	*Frequency (Hz)*	*Fault current duty (kA)*
Primary distribution	66000 ± 5%	3	50 ± 0.2%	31.5 kA
MV distribution and utilization	11000 ± 5% 6000 ± 5%	3 3	50 ± 0.2% 50 ± 0.2%	31.5 kA 25 kA
Emergency power source	6000 ± 5%	3	50 ± 0.2%	31.5 kA
LV distribution and utilization	400/230V±5%	3+N	50 ± 0.2%	50 kA
Uniterruptible power supply	230 ± 1% (from UPS)	2	50 ± 0.2%	12.5 kA
DC control services	110 + 10%-15%	2	-	-
DC power services	220 + 10%-15%	2	-	-

3.7 Plant Life

The IGCC Plant is designed for a 25 years life, with the following considerations:

- Design life of vessels, equipment and components of equipment will be as follows:
  - 25 years for pressure containing parts;
  - 5 years for replaceable parts internal to static equipment.

- Design life of piping will be 10 years.

- For rotating machinery a service life of 25 years is to be assumed as a design criterion, taking into account that cannot be applicable to all parts of machinery for which replacement is recommended by the manufacturer during the operating life of the unit, as well as to small machinery, machines on special or corrosive/erosive service, some auxiliaries and mechanical equipment other than rotating machinery.

3.8 Codes and standards

The project shall be in accordance to the International and EU Standard Codes.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : BASIC INFORMATION FOR THE IGCC COMPLEX

ISSUED BY : L. VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev. 1	L. Valota	P. Cotone	S. Arienti

SECTION C

BASIC INFORMATION FOR THE IGCC COMPLEX

I N D E X

SECTION C

1.0 Gasification Island Below are listed the component parts of this section.

1.1 Shell Technology

The purpose of the attached document “Shell Gasification Island” is to summarize the information used for the Hydrogen and Electricity Co-production study. In particular these data were the basis in the first step of the study for the selection of the gasification technology for the IGCC (section D.2) and for configurations with and without hydrogen production (section G).

Technical data of the IGCC have been taken from a previous study made by FWI for IEA GHG (Gasification Power generation study (2003)) and have been reviewed with Shell including minor changes and slight improvements. Investment data have been updated to 2007 by FWI Estimate Department and finally approved by Shell.

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME	:	GASIFICATION POWER GENERATION STUDY
CONTRACT NO.	:	1-BD-0337A
UNIT NO.	:	1000
DOCUMENT NAME	:	SHELL GASIFICATION ISLAND

ISSUED BY	:	L.VALOTA
CHECKED BY	:	P. COTONE
APPROVED BY	:	S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

SHELL GASIFICATION ISLAND

I N D E X

1.0 INTRODUCTION

Purpose of this document is to summarize the information received from SHELL for the first step of the Gasification Power Generation Study.

Technical relevant information of the IGCC have been taken from Gasification Power Generation study that FWI performed for IEA GHG in 2003. That study was based on the same coal as the present study. In 2004, for a second study that FWI performed for IEA GHG, Shell communicated as improvement that the steam generation pressure in WHB could be reduced from HP (160 barg) to MHP (70 barg) in order to reduce the investment cost. In conclusion, for the present, the 2003 study has been considered only changing the pressure level generation in WHB.

Investment data have been updated to 2007 by FWI Estimate Department.

They are the basis for the selection of the gasification technology and for the IGCC configurations with and without CO_2 capture, with and without hydrogen production.

2.0 GASIFICATION ISLAND PROCESS DESCRIPTION AND BLOCK FLOW DIAGRAM

2.1 General description of the Shell Coal Gasification Process

The basic concepts selected for the Shell Coal Gasification Process (SCGP) are:
- Pressurised: compact equipment;
- Entrained flow: compact gasifier;
- Oxygen blown: compact equipment, high gasification efficiency;
- Membrane wall, slagging gasifier: robustness, high temperature, insulation by slag layer;
- Opposed burners: good mixing, high conversion, scale-up possibility;
- Dry feed of pulverised coal: high gasification efficiency, high feed flexibility.

The process can handle a wide variety of solid fuels, ranging from bituminous coal to lignite, as well as petroleum coke (petcoke) in an environmentally acceptable way. The process produces a raw syngas and after gas treatment the high purity, medium-calorific gas can be used as a fuel for power generation, as a chemical feedstock or as a source of hydrogen.

The oxygen required in the SCGP gasification step is supplied by an air separation plant. Nitrogen from the air separation unit provides low-pressure and high-pressure nitrogen for use in the gasification plant, e.g., for transporting coal in the feed system. Milled and dried coal from the coal milling and drying unit is transported (pneumatically or by gravity) to the coal pressurisation and feeding system. Pressurised coal, oxygen and steam enter the gasifier through pairs of opposed burners. “Flux” can be added to a coal feed to ensure an appropriate slag flow from the gasifier, if it is required.

The gasifier operates at a pressure of 20 to 40 bar. The gasifier consists of a pressure vessel with a gasification chamber inside. The inner gasifier wall temperature is controlled by circulating water through the membrane wall to generate saturated steam. The membrane wall encloses the gasification zone from which two outlets are provided. One opening at the bottom of the gasifier is used for the removal of slag. The other outlet allows hot raw gas and fly slag to exit from the top of the gasifier.

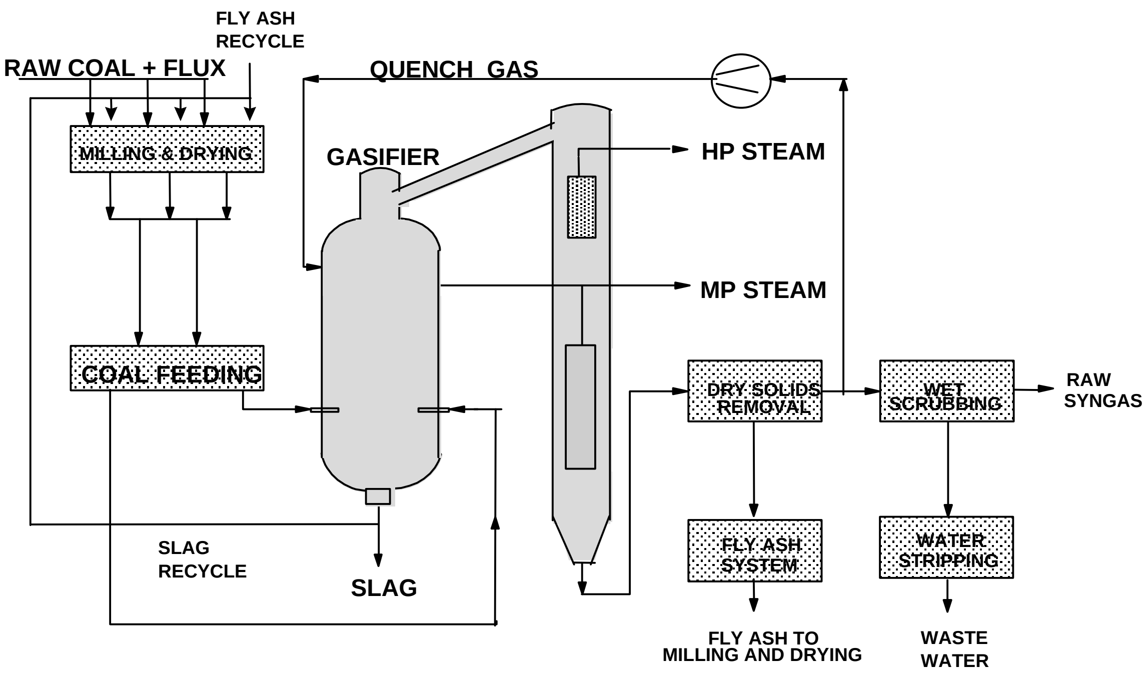
Most of the mineral content of the feed leaves the gasification zone in the form of molten slag. The high gasifier temperature (over 1500°C) ensures that the molten slag flows freely down the membrane wall into a water-filled compartment at the bottom of the gasifier. High carbon conversions (above 99%) are obtained, and the high temperature ensures that no organic components heavier than methane are in the raw syngas. The insulation provided by the slag layer in the gasifier inner membrane wall minimises heat losses, such that cold gas efficiencies are high and CO_2 levels in the syngas are low.

As the molten slag contacts the water bath, the slag solidifies into dense, glassy granules. The slag is washed, de-pressurised and then fed to intermediate storage for recycle (if required) and disposal.

The hot raw synthesis gas leaving the gasification zone is quenched with cooled, recycled synthesis gas to convert any entrained molten slag to a hardened solid material prior to entering the syngas cooler. The syngas cooler recovers high-level heat from the quenched raw gas by generating high-pressure steam, and steam at other desired pressure levels.

Virtually all fly slag contained in the raw gas leaving the syngas cooler is removed from the gas using commercially available equipment such as filters or cyclones. The recovered fly slag can be recycled back to the gasifier via the coal feeding system (if required). The syngas then goes to a scrubbing system, where the remaining traces of solids and water soluble contaminants are removed.

A bleed from the scrubbing system is sent to a sour slurry stripper. The water is then clarified and can be partially recycled to minimise the volume of effluent water.



THE SHELL COAL GASIFICATION PROCESS

2.2 Brief description of various process blocks

Reference is made to the attached Block Flow Diagram.

Coal Pressurisation and Feeding

Milled and dried coal from the coal milling and drying unit is pneumatically transported to the coal pressurisation and feeding system. This system consists of lock hoppers and feed hoppers. Once a lock hopper has been charged with coal, it is pressurised with nitrogen and its contents discharged into a feed hopper.
Pressurised coal is withdrawn from the feed hoppers and pneumatically conveyed with nitrogen to the gasifier's coal burners.
Lock hoppers are widely utilised in materials handling applications. They have proven to be a safe and reliable method for transferring solids under pressure.
The valves required for commercial scale lock hopper systems have been extensively demonstrated.

Gasification, Gas Quench and Slag Removal

A line-up of a single-train gasifier, hot-gas quench has been proposed.
In the top part of the gasifier, a solid-free cold syngas stream is injected to the hot syngas, so that the syngas is quenched to a temperature at which the flyash solidifies. The recycle quench gas is withdrawn from downstream of the dry solids removal unit. A recycle gas compressor is applied for this service.
At the bottom of the gasifier, as the molten slag falls into a water bath, the slag solidifies into dense, glassy granules. These slag granules fall into a collecting vessel located beneath the slag bath and are transferred to a lock hopper which operates on a timed cycle to receive the slag. After the lock hopper is filled, the slag is washed with clean make-up water to remove entrained gas and any surface impurities. After washing, the lock hopper is de-pressurised and the slag is fed to a de-watering bin. Commercially sized slag sluicing valves have been applied for this service.
This dewatering bin is equipped with a mechanical conveyor (drag chain) to lift the settled solids off the bottom of the vessel and deposit them on a conveyor belt for delivery to intermediate storage (conveyor belt and storage outside scope of this proposal).

High Temperature Gas Cooling

The hot raw syngas leaving the gasification zone is quenched with cooled, recycled quench gas to convert any entrained molten slag to a hardened solid material prior to entering the syngas cooler. The syngas cooler recovers high-level heat from the quenched raw syngas by generating steam. The gasifier and syngas cooler included in the SCGP plant operates similar to the water wall boilers which are widely used in

other utility processes.
A syngas cooler line-up has been selected for this proposal to maximise the heat recovery while maintaining operability. The steam system has been designed bearing efficiency and intrinsic safety in mind. The choice for three steam levels (HP, MP and LP) ensures a high efficiency. The HP and MP steam pressure levels have been selected higher than the syngas pressure in order to maximise safety and integrity. The MP steam pressure level has been selected as high as the HP in order to leave a positive effect on investment cost. LP steam is not produced inside the SGC for this reason but via a separate boiler. An economiser is installed to booster the efficiency further.

Dry Solids Removal

The bulk of the flyash contained in the raw gas leaving the syngas cooler is removed from the gas using a commercially demonstrated high pressure, high temperature (HPHT) filter. The flyash leaving the process is conveyed to a flyash lock hopper. After the lock hopper is filled, the flyash is purged with high-pressure nitrogen to remove any entrained raw gas. After purging the lock hopper, the flyash is pneumatically conveyed to a silo for intermediate storage. All vent gases from the flyash lock hopper and the storage silo are filtered of particles.
Flyash is finally disposed and could be sold and used as filling materials.
Normally, in case of coke gasification, flyash could be recycled and used as fluxant. On the contrary in case of coal gasification, it is not necessary to recycle back the flyash and it is possible to sell it.

Wet Scrubbing

The gas leaving the dry solids removal is further purified by passing through a wet scrubbing unit where any residual flyash is removed to a level of less than 1 ppm. This wet scrubbing system also removes other minor contaminants such as soluble alkali salts and hydrogen halides.
Make-up water is continuously added to the wet scrubbing unit to control the concentration of contaminants. To minimise the water use for the plant, recycle water from the sour water stripper unit is used for this make-up and this comprises the majority of the make-up water stream. A small bleed flow of the contaminated water is sent to the sour slurry stripping unit to remove the contaminants.
A scrubber outlet temperature of 128 °C has been generally selected. Higher exit temperatures are however possible by optimizing the heat recovery in the SGC. For the study alternatives with CO_2 capture and sour shift reaction, the temperature is increased up to 160 °C, with the consequent elimination of LP steam production in syngas cooling section.

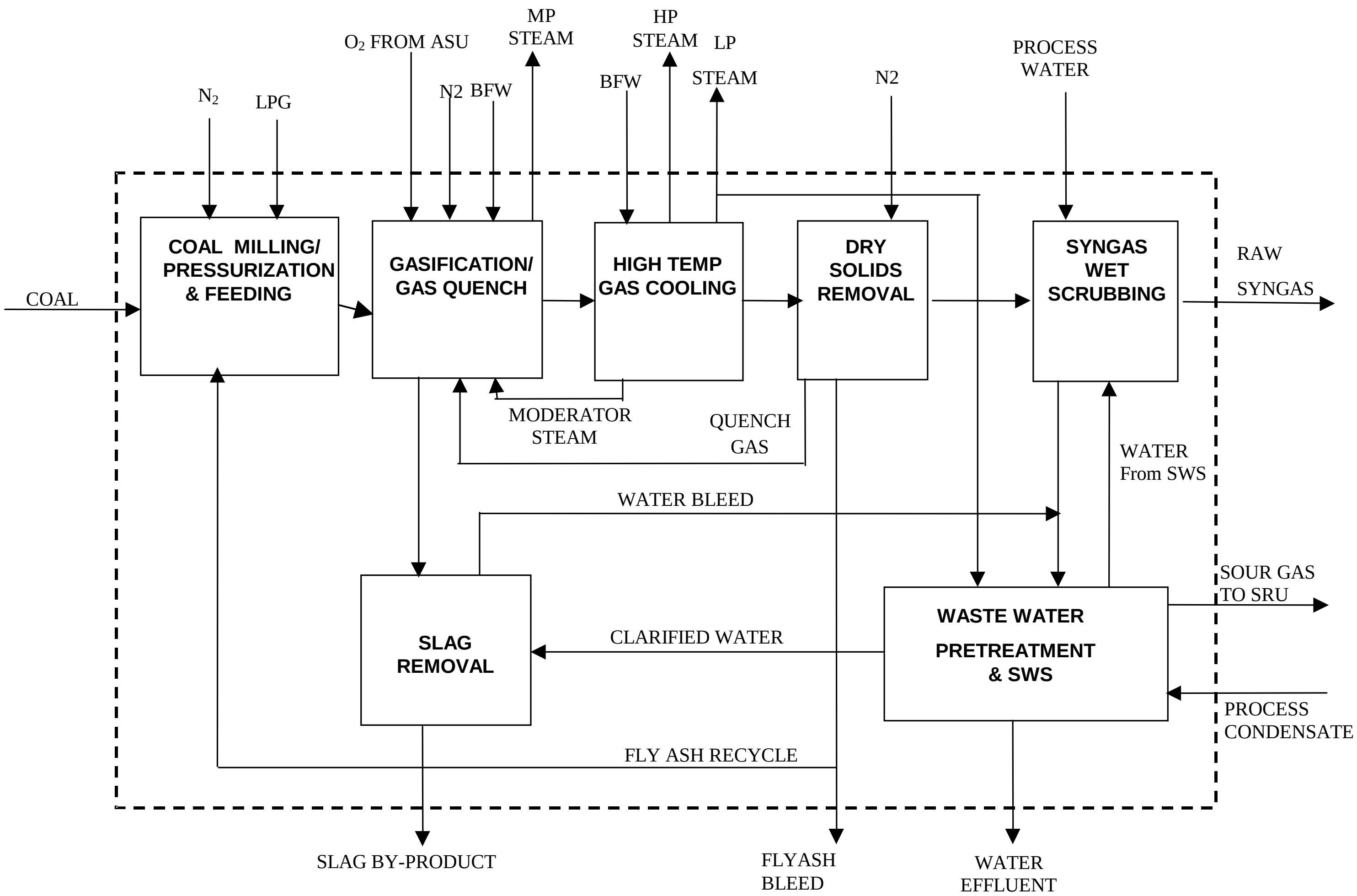
Sour Slurry Stripper (Waste Water Pretreatment)

The blow-down water from the wet scrubbing unit and a bleed from the slag bath are fed to a stripper for the removal of hydrogen sulphide, dissolved raw gases and to reduce the ammonia level in the water to an environmentally acceptable level. In this unit, low-pressure steam provides the necessary heat and stripping medium. A large portion of the effluent water from the stripper is recycled after clarification to the slag bath as make-up water. Only a small effluent water stream is sent to the OSBL Effluent Treating facilities (e.g. biotreater). In this way, the consumption of process water has been minimised.

Sour Water Stripper

Sour water streams from several sources in downstream OSBL units are stripped in this unit. Since we have no insight in all downstream units, we have assumed that any water condensed out of the syngas prior to the Acid Gas Removal unit will be supplied to this unit. In actual practice we expect a slightly higher volume of water to be treated. Since the column operates under non fouling conditions, the necessary stripping steam is supplied via a LP steam re-boiler. The vapour leaving the SWS column is sent to an overhead system. In this overhead system the overhead vapours are condensed and the sour gases are separated from the condensate in the gas/liquid separator. The condensed water is routed back to the SWS column as reflux, above the rectifying bed. The sour gases are routed to the battery limit. The SWS effluent has been used as make-up water in the wet scrubbing systems.

FIGURE 1
GASIFICATION ISLAND BLOCK FLOW DIAGRAM



3.0 PROCESS FLOW DIAGRAMS

The preliminary Process Flow Diagrams provided by SHELL are attached.

4.0 CHARACTERISTICS OF STREAMS AT GASIFICATION ISLAND BATTERY LIMITS.

The following Tables summarize the characteristics of Streams at Gasification Island Battery Limits for the cases 1 and 2.
The Cases differ for plant configuration and gasification pressure as follows:

1 Low Gasification pressure, IGCC w/o CO_2 capture
2 Low Gasification pressure, IGCC with CO_2 capture

Shell consider those cases as entirely proven concept.

TABLE 1

OVERALL PERFORMANCE

	Case 1	**Case 2**
Fresh Coal to Coal Grinding		
Proximate Analysis (%wt)		
Inherent moisture	9.5	9.5
Ash	12.2	12.2
Coal (dry, ash free)	78.3	78.3
Total	100.00	100.00
Flowrate (fresh, Air Dried Basis), t/h	250.6	273.1
Ultimate Analysis (%wt) (dry, ash free)		
Carbon	82.5	82.5
Hydrogen	5.6	5.6
Nitrogen	1.77	1.77
Sulphur	1.1	1.1
Oxygen	9	9
Chlorine	0.03	0.03
Total	100.00	100.00
Coal HHV (Air Dried Basis), kcal/kg	6464	6464
Coal LHV (A.D.B.), kcal/kg	6180	6180
Thermal Pow, MWt (LHV)	1800.8	1962.5

TABLE 1 (c'd)

	Case 1	Case 2
Characteristics of Syngas Ex Scrubber (Total)		
Composition, % mol		
CO	56.4	49.6
H_2	29.7	26.3
CO_2	1.4	1.2
H_2O	7.0	18.1
Ar	0.7	0.6
N_2	4.53	3.96
H_2S	0.24	0.21
COS	0.02	0.02
HCN	0.01	0.01
	100.00	100.00
Flowrate, kmol/h (1)	23,260	28,850
t/h	463.5	568.2
Pressure @ B.L., bar g	33	36
Temperature @ B.L., °C	126	160
Raw Syngas LHV, dry kcal/kg	2981.6	2490.6
Raw Syngas Thermal Power (LHV), MWt	1504.4	1638.2
Gasification eff. (LHV), %	83.5	83.5
Oxygen Consumptions		
O_2 Flowrate, t/h	197.0	214.6
O_2 Press @ B.L., barg	39.4	39.4
O_2 Temp @ B.L., °C	100	100
Nitrogen Consumptions		
HP N_2 Flowrate, t/h	82.0	87.0
HP N_2 Press @ B.L., barg	68	68
HP N_2 Temp @ B.L., °C	80	80
LP N_2 Flowrate, t/h	31.8	33.7
LP N_2 Press @ B.L., barg	6.5	6.5
LP N_2 Temp @ B.L., °C	70	70

(1) Clean syngas consumption for coal drying included

TABLE 2

STEAM PRODUCTIONS/BFW CONSUMPTIONS

	Case 1	Case 2
MHP Steam Production		
Flowrate, t/h	291.4	317.4
Pressure @ Unit B.L, barg	70	70
Temperature, °C	sat	sat
LP Steam Production		
Flowrate, t/h	57	-
Pressure @ Unit B.L, barg	6.5	-
Temperature, °C	168	-
MHP BFW Consumption		
Flowrate, t/h	403.7	390.9
Pressure @ Unit B.L., barg	85	85
Temperature, °C	160	160
LP BFW Consumption		
Flowrate, t/h	11.3	-
Pressure @ Unit B.L, barg	17	-
Temperature, °C	160	-
Steam Condensate		
Flowrate, t/h	37.6	41.3

TABLE 3

	Case 1	Case 2
Slag		
Total Dry, kg/h	37,200	40,500
Water, % wt	10	10
Fly ash		
Flowrate, kg/h	1200	1330
Temperature, °C	80	80

5.0 UTILITY AND CHEMICAL CONSUMPTIONS.

Table 4 summarizes the utility continuous consumptions (other than steam and Nitrogen) estimated for the two cases.

TABLE 4

	Case 1	Case 2
Fresh Cooling Water, m^3/h	233	248
Absorbed Electric Pow, kW	12,000	12,700
Instrument Air, Nm^3/h	700	700

Caustic solution is injected to the wet scrubbing unit to maintain the pH value of the circulating water slightly above neutral. For the same reason, HCl is added to the primary water treatment unit to prevent fouling.

6.0 EQUIPMENT LIST

Major Equipment related to the SHELL Gasification Island are presented in the attached Equipment List.

The main process units consist of two 50% trains as detailed in the Equipment List. Even if the capacity of each gasifier is significantly higher than the Buggenum capacity, the required scale up (approx. + 60%) is not seen by Shell as a risk. They have designed and offered gasifiers at even higher throughput.
For IGCC generating electric power only, Shell do not recommend to install overcapacity in the Gasification Island, but only to have natural gas available as back-up for the Combined Cycle.

MAIN EQUIPMENT LIST

The first numbers give the number of systems, the second number gives the fraction of the total plant capacity.

Unit 1100 - Coal Milling and Drying (4 x 33% trains)

4	33.3%	Raw Coal Bunker
4	33.3%	Raw Coal Bunker Bag Filter and Exhaust Fan
4	33.3%	Gravimetric Coal Weigh Feeder
4	33.3%	Flux Bunker(*)
4	33.3%	Flux Bunker Bag Filter and Exhaust Fan (*)
4	33.3%	Gravimetric Flux Weigh Feeder
4	33.3%	Coal Mill
4	33.3%	Rotary Classifier
4	33.3%	Inert Gas Generator
4	33.3%	Circulation Gas Fan
4	33.3%	Combustion Air Blower
4	33.3%	Seal Air Fan
4	33.3%	Dilution Air Fan
4	33.3%	Pulverised Coal Bag Filter
8	17%	Pulverised Coal Bag Filter Discharge Screws
8	17%	Pulverised Coal Rotary Feeders
8	17%	Pulverised Coal Screw Conveyors

(*) These are required when gasifying coals need fluxing, as in the present case.

Unit 1200 - Coal Pressurisation & Feeding (6 x 20% trains)

6	20%	Pulverised Coal Storage Vessel
6	20%	Pulverised Coal Storage Vessel Bag Filter
6	20%	Pulverised Coal Storage Bag Filter Discharge Screw
6	20%	Pulverised Coal Storage Bag Filter Rotary Feeder
6	20%	Coal Sluice Vessel
6	20%	Coal Sluice Vessel HP Filter
6	20%	Coal Feed Vessel
2	50%	Flyash Buffer Vessel
6	20%	Flyash Buffer Vessel Rotary Feeder

Unit 1300 - Gasification, Quenching, Syngas Cooling (2 x 50 trains)

2	50 %	Gasifier, which includes MHP evaporator membrane wall Quench section with MHP evaporator Duct between gasifier and SGC with MHP evaporator Slag bath
2	50%	Syngas Cooler (SGC) which includes MHP superheater MHP evaporator MHP economiser
2	50%	LP Steam Generator
4	50%	MHP Circulation Pump for syngas cooler and syngas duct sections
6	25%	MHP Circulation Pump for gasifier membrane wall
4	50%	MHP Circulation Pump for syngas cooler economiser
2	50%	MHP Steam Drum
2	25%	MHP Steam Drum
12	16.7%	Coal Burners (6 per each gasifier)
2	100%	Start up Burner (1 per each gasifier)
2	100%	Ignition Burner (1 per each gasifier)
2	50%	Oxygen Preheater
2	130%	Quench Gas Compressor
4	50%	Burner Cooling Water Circulation Pump
2	50%	Burner Cooling Water Buffer Vessel
2	50%	Burner Cooling Water Circulation Heater

Unit 1400 - Slag Removal (2 x 50% trains)

2	50%	Slag Crusher
2	50%	Slag Accumulator
2	50%	Slag Sluice Vessel
2	50%	Slag De-watering Silo with Drag Chain
2	50%	Slag Conveyor (outside Shell scope)
4	50%	Slag Bath Circulation Pump
4	25%	Slag Bath Circulation Cooler
2	50%	Slag Sluice Water Clarifier
4	50%	Clarifier Overflow Pump
4	50%	Clarifier Bottom Pump
2	50%	Slag Sluice Water Buffer Tank
4	50%	Slag Sluice Vessel Fill Pump
4	50%	Slag Sluice Support Pump

4	50%	Slag De-watering Silo Slurry Pump
4	50%	Slag Sludge Pump

Unit 1500 - Dry Solids Removal (2 x 50% trains)

2	50%	HPHT Ceramic Candle Filter includes Cleaning system with buffer volume
2	50%	Flyash Sluice Vessel
2	50%	Flyash Sluice Vessel Vent Filter
1	100%	Flyash Sluice Vessel Nitrogen Buffer Vessel
1	100%	Flyash Stripping/cooling Vessel
1	100%	Flyash Stripping/cooling Vessel filter
1	100%	Flyash Stripping/cooling Vessel Nitrogen Buffer Vessel
1	100%	LP Nitrogen Buffer Stripper Filter
1	100%	LP Nitrogen Buffer Storage Filter
1	100%	LP Nitrogen Heater
1	100%	Flyash Intermediate Storage Silo
1	100%	Flyash Intermediate Storage Silo Filter
1	100%	Flyash Blow Egg
1	100%	Flyash Pick-up
1	100%	Flyash Storage Silo
1	100%	Flyash Storage Silo Filter
1	100%	Rotary Ash Feeder
4	50%	Flyash Recycle or Disposal System

Unit 1600 - Wet Scrubbing (2 x 50% trains)

2	50%	Scrubber Column
2	50%	Scrubber Circulation Cooler
4	50%	Scrubber Top Circulation Pump
4	50%	Scrubber Bottom Circulation Pump
2	50%	Start up Steam Ejector
4	50%	Caustic Dosing Pump

Unit 1700 - Sour Slurry Stripper (1 x 100% train)

1	100%	Sour Slurry Stripper (SSS) column
1	100%	SSS Feed Vessel
3	50%	SSS Effluent Cooler
2	100%	SSS Feed Pump
2	100%	SSS Effluent Pump
2	100%	Acid Dosing Pump
1	100%	Drains Collection Vessel
2	100%	Drain Pump
1	100%	SSS Effluent Clarifier
2	100%	SSS Effluent Clarifier Bottom Pump
2	100%	SSS Effluent Clarifier Overflow Pump
1	100%	Sludge Storage Tank
2	100%	Sludge Storage Tank Bottom Pump
1	100%	Vacuum Belt Filter
2	100%	Filtrate Recycle Pump
1	100%	Filtrate Vacuum Pump

Unit 1800 - Sour Water Stripper (1 x 100% train)

2	100%	Feed/Effluent Heat Exchanger
1	100%	Sour Water Stripper
1	100%	SWS Overhead Condenser
1	100%	SWS Reflux Vessel
2	100%	Reflux SWS Pump
1	100%	SWS Reboiler
2	100%	SWS Effluent Pump
1	100%	SWS Effluent Cooler

7.0 REFERENCES

The following Table 5 "Overview of reference SCGP Projects" summarizes the status and operating data of all the plants adopting the Shell Coal Gasification Process, i.e. the pilot plants (Amsterdam and Hamburg), the demonstration plant (SCGP – Germany, Houston (USA), the operating plant (Demkolec, Buggenum (the Netherlands)) and the plants under design/engineering/development which Shell are allowed to refer to.

TABLE 5

Shell gasification reference list

Owner	Location	Feedstock (t/d)	Final Product	Start-up date
Shell/Koppers	Harburg, Germany	70	Syngas	1980
Shell	Houston, USA	200	Syngas	1985
NUON Power	Buggenum, The Netherlands	2000	Power	1994
Shuanghuan Chem.	Yingcheng, Hubei	900	Ammonia	2006
Sinopec/Shell	Dongting, Hunan	2000	Ammonia	2006
Sinopec	Zhijiang, Hubei	2000	Ammonia	2006
Sinopec	Anqing, Anhui	2000	Ammonia	2006
Liuhua Chem.	Liuzhou, Guanxi	1200	Ammonia	2007
Dahua Chem.	Dalian, Liaoning	1100	Methanol	2007 projected
Yuntianhua	Anning, Yunnan	2700	Ammonia	2007 projected
Yunzhanhua	Huashan, Yunnan	2700	Ammonia	2007 projected
Shenhua	Majiata, Inner Mongolia	2x2250	Hydrogen	2007 projected
Yongcheng Chem	Yongcheng, Henan	2150	Methanol	2007 projected

Zhongyuan Dahua	Puyang, Henan	2100	Methanol	2007 projected
Kaixiang	Yima, Henan	1100	Methanol	2007 projected
Datang	Duolun, Inner Mongolia	3x4000	Methanol	2009
Tianjin Soda Plant	Tianjin	2*2050	Ammonia/Methanol	2010
Guizhou Tianfu	Fuquan, Guizhou	2050	Ammonia/DME	2010
Magnum	Eemshaven, NL	3x2000 coal biom.	Power	2010

8.0 INVESTMENT COSTS

Table 6 summarizes the estimated total FOB costs estimated by FWI for the Gasification Island, as defined in para 2.0 for the two cases, based on 2007 costs in the Netherlands. Excluded are Coal Yard and Handling/Conveying facilities and general facilities (i.e. building, control room, DCS utilities etc.).

TABLE 6

	Case 1 MM Euro	Case 2 MM Euro
Direct Materials	129.8	137.4
Construction	58.7	62.1
Total	188.5	199.5

1.2 GEE Technology

The purpose of the attached document "GEE Gasification Island" formerly the Texaco, is to summarize the information used for the Hydrogen and Electricity Co-production study. In particular these data were the basis in the first step of the study for the selection of the gasification technology for the IGCC (section D.1).

Technical data of the IGCC have been taken from a previous study made by FWI for IEA GHG (Gasification Power generation study (2003)) and have been reviewed with GEE. Investment data have been updated to 2007 by FWI Estimate Department and finally approved by GEE.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : GASIFICATION POWER GENERATION STUDY
CONTRACT NO. : 1-BD-0337A
UNIT NO. : 1000
DOCUMENT NAME : GEE GASIFICATION ISLAND

ISSUED BY : L.VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

GEE GASIFICATION ISLAND

I N D E X

1.0 Introduction

The purpose of this chapter is to summarize the information received from GEE for the Gasification Power Generation Study that GEE allows to be disclosed to IEA GHG R&D without a non-disclosure agreement between IEA and GEE. They are the basis for the selection of the gasification technology for the IGCC configurations considered in the hydrogen and electricity coproduction study.

2.0 Gasification Island Process Description And Block Flow Diagram

2.1 Overall GEE Gasification Process Description

The Gasification Unit employs the GEE Gasification Process (TGP) formerly the Texaco Gasification Process (TGP), to convert feedstock coal into syngas. Facilities are included for scrubbing particulates from the syngas as well as removing the coarse and fine slag from the quench and scrubbing water.

The Gasification Unit includes the following sections, which are described briefly hereinafter:

Section	Description
1	Coal Grinding/Slurry Preparation
2	Gasification
3	Slag Handling
4	Black Water Flash
5	Black Water Filtration

The following description refers to a single train.

2.1.1 Coal Grinding/Slurry Preparation (PFD-01)

The Coal Grinding & Slurry Preparation System provides a means to prepare the coal as a slurry feed for the gasifier. Coal is continuously fed to the Coal Weigh Feeder, which regulates and weighs the coal fed to the Grinding Mill. Grey water from Black Water Filtration is used for slurrying the coal feed. Slurrying water is added to the grinding mill with a feed ratio controller to control the desired slurry concentration. The Grinding Mill may also utilize coal dust recovered by dust collection systems in the coal storage areas. The Grinding Mill is either a rod-type or ball-type with an overflow discharge. The Grinding Mill reduces the feed coal to the design particle size distribution.

Slurry discharged from the Grinding Mill passes through a coarse screen and into the Mill Discharge Tank, and is then pumped into the Slurry Run Tank. The Slurry Run Tank holds enough capacity to sustain full rate operation of the gasifier train during routine maintenance of the Grinding Mill. Coal slurry is pumped from the Slurry Run Tank to the Gasifier by the Slurry Charge Pumps, which are high pressure metering pumps. These pumps supply a steady, controlled flow of slurry to the Gasifier Feed Injector.

A below grade Grinding Area Sump is located centrally within the Coal Grinding and Slurry Preparation section to allow for handling of drains and spills in this area.

2.1.2 Gasification (PFD-02)

The Gasifier is a refractory-lined vessel capable of withstanding high temperatures and pressures. The coal slurry from the Slurry Run Tank and oxygen from the Air Separation Plant react in the gasifier at very high temperatures (approximately 1400 ^{o}C) and under conditions of insufficient oxygen to produce syngas. Syngas consists primarily of hydrogen and carbon monoxide with lesser amounts of water vapor, carbon dioxide, hydrogen sulfide, methane, and nitrogen. Traces of carbonyl sulfide (COS) and ammonia are also formed. Ash, which was present in the coal, melts in the gasifier and transforms into slag.

Hot syngas and molten slag from the Gasifier flow downward into a water filled quench chamber, where the syngas is cooled and the slag solidifies. Raw syngas then flows to the Syngas Scrubber for removal of entrained solids. The solidified slag flows to the bottom of the quench chamber, where the Slag Crusher is located. The coarse fraction of the slag is then removed from the quench section through a water-filled lockhopper system, after being ground through the Slag Crusher.

The Feed Injector is protected from the high temperatures prevailing in the gasifier by cooling coils through which cooling water is continuously circulated. Feed injector cooling water is stored in the Feed Injector Cooling Water Drum and pumped by the Feed Injector Cooling Water Pump to the Feed Injector Cooling Water Cooler and then to the feed injector cooling coils. After the cooling water exits the cooling coils, it flows to the Feed Injector Cooling Water Drum by gravity.

Syngas from the Gasifier quench chamber is fed to a Nozzle Scrubber. In the Nozzle Scrubber, the syngas is mixed with a portion of the Syngas Scrubber bottoms in order to wet the entrained solids so they can be removed in the Syngas Scrubber. The spray water is supplied by the Syngas Scrubber Circulating Pump.

The water/syngas mixture enters the Syngas Scrubber, where all of the solids are removed from syngas. Process condensate from the Syngas Treatment and Conditioning Line is fed into the Syngas Scrubber to remove particulates in the syngas. Then, the syngas from the overhead of the Syngas Scrubber is routed to the Syngas Treatment and Conditioning Line.

The Syngas Scrubber bottoms stream contains all the solids, which were not removed in the Gasifier quench chamber. In order to reduce the amount of solids recycled to the Nozzle Scrubber and Gasifier quench ring, a portion of the scrubber bottoms stream is sent to the Black Water Flash Section.

2.1.3 Slag Handling (PFD-03)

The Slag Handling System removes the majority of solids from the gasification process equipment. These solids are made up from the coal ash and unconverted coal components that exit the gasifier in the solid phase.

Coarse slag and some of the fine solids flow by gravity from the Gasifier quench chamber into the Lockhopper. Flow into the Lockhopper is assisted by the Lockhopper Circulation Pump which takes water from the top of the Lockhopper and returns it to the Gasifier quench chamber. After the solids enter the Lockhopper, the particles settle to the bottom. Thus, the Lockhopper acts as a clarifier, separating solids from the water. Solids are collected in this manner for a set period of time, typically about 30 minutes.

When the solids collection time is over, the Lockhopper is isolated from the quench chamber and depressured. Then, the solids, which have accumulated in the Lockhopper, are flushed with water into the Slag Sump. The water flush is then discontinued and the Lockhopper is filled with water and repressured, and the next solids collection period begins.

In the Slag Sump, slag settles onto a submerged conveyor, which drags the slag out of the water. It is passed over a screen, which allows surface water to drain. The slag is then transported by trucks to offsite for disposal. The water removed from the slag is pumped by the Slag Sump Overflow Pump to the Vacuum Flash Drum in the Black Water Flash Section.

Water used to flush the Lockhopper of collected solids is supplied to the Lockhopper Flush Drum from the Grey Water Tank in the Black Water Filtration Section. The water is cooled in the Lockhopper Flush Water Cooler so that the water in the Lockhopper will be cool at the start of the solids collection period and not get excessively hot during the solids collection period.

2.1.4 Black Water Flash (PFD-04)

The purpose of the Black Water Flash Section is to recover heat from the black water, as well as to remove dissolved syngas. Gas evolved from the flashes is routed to the Sulfur Recovery Unit, since it contains traces of hydrogen sulfide and ammonia. The cooled and flashed black water is sent to Black Water Filtration.

Black Water from the Gasifier quench chamber and the Syngas Scrubber is first routed to the LP Flash Drum. The overhead vapor is first used to heat the grey water return from the Black Water Filtration Section before it is condensed by the LP Flash

Condenser. Then, both of the vapor and condensate are routed to the Vacuum Pump Knockout Drum. From the LP Flash Drum, the black water stream goes to the Vacuum Flash Drum along with the black water from the Overflow Slag Sump. The Vacuum Flash Drum flashes out additional dissolve gases and liquid of which most of the liquid is condensed by the Vacuum Flash OH Condenser and separated in the Vacuum KO Drum. Then, both the vapor and condensate are routed to the Vacuum Pump Knockout Drum. Most of entrained gas in the black water is removed in the Vacuum Pump Knockout Drum and flows to the Sulfur Recovery Unit. Any liquid condensed in this vapor stream is also removed in Vacuum Pump Knockout Drum and flows to the Grey Water Tank.

2.1.5 Black Water Filtration (PFD-05)

The Black Water Filtration Section processes flashed black water from the Black Water Flash Section. The flashed black water from the Vacuum Flash Drum is sent to the LP Settler, where the suspended solids are settled at the bottom of the tank. The solids-free overflow is sent back to the Grey Water Tank, and the underflow is pumped by the LP Settler Bottom Pump to the Rotary Filter. The solids are removed, and the filtrate is sent to the Grey Water Tank. The filter cake is removed for disposal.

The water in the Grey Water Tank is essentially free of particulates. Some portion of the grey water is pumped by the LP Grey Water Return Pump to the Lockhopper Flush Drum, to the Coal Grinding Section and to offsite. The HP Grey Water Return Pump pumps grey water to the Grey Water Heater and then to the Syngas Scrubber.

3.0 **Process Flow Diagrams**

The simplified Process Flow Diagrams provided by GEE are attached.

IEA GHG

Hydrogen and Electricity Co-Production

Revision no.: Draft
Date: March 2007
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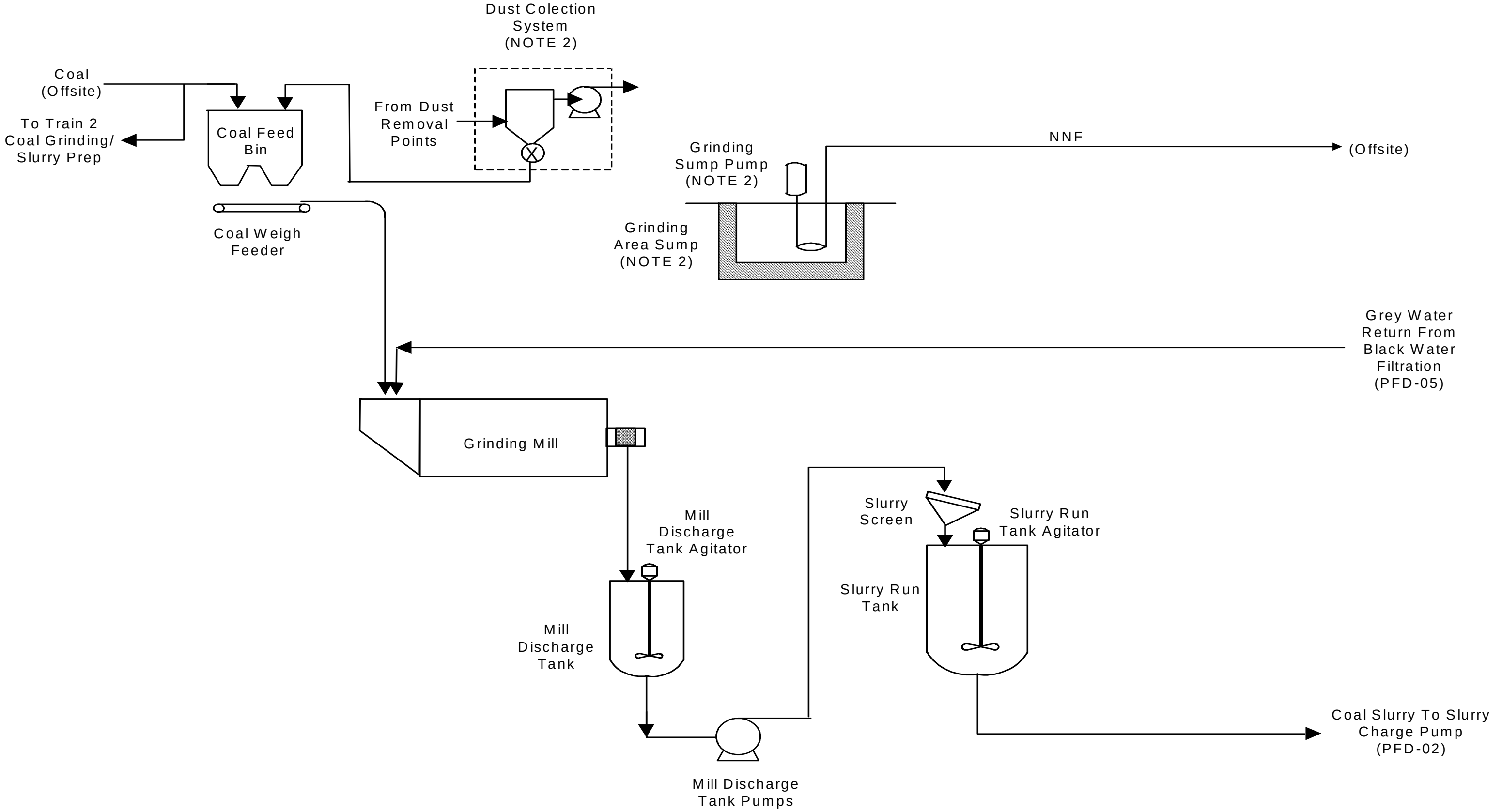


TEXACO DEVELOPMENT CORPORATION
HOUSTON, TEXAS

NOTES:

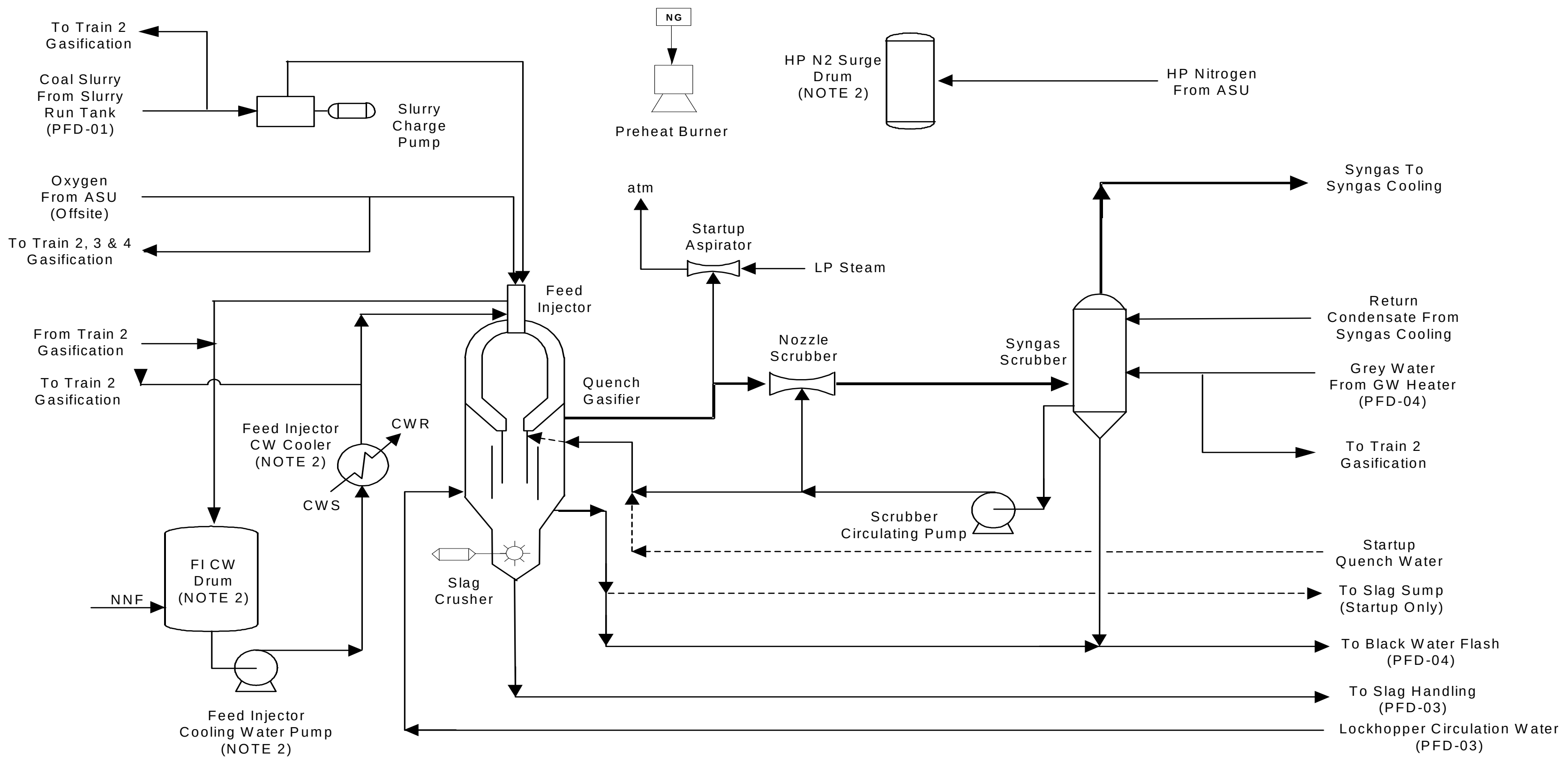
1. THE EQUIPMENT SHOWN IS FOR ONE TRAIN. TWO TRAINS AT 66.6% CAPACITY ARE REQUIRED UNLESS NOTED OTHERWISE.

2. THE GRINDING SUMP AND DUST COLLECTION SYSTEM ARE COMMON TO ALL TRAINS.

TEXACO GASIFICATION POWER SYSTEMS

COAL GRINDING/SLURRY PREP SECTION

DRAWN	DATE	PROJECT NO.		REV.
DESIGNER	DATE			
CHECKED	DATE	DRAWING NO.	PFD-01	
APPROVED	DATE			



TEXACO DEVELOPMENT
CORPORATION
HOUSTON, TEXAS

NOTES:

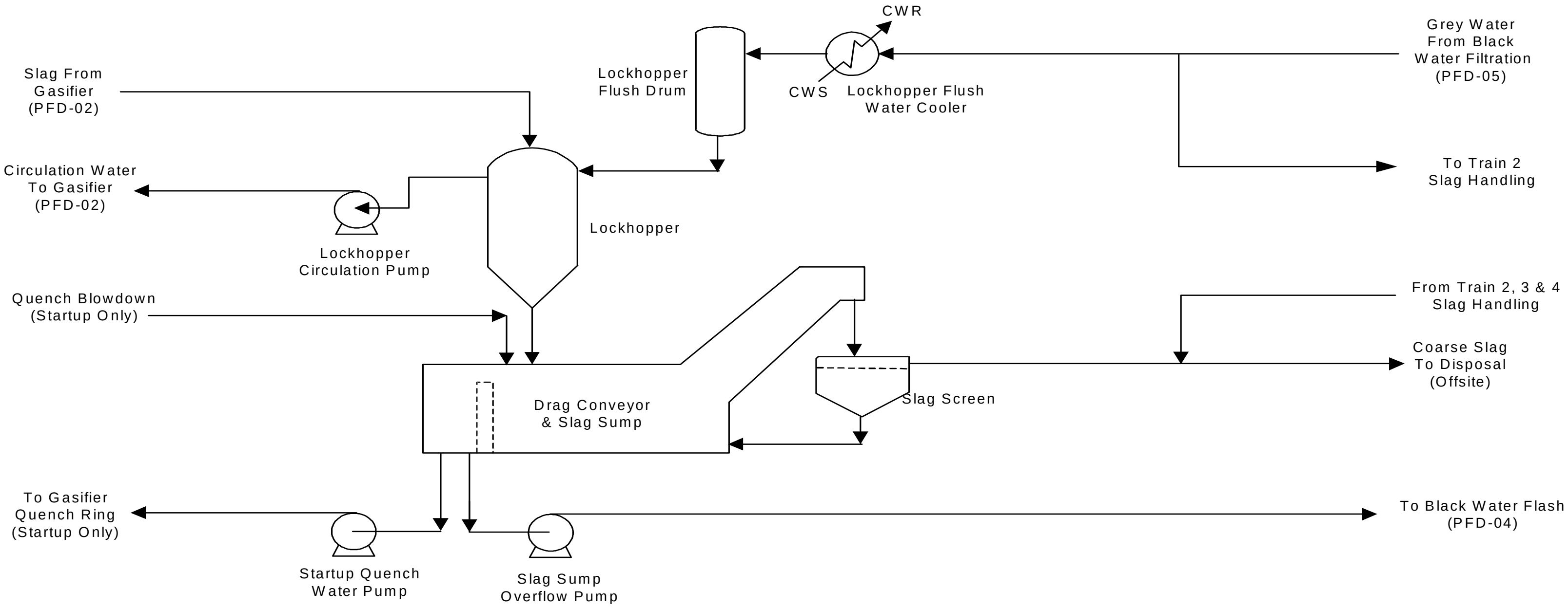
1. THE EQUIPMENT SHOWN IS FOR ONE TRAIN. FOUR TRAINS AT 33.3% CAPACITY ARE REQUIRED UNLESS NOTED OTHERWISE.

2. TWO TRAINS AT 66.6% CAPACITY ARE REQUIRED FOR HP N2 SURGE DRUM, FI CW DRUM, FI CW PUMP, AND FI CW COOLER.

TEXACO GASIFICATION POWER SYSTEMS

GASIFICATION SECTION

DRAWN	DATE	PROJECT NO.		REV.
DESIGNER	DATE			
CHECKED	DATE	DRAWING NO.	PFD-02	
APPROVED	DATE			



TEXACO DEVELOPMENT CORPORATION
HOUSTON, TEXAS

NOTES:

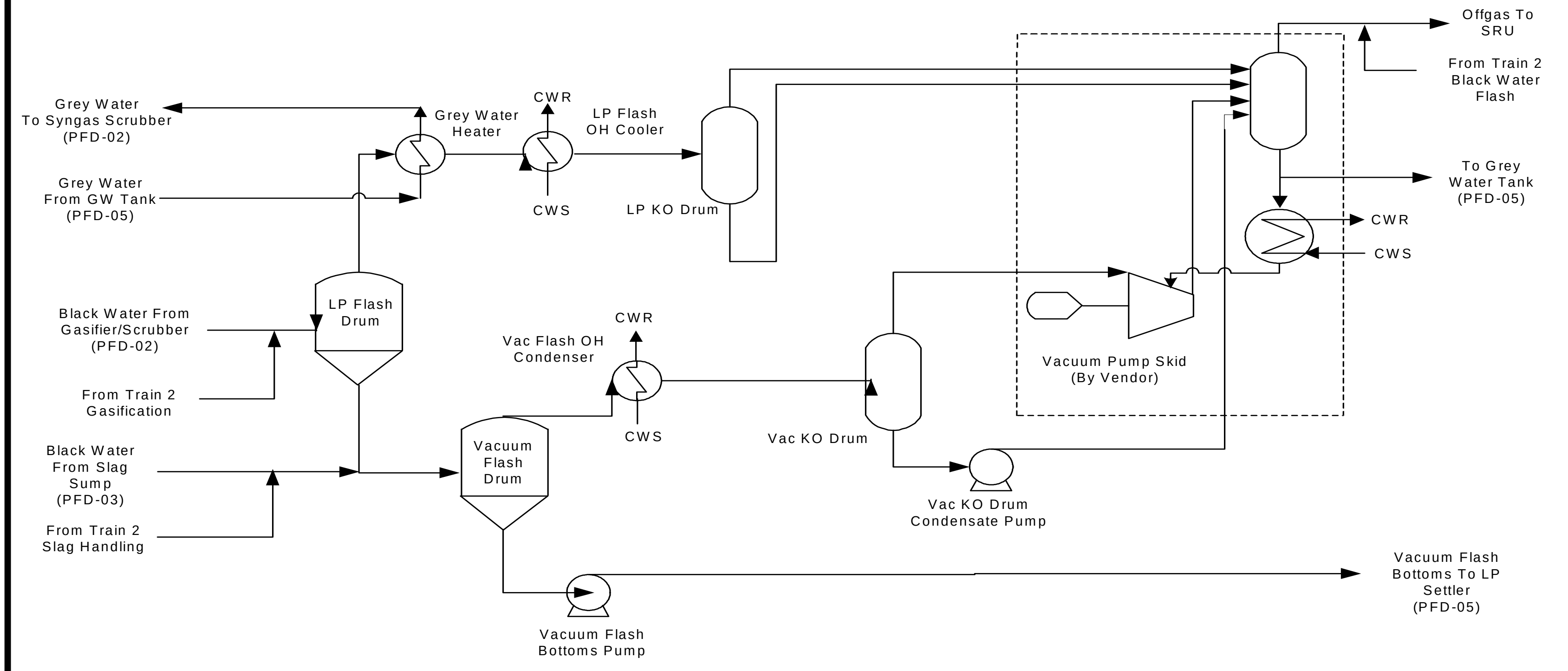
1. THE EQUIPMENT SHOWN IS FOR ONE TRAIN. FOUR TRAINS AT 33.3% CAPACITY ARE REQUIRED.

TEXACO GASIFICATION POWER SYSTEMS				
SLAG HANDLING SYSTEM				
DRAWN	DATE	PROJECT NO.		REV.
DESIGNER	DATE			
CHECKED	DATE	DRAWING NO.	PFD-03	
APPROVED	DATE			

IEA GHG

Hydrogen and Electricity Co-Production

Revision no.: Draft
Date: March 2007
Sheet: 12 of 21



TEXACO DEVELOPMENT CORPORATION
HOUSTON, TEXAS

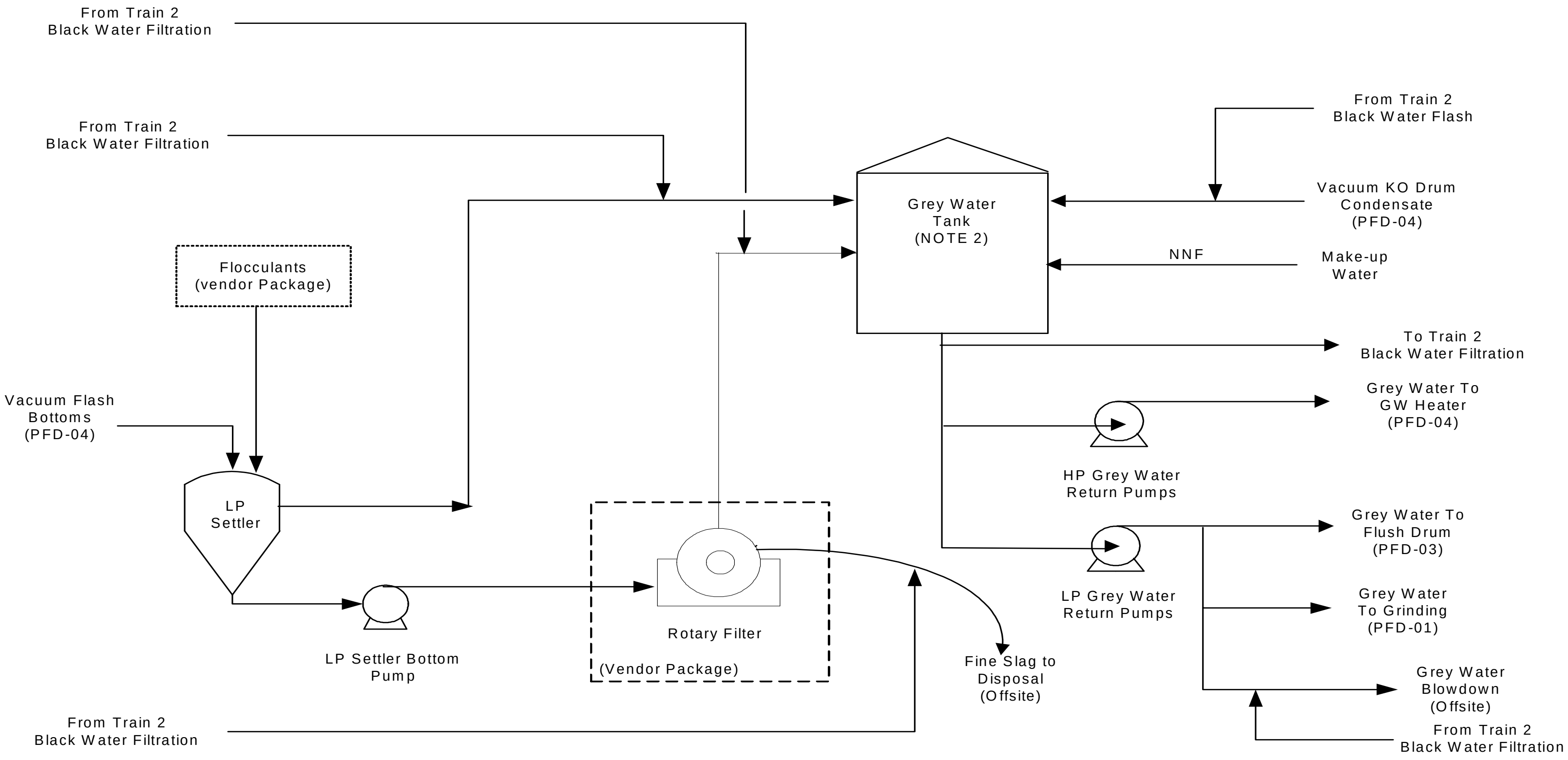
NOTES:

1. THE EQUIPMENT SHOWN IS FOR ONE TRAIN. TWO TRAINS AT 66.6% CAPACITY ARE REQUIRED UNLESS NOTED OTHERWISE.

TEXACO GASIFICATION POWER SYSTEMS

BLACK WATER FLASH SECTION

DRAWN	DATE	PROJECT NO.	REV.
DESIGNER	DATE		
CHECKED	DATE	DRAWING NO. PFD-04	
APPROVED	DATE		



TEXACO DEVELOPMENT CORPORATION
HOUSTON, TEXAS

NOTES:

1. THE EQUIPMENT SHOWN IS FOR ONE TRAIN. TWO TRAINS AT 66% CAPACITY ARE REQUIRED UNLESS NOTED OTHERWISE.

2. THE GREY WATER TANK IS COMMON TO ALL TRAINS.

TEXACO GASIFICATION POWER SYSTEMS

BLACK WATER FILTRATION SECTION

DRAWN	DATE	PROJECT NO.		REV.
DESIGNER	DATE			
CHECKED	DATE	DRAWING NO.	PFD-05	
APPROVED	DATE			

4.0 Characteristics of Streams at Gasification Island Battery Limits.

The following Tables summarize the characteristics of Streams at Gasification Island Battery Limits for the considered case of high Gasification pressure with CO_2 capture

TABLE 1

OVERALL PERFORMANCE

Fresh Coal to Coal Grinding	
Flowrate (fresh, Air Dried Basis), t/h	323.1
Ultimate Analysis (%wt) (Dry, ash free)	
Carbon	82.5
Hydrogen	5.6
Nitrogen	1.77
Sulphur	1.1
Oxygen	9.0
Ash	0.03
Total	100.0
Coal LHV (Air Dried Basis), kcal/kg	6180
Total Thermal Power (LHV), MWt	2321.8
Oxygen Conditions	
95% Oxygen Flowrate, t/h	278.7
Oxygen Pressure @ B.L., bar g	79
Oxygen Temperature @ B.L., °C	149
Gasification Conditions	
Pressure, bar g	65
Temperature, °C	~ 1400

TABLE 1 (c'd)

Characteristics of Syngas Ex Scrubber (Total)	
Composition, % mol	
CO	15.6
H_2	15.1
CO_2	7.3
H_2O	61.0
Ar + N_2	0.8
H_2S + COS	0.12
Others	0.08
Flowrate, kmol/h	72,260
Pressure @ B.L., bar g	62
Temperature @ B.L., °C	243
Raw Syngas LHV, kcal/kg	1015
Gasification Efficiency (LHV), %	70.5

TABLE 2

Coarse Slag	
Water, % wt	50
Total Wet, kg/h	76,300
Filter Cake	
Water, % wt	70
Total Wet, kg/h	31800

5.0 Utility Consumptions

Table 3 summarizes the utility continuous consumptions estimated for the two cases.

TABLE 3

HP Steam, t/h	5
MP Steam, t/h	0
LP Steam, t/h	0
Fresh Cooling Water, m^3/h	3100
Absorbed Electric Power, kW	13900

6.0 Equipment list

Only major equipment in TGP's Battery Limit are presented.

Coal Handling/Slurry Preparation

Coal Weigh Feeder	2 x	66%
Coal Feed Bin	2 x	66%
Dust Collection System	1 x	100%
Grinding Area Sump	1 x	100%
Grinding Sump Pump	1 x	100%
Grinding Mill	2 x	66%
Mill Disch Tank Agitator	2 x	66%
Mill Discharge Tank	2 x	66%
Mill Discharge Tank Pump	2 x	66%
Slurry Screen	2 x	66%
Slurry Run Tank Agitator	2 x	66%
Slurry Run Tank	2 x	66%

Gasification

Slurry Charge Pump	4 x	33%
Feed Injector CW Drum	2 x	66%
Feed Injector CW Cooler	2 x	66%
Feed Injector CW Pump	2 x	66%
Feed Injectors	9	Total
Preheat Burner	4	Total
Quench-type Gasifier	4 x	33%
Gasifier – Refractory	4	Total
Slag Crusher	4 x	33%
Syngas Scrubber	4 x	33%
Nozzle Scrubber	4 x	33%
Scrubber Circulation Pump	4 x	33%
HP Nitrogen Surge Drum	2 x	66%
Safety System PLC	1	
Start-Up Aspirator	4 x	33%

Slag Handling

Lockhopper	4 x	33%
Lockhopper Circ Pump	4 x	33%
Lockhopper Flush Drum	4 x	33%
Lockhopper Flush Water Cooler	4 x	33%
Start Up Quench Water Pump	4 x	33%
Drag Conveyor/Slag Sump	4 x	33%
Slag Screen	4 x	33%
Slag Sump Overflow Pump	4 x	33%

Black Water Flash

Grey Water Heater	2 x	66%
LP Flash OH Cooler	2 x	66%
LP Knockout Drum	2 x	66%
LP Flash Drum	2 x	66%
Vacuum Flash Drum	2 x	66%
Vacuum Flash OH Condenser	2 x	66%
Vacuum KO Drum	2 x	66%
Vacuum KO Drum Condensate Pump	2 x	66%
Vacuum Flash Bottoms Pump	2 x	66%
Vacuum Pump Skid	2 x	66%

Black Water Filtration

LP Settler	2 x	66%
LP Settler Bottoms Pump	2 x	66%
Rotary Filter	2 x	66%
Grey Water Tank	1 x	100%
HP Grey Water Return Pump	2 x	66%
LP Grey Water Return Pump	2 x	66%

7.0 References

As of January 2001 the total plants licensed by Texaco are 127, with a total of 69 plants in operation and engineering, construction or start-up phases.
Table 4 shows the split among different feedstocks.

TABLE 4

Feedstock	Plants in operation	Plants in Eng./ Constr./Start-up Phases	Total
Coal/Petcoke	13	2	15
Liquid	20	12	32
Natural Gas	19	3	22
TOTAL	49	20	69

Table 5 lists coal gasification plants presently in operation.

TABLE 5

Texaco Coal/Petcoke Gasification Process

Customer	Location	No. of Gasifiers Op/spare	Type Quench (Q) WHB (FHR)	Solid Feedstock	Product	Start Date
Eastman Chemical	Kingsport, TN – USA	1/1	Q	Bituminous Coal	Oxochemicals	1983
Ube Ammonia Industry	Ube City – Japan	3/1	Q	Coal/Petcoke	Ammonia	1984
Rheinbraun	Ville – Germany	3/0	Q/FHR	Coal/oil	Methanol	1986
Lu Nan Chemical Industry	Tengxian, Shandong – China	2/0	Q	Bituminous Coal	Ammonia	1993
Shanghai Pacific Chemical	Wujing, Shanghai – China	3/1	Q	Anthracite Coal	Methanol/ Town gas	1995
Tampa Electric	Lakeland, FL – USA	1/0	FHR	Coal	Electricity	1996
GEE Gasification Power Systems	El Dorado, KS – USA	1/0	Q	Petcoke	Electricity/ Steam	2000
Weihe Fertilizer	Xian, Shaanxi – China	2/1	Q	Coal	Acetic Acid	1996
Farmland Industries	Coffeyville, KS – USA	1/0	Q	Petcoke	Ammonia/ UAN	2000
Huainan	Anhui – China	2/1	Q	Coal	Ammonia	2000
Motiva Enterprises	Delaware City, DE – USA	2/0	Q	Petcoke	Electricity/ Steam	2000

8.0 Investment costs

Table 6 summarizes the estimated Investment Cost provided by Texaco for the Gasification Island for the two cases, split into the main sections and escalated by FWI to 2007. This cost includes materials and construction only.

TABLE 6

	MM Euro
Direct Materials	184.6
Construction	62.8
Total	247.4

1.3 Siemens Technology

The purpose of the attached document "Siemens Gasification Island" is to summarize the information used for the Hydrogen and Electricity Co-production study. In particular these data were the basis in the first step of the study for the selection of the gasification technology for the IGCC (section D.3).

Technical and economical data of the IGCC have been taken from FWI in house data relevant to previous projects and have been reviewed with Siemens.

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME	:	GASIFICATION POWER GENERATION STUDY
CONTRACT NO.	:	1-BD-0337A
UNIT NO.	:	1000
DOCUMENT NAME	:	SIEMENS GASIFICATION ISLAND

ISSUED BY	:	L.VALOTA
CHECKED BY	:	P. COTONE
APPROVED BY	:	S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

SIEMENS GASIFICATION ISLAND

I N D E X

1.0 INTRODUCTION

The purpose of this document is to summarize the information received from Siemens for the first step of the Gasification Power Generation Study. They are the basis for the selection of the gasification technology for the IGCC configurations considered in the hydrogen and electricity coproducion study.

2.0 GASIFICATION ISLAND PROCESS DESCRIPTION AND BLOCK FLOW DIAGRAM

2.1 General description of the Siemens Coal Gasification Process

The Siemens gasifier vessel is a cooling screen design gasifier consisting of an outside pressure wall and an inside cooling screen cooled by pressurized water.

The feed system is pneumatic (high density-low velocity). Pulverized coal is pressurized and transported pneumatically to the gasifier.

The dry feed minimizes the O_2 requirement and makes the gasifier more efficient than entrained flow gasifiers using wet feed systems. A penalty is however paid because the dry feed is more costly and operationally more complex.

The raw gas leaving the gasifier at high temperature contains molten ash and a small quantity of unburned carbon (soot). This stream is directly quenched in water to cool the gas and remove solidified particles, prior to water scrubbing.

The major advantage of the quench variant is a lower cost and higher reliability. The quench provides in the syngas all the water needed by the downstream shift reaction.

The gasification unit includes the following sections, which are described briefly hereinafter:

- Dense flow feeding
- Gasifier
- Quench
- Slag discharge system
- Gas scrubbing
- Waste water treatment

2.2 Process Description

Reference is made to the attached Block Flow Diagram.

Feeding system

The coal feeding system consists of one coal silo, mills and conveyor system and dosing unit for each gasifier.
The mills reduce the coal size to a fine powder.

By means of conveyor systems the pulverized coal is passed to a dense-flow feeding system consisting of a sequence of an atmospheric fuel bunker, lock hoppers and a feeder vessel.
The pulverized fuel settles in the fuel bunker and the carrier gas and purging gas are vented over the bunker top. The full lock hopper is pressurized with purge gas.
The fuel in the feeder vessel is partially fluidised by means of a carrier gas (N_2 or CO_2) in the vortex shaft of the feeder vessel, in which the fuel conveying lines are immersed. Finally the fuel is pneumatically transported in a dense flow to the gasifier burners.

Gasifier

The feedstock is gasified in a patented cooling screen design gasifier. This design lowers the risk of slag attack to a refractory lining and offers long lifetime and low maintenance cost operation. For safe capture of slag and solids a full-quench system is proposed.
The gasifier consists of an outside pressure wall and an inside cooling screen cooled by pressurized water to protect the outside wall against chemical and thermal attacks.

The reactants, pulverized fuel and oxygen are fed into the reaction chamber in parallel flow through the combination burners at the gasifier top. The latter are converted in a heterogeneous flame reaction in entrained flow at temperatures exceeding slag melting temperatures.

At the top of the reactor a combined burner consisting of a pilot burner and a coal burner (main burner) is arranged. Each main burner is equipped with one feed line.

The partial oxidation reaction converts the coal into hydrogen and carbon monoxide. The inert components in the feed are forming a slag.

Quench

The hot raw synthesis gas and the liquid slag leave the gasifier reaction chamber and flow in parallel vertically downward and discharge directly into the quench section where the raw gas is cooled down by injection of water. Slag produced is granulated in the water bath in the bottom of the quench system.
The raw gas is saturated with steam. This water becomes gas condensate in the following cooling steps of the syngas treatment and it will be recycled back as quench water.
The water of the quench, which is not vaporized, is flashed together with suspended solids (slag, fine ash, coke, soot and salts) and sent to the waste water treatment.

Slag Handling

The slag discharged from the quench sump falls into a water-filled pressurized lock hopper. When the lock hopper is filled with slag, it is cooled, depressurised and the slag and any water remaining in the hopper are discharged into a slag-receiving tank.
The major portion of the slag settles in the slag-receiving tank from where it is discharged by means of a drag chain conveyor. The slag is then washed on a slag wash conveyor to remove fines and quench water and is passed to a conveyor that transports the slag to a slag storage bin/container.
Waters carried out of the slag discharge system are collected in a conveyor overflow wet well and pumped to the waste water treatment plant via a hydro cyclone. Water that is needed in the slag discharge system is recycled from the waste water treatment plant.

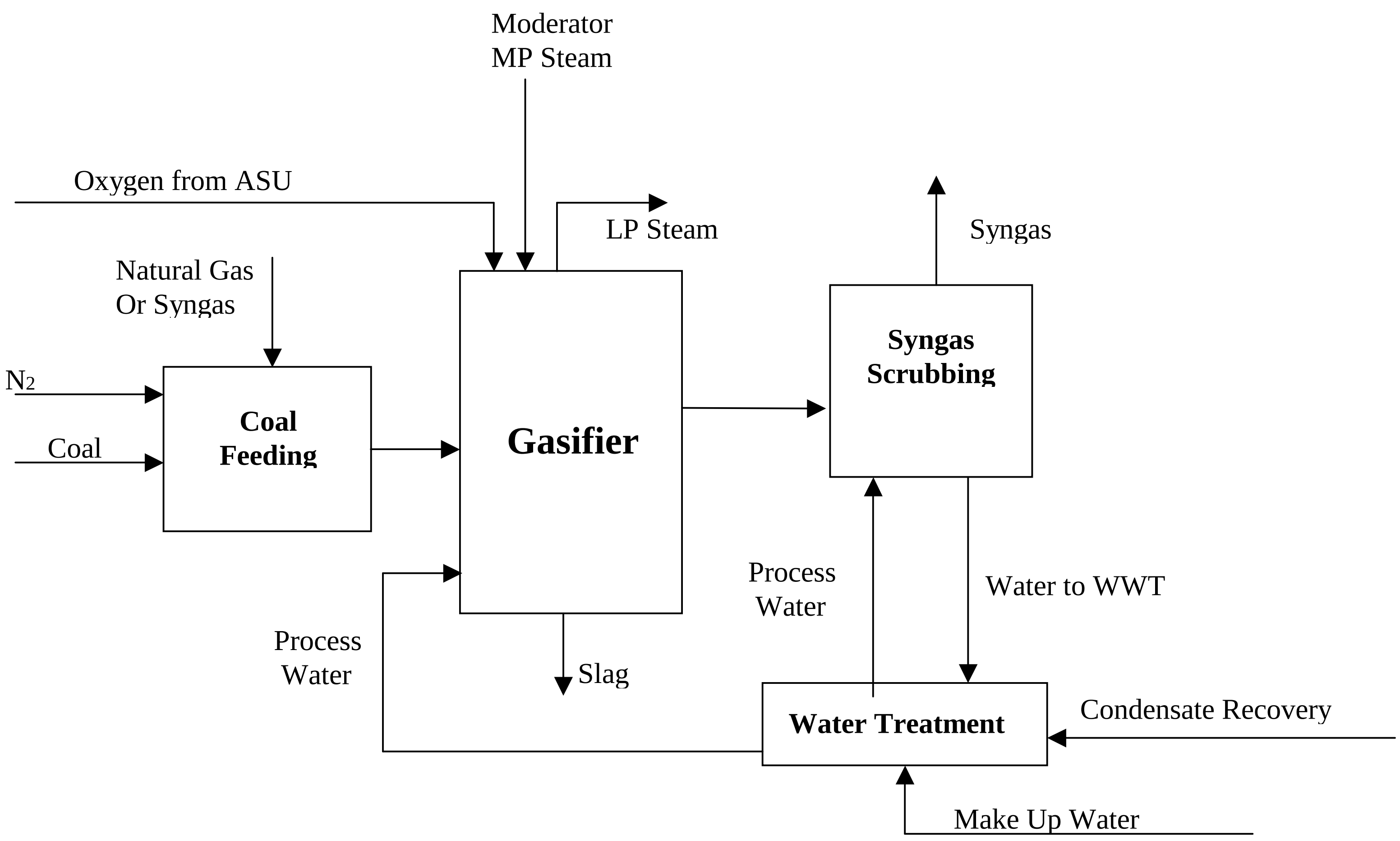
Gas Scrubbing

The wet raw gas from the quench is cleaned in a venturi scrubber, where fine ash and soot particles are removed from the raw gas by water. Scrubber water is directed to the Quench water vessel. Remaining solid particles in the raw syngas are separated from the gas in a double ventury system followed by a partial condenser with K/O-drum in order to minimize the dust content in syngas before sending to the Syngas Cooling and conditioning line.

Waste Water Treatment

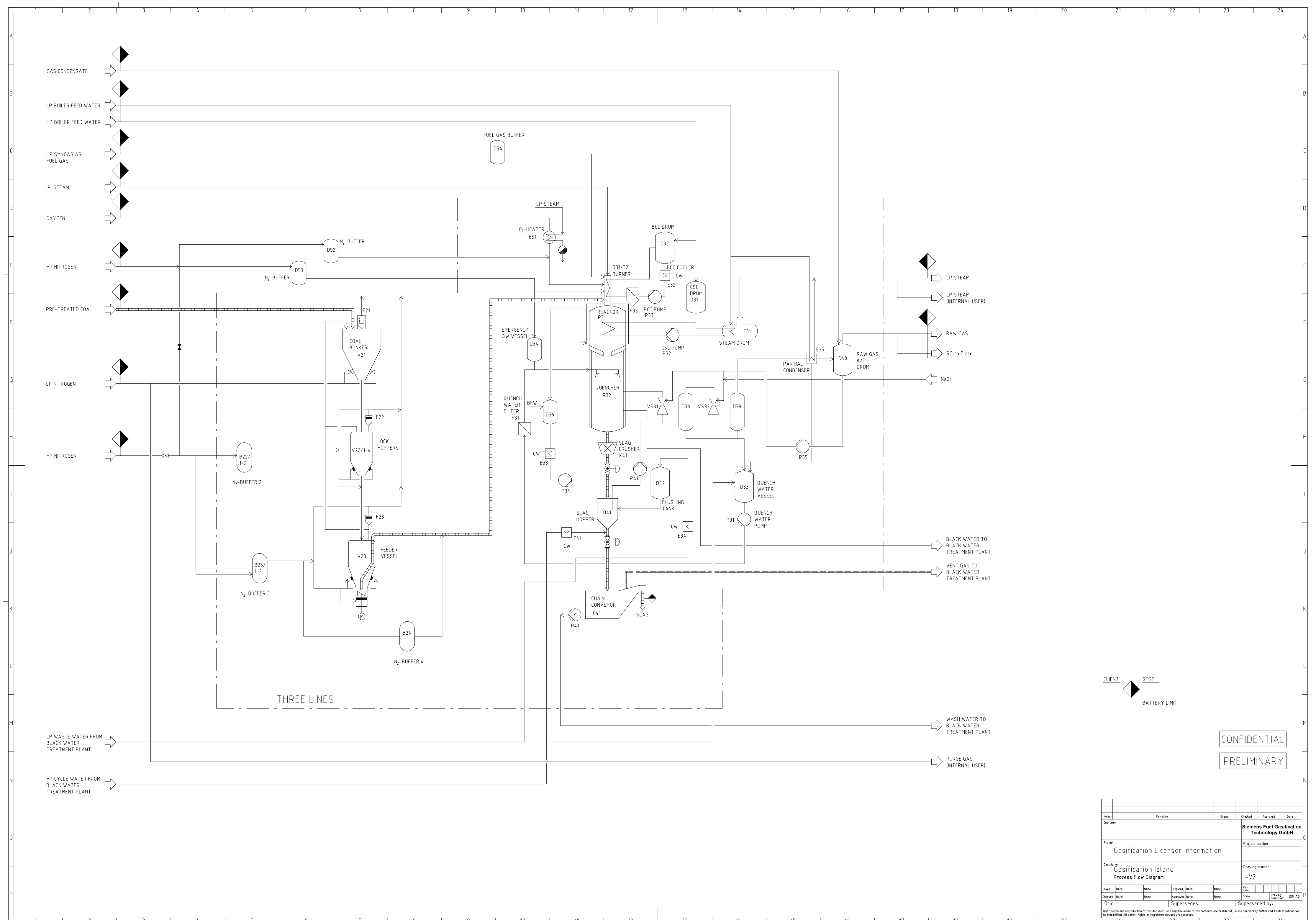
The liquid effluents from the quench systems and water from the slag separation contaminated with fine particulate matter, soot and salts are treated in this section. Waste water from the quench circuit is first depressurized in a thickener. Most of the pre-cleaned quench water is returned to the quench system.
The remaining part of the pressurized waste water is sent to a two step flash system followed by a thickener. Clean water is sent back to the gasifier as quench water and a small amount of water is discharged as waste water for later treatment.

SIEMENS GASIFICATION TECHNOLOGY



3.0 PROCESS FLOW DIAGRAMS

The preliminary Process Flow Diagrams provided by SIEMENS are attached.



GAS CONDENSATE
LP BOILER FEED WATER
HP BOILER FEED WATER
HP SYNGAS AS FUEL GAS
IP-STEAM
OXYGEN
HP NITROGEN
PRE-TREATED COAL
LP NITROGEN
HP NITROGEN
LP WASTE WATER FROM BLACK WATER TREATMENT PLANT
HP CYCLE WATER FROM BLACK WATER TREATMENT PLANT
FUEL GAS BUFFER
D54
LP STEAM
O_2-HEATER
E51
N_2-BUFFER
D52
D53
N_2-BUFFER
F21
COAL BUNKER
V21
F22
LOCK HOPPERS
V22/1-4
B22/1-2
N_2-BUFFER 2
F23
V23
FEEDER VESSEL
B23/1-2
N_2-BUFFER 3
M
B24
N_2-BUFFER 4
THREE LINES
BCC DRUM
D32
BCC COOLER
CW
E32
B31/32
BURNER
F33
BCC PUMP
P33
CSC DRUM
D31
REACTOR
R31
E31
STEAM DRUM
CSC PUMP
P32
EMERGENCY QW VESSEL
D34
QUENCHER
R32
QUENCH WATER FILTER
F31
BFW
D36
CW
E33
P34
SLAG CRUSHER
X41
P41
D42
FLUSHING TANK
SLAG HOPPER
D41
CW
E34
E41
CW
CHAIN CONVEYOR
C41
SLAG
P41
VS31
D38
VS32
D39
E35
PARTIAL CONDENSER
D40
RAW GAS K/O DRUM
P35
D33
QUENCH WATER VESSEL
P31
QUENCH WATER PUMP
LP STEAM
LP STEAM (INTERNAL USER)
RAW GAS
RG to Flare
NaOH
BLACK WATER TO BLACK WATER TREATMENT PLANT
VENT GAS TO BLACK WATER TREATMENT PLANT
WASH WATER TO BLACK WATER TREATMENT PLANT
PURGE GAS (INTERNAL USER)
CLIENT
SFGT
BATTERY LIMIT
CONFIDENTIAL
PRELIMINARY
Index
Revisions
Drawn
Checked
Approved
Date
Costumer
Siemens Fuel Gasification Technology GmbH
Project
Gasification Licensor Information
Project number
Description
Gasification Island
Process Flow Diagram
Drawing number
-V2
Drawn
Date:
Name:
Prepared
Date:
Name:
Rev. Index
-
Checked
Date:
Name:
Approved
Date:
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Scale
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Drawing dimension
DIN A0
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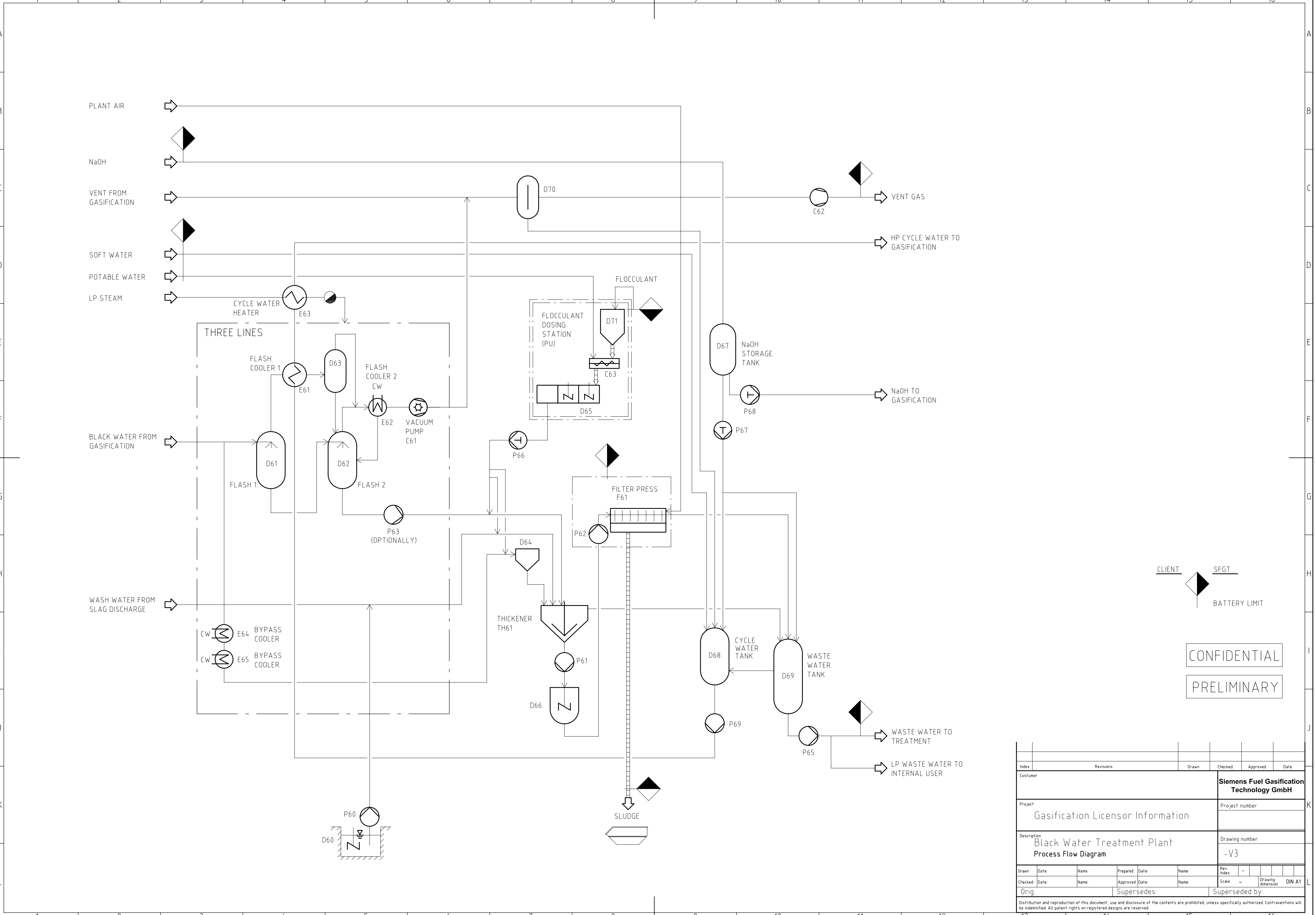

PLANT AIR
NaOH
VENT FROM GASIFICATION
SOFT WATER
POTABLE WATER
LP STEAM
BLACK WATER FROM GASIFICATION
WASH WATER FROM SLAG DISCHARGE
THREE LINES
CYCLE WATER HEATER
E63
FLASH COOLER 1
E61
D63
FLASH COOLER 2
CW
E62
VACUUM PUMP
C61
D61
FLASH 1
D62
FLASH 2
P63
(OPTIONALLY)
CW
E64
BYPASS COOLER
CW
E65
BYPASS COOLER
P60
D60
D70
C62
VENT GAS
HP CYCLE WATER TO GASIFICATION
FLOCCULANT
FLOCCULANT DOSING STATION (PU)
D71
C63
D65
P66
D64
THICKENER
TH61
P61
D66
FILTER PRESS
F61
P62
SLUDGE
D67
NaOH STORAGE TANK
P68
NaOH TO GASIFICATION
P67
D68
CYCLE WATER TANK
P69
D69
WASTE WATER TANK
P65
WASTE WATER TO TREATMENT
LP WASTE WATER TO INTERNAL USER
CLIENT
SFGT
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Project number
Description
Black Water Treatment Plant
Process Flow Diagram
Drawing number
-V3
Drawn
Date:
Name:
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Date:
Name:
Rev. Index
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4.0 CHARACTERISTICS OF STREAMS AT GASIFICATION ISLAND BATTERY LIMITS.

The following Tables summarize the characteristics of Streams at the Gasification Island Battery Limits for the considered case of high gasification pressure with CO_2 capture

TABLE 1

OVERALL PERFORMANCE

Fresh Coal to Coal Grinding	
Flowrate (fresh, Air Dried Basis), t/h	295.3
Flowrate (dryed coal, 2% H_2O), t/h	272.7
Ultimate Analysis (%wt) (dry, ash free)	
Carbon	82.5
Hydrogen	5.6
Nitrogen	1.77
Sulphur	1.1
Oxygen	9
Chlorine	0.03
Total	100.00
Coal HHV (Air Dried Basis), kcal/kg	6464
Coal LHV (A.D.B.), kcal/kg	6180
Thermal Pow, MWt (LHV)	2122

TABLE 1 (c'd)

Characteristics of Syngas Ex Scrubber (Total)	
Composition, % mol	
CO	29.2
H_2	11.3
CO_2	1.9
H_2O	54.0
Ar	0.4
N_2	3.0
H_2S	0.2
	100.00
Flowrate, kmol/h	53,870
t/h	1,075.6
Pressure @ B.L., bar g	36
Temperature @ B.L., °C	216
Raw Syngas LHV, dry kcal/kg	1327
Raw Syngas Thermal Power (LHV), MWt	1659.3
Gasification eff. (based on Air Dried coal LHV), %	78.2
Gasification eff. (based on Dried coal @ 2% H2O, LHV), %	77.6
Oxygen Consumptions	
O_2 Flowrate, t/h	233
O_2 Press @ B.L., barg	47
O_2 Temp @ B.L., °C	120
Nitrogen Consumptions	
HP N_2 Flowrate, t/h	72
HP N_2 Press @ B.L., barg	54
HP N_2 Temp @ B.L., °C	70
LP N_2 Flowrate, t/h	19
LP N_2 Press @ B.L., barg	6.5
LP N_2 Temp @ B.L., °C	15
Natural Gas Consumption (pilot)	
NG Flowrate, t/h	1.2
NG Pressure @ B.L., bar g	49
NG Temperature @ B.L., °C	15

TABLE 2

STEAM PRODUCTIONS/BFW CONSUMPTIONS

LP Steam Net Production	
Flowrate, t/h	28
Pressure @ Unit B.L, barg	6.5
Temperature, °C	sat
Steam Condensate	
Flowrate, t/h	65
Pressure @ Unit B.L, barg	5
Temperature, °C	150
LP BFW Consumption	
Flowrate, t/h	98
Pressure @ Unit B.L, barg	17
Temperature, °C	160
HP BFW Consumption	
Flowrate, t/h	19
Pressure @ Unit B.L., barg	40
Temperature, °C	sat

5.0 UTILITY AND CHEMICAL CONSUMPTIONS

Table 3 summarizes the utility continuous consumptions (other than steam and Nitrogen) estimated for the two cases.

TABLE 3

Fresh Cooling Water, m^3/h	1,300
Absorbed Electric Pow, kW	7500
Instrument Air, Nm^3/h	700

6.0 INVESTMENT COSTS

Table 4 summarizes the estimated total FOB costs estimated by FWI for the Gasification Island, for the two cases, based on 2007 costs in the Netherlands. Excluded are Coal Yard and Handling/Conveying facilities and general facilities (i.e. building, control room, DCS utilities etc.).

TABLE 4

	MM Euro
Direct Materials	86.4
Construction	31.9
Total	118.3

2.0 Coal Handling and Storage

Coal Handling and Storage consists of one dome with a coal storage capacity equivalent to approx. 21 days at IGCC full capacity, one conveyor connecting the pier with the dome sized for 1200 t/h, and one conveyor connecting the dome with the milling system in the Gasification Island sized for the actual coal flowrate.

In case of a Shell and Siemens gasification technology, a coal milling and drying section is also present. It includes a conventional mill, similar to those used in a pulverised coal boiler. The mill grinds the coal to a size range suitable for efficient gasification. As the coal is being ground, it is simultaneously dried utilising a heated inert gas stream. The gas stream carries the evaporated water from the system as it sweeps the pulverised coal through an internal classifier to collection in a bag house.
The heat required for drying the coal is supplied by burning Natural Gas (Siemens gasification technology) or syngas (Shell gasification technology).

The Unit is designed in order to minimize particulate emissions, with both closed storage (dome) and closed conveyors.

3.0 Air Separation Unit

The Air Separation Unit (ASU, Unit 2100) is installed to produce oxygen and nitrogen through cryogenic distillation of atmospheric air.
The oxygen produced is delivered to the Gasification Island to be used as reaction oxidant. A small quantity is also used by the Sulphur Recovery Unit.

As a by-product nitrogen is obtained:

- for GEE alternatives nitrogen is routed to the gas turbines of the combined cycle for power augmentation and NO_x control;
- for Shell and Siemens alternatives nitrogen is used for the pneumatic transport of dried pulverized coal to the gasifiers; the excess is routed to the gas turbines for power augmentation and NO_x control.

The plant consists of two air separation trains and at the same time is able to produce additional oxygen and nitrogen products to maintain the desired inventories in the storage systems of liquid and gaseous products used as back-up; these systems are common to both trains.

The ASU, for each different case, can be stand alone or partially integrated with the gas turbines with a certain percentage; it consists of the percentage of air required by the air separation that is supplied by the gas turbine. Integration means recovery of the waste energy available, improvement of the efficiency and reduction of investment cost, but also a possible reduction of operating flexibility that can affect the IGCC availability. Considerations regarding the integration have been made in order to optimize the configuration to reach the best overall IGCC performance. Considerations on IEA GHG Gasification Power generation study (2003) show an optimised integration for two 9FA of 30% for Shell technology with CO_2 capture, 50% for Shell technology without CO_2 capture and 50% for GEE (with CO_2 capture).
In the current study the same configuration has been considered: when the power island is based on only one Gas Turbine, only half of the integration has been considered.
Siemens technology presents the same value of integration as Shell technology.
In case G3 (maximum hydrogen production), the power island is based on one GT 6FA only. As the maximum flowrate that can be extracted from such gas turbine is much lower than from 9FA, no air integration has been considered.

The streams listed in Table C.3.1 are produced according to the requirement of each gasification technology.

Table C.3.1

	Product	Use	Details	Gasification Technology
1	Oxygen	C	High Pressure Gaseous Oxygen for Gasifiers	Sh/GEE/Si
2	Oxygen	C	Low Pressure Gaseous Oxygen for Sulphur Recovery Claus Units	Sh/GEE/Si
3	Nitrogen	C	Medium Pressure Gaseous Nitrogen for Syngas Dilution at Gas Turbines	Sh/GEE/Si
4	Nitrogen	C	Very High Purity High Pressure Gaseous Nitrogen for dried coal transport	Sh/Si
5	Nitrogen	C	Very High Purity Low Pressure Gaseous Nitrogen for died coal transports	Sh/Si
6	Nitrogen	C	Very High Purity Low Pressure Gaseous Nitrogen for blanketing, equipment purging, etc	Sh/GEE/Si
7	Nitrogen	D	Very High Purity High/Low Pressure Gaseous Nitrogen for Purging under Gasifiers and Gas Turbine Shutdown	Sh/GEE/Si
8	Air	C	Low Pressure Dry Gaseous Air to Plant and Instrument Air System	Sh/GEE/Si

Notes (1): Sh = Shell
GEE = GE Energy
Si= Siemens

(2) C = Continuous
D = Discontinuous

3.1 Capacity

The Air Separation Unit capacity is defined per each alternative by the required oxygen production (sum of flowrates to the gasification island and to the sulphur plant).

3.2 Compressed Air

When the gasification operates at full load, 15% (25% for GEE case; 50% in case G1; 30% in case G2) of the air required by the ASU to obtain the design oxygen production is derived from gas turbine compressor; the integration between the gas turbine operation and the ASU is achieved at a level where 85% (75% for GEE case; 50% in case G1; 70% in case G2) of the atmospheric air is compressed with selfstanding units and the difference comes already pressurized from the compressors of the gas turbines in the combined cycle.
The air extracted from the gas turbine at high temperature is cooled by exchanging heat with nitrogen for syngas dilution before being fed to the Air Separation Unit.

3.3 Product Characteristics

Oxygen For Gasifiers and Sulphur Plant

Purity

O_2	95 mol%
Ar	3.5 mol%
N_2	1.5 mol%
H_2O	1 ppm (max)
CO_2	1 ppm (max)

Nitrogen For Syngas Dilution at Gas Turbines

The gas turbines require a continuous gaseous nitrogen supply to dilute Syngas and maximise power output. The maximum oxygen content of nitrogen stream is 2% mol.

Other Nitrogen Streams

Purity	
N_2	99.99 mol% (1)
Cl_2	Absent
Ar	300 ppm (max)
CO_2	5 ppm (max)
HC	5 ppm (max)
Oxygenated Compounds	100 ppm (max)
Dew Point	-50 °C @ 7 barg
CO (No. of times the content in ambient air)	1.5 max

Note (1): including Argon

These streams perform the following functions:

a. Nitrogen for Pneumatic Transport of dried coal

b. Nitrogen for Blanketing and Purging

   The IGCC plant requires a continuous supply of gaseous nitrogen for tank blanketing and other small purging.

c. Nitrogen for Purging Under Gasifier and Gas Turbine Shutdown

   The instantaneous shutdown of one gasifier or of one gas turbine requires a purging supply of gaseous nitrogen. To ensure a secure supply to the gasifiers, as well as the two gas turbines requires a dedicated high pressure local storage of gaseous nitrogen, to be fed by the ASU. The refilling of these storage vessels is intermittent. A vaporiser, two pumps and/or compressors are to be provided if required to meet this demand.

Dry Air For Plant and Instrument Air System

All plant and instrument air requirements for the IGCC are met by extracting air from each main air compressor of ASU. An air receiver will be provided common to both trains, sized for 10 minutes hold up at the flow given below. Each air compressor is sized for the extraction of 5,000 Nm^3/hr, however under normal circumstances the compressors shall share the duty equally.

Flow	5,000 Nm^3/h
Dew Point	- 20°C @ 7.0 bar g

3.4 Product Storage

The continuity of supply of oxygen and nitrogen to the IGCC Plant is extremely critical.
The Air Separation Unit can be considered as an essential service since in case of complete failure it will result in the entire IGCC Complex not being available. For this reason two 50% Air Separation trains are installed and no equipment, except for the back-up systems, is shared between these two production trains.
In addition a liquid oxygen storage equivalent to at least 12 hours of a single ASU train and a back-up system shall be provided. This storage is sufficient to cover the majority of the ASU emergency failures ensuring a high availability (more than 98%).

In order to refill these systems in the time periods specified, ASU is “over-designed” above the normal oxygen and nitrogen requirements at 100% IGCC operation.

The liquid oxygen storage facilities have two pumps and one vaporiser during the period necessary to reach the steady flowrate of the back-up vaporiser, a gaseous buffer tank with a capacity of at least two minutes of 50% ASU design capacity shall ensure the required oxygen flowrate.
Also the nitrogen system is provided with a liquid storage designed to ensure for Shell gasification cases 12 hours of a single ASU train continuous nitrogen requirements of the Gasification Island. In addition for both technologies the liquid storage is suitable to ensure low pressure nitrogen required for purging, blanketing etc. for 12 hours continuous operation of the IGCC Complex, and a safe shutdown in case of gasifier failure.

4.0 Syngas Treatment and Conditioning Line

This Unit receives the raw syngas from the gasification section, which is hot, humid and contaminated with acid gases, CO_2 and H_2S, and other chemicals, mainly COS, HCN and NH_3.
Before using this syngas as fuel in the gas turbines it is necessary to remove all the contaminants and prepare the syngas at the proper conditions of temperature, pressure and water content in order to achieve in the combustion process of the gas turbine the desired environmental performance and stability of operation.

Depending on the design alternative under consideration, amongst the different cases, this unit includes the following processing steps:

- catalytic conversion of CO to H_2 and CO_2 (shift reaction; based on a catalyst that can be suitable to process either sulphur containing syngas (sour shift);
- syngas cooling in waste heat boilers, recovering HP, MP, LP and VLP steam;
- further cooling of syngas by preheating process condensate;
- reduction of pressure from the gasification pressure to the pressure required by the gas turbine. For GEE gasification technology, this pressure reduction is achieved by an expansion turbine, recovering energy, or by control valve;
- preheating of clean syngas before entering the gas turbine combustion chamber.

Each of the cases examined in the study has a different combination and sequence of the above listed processing steps.
Section D and G of the study provides for each case a description of this unit, with the support of process flow diagrams.

The shift of CO to H_2 and CO_2 is a catalytic step necessary when the IGCC must reduce the CO_2 discharged to the atmosphere thus it's considered critical for the environmental impact of the IGCC. In fact the addition of CO shift brings the following benefits:

- CO shift reaction is exothermic and eliminates part of the syngas water coming from the quench. This results, downstream, in more availability of high temperature heat, for HP steam production, and less low temperature heat for LP steam production.
  With a quench gasifier without shift, heat can only be recovered as MP and LP steam.
- CO shift catalyst also hydrolyses COS to H_2S and there is no need of a separate COS hydrolysis system.

- The greater mass flow of syngas, due to CO_2, increases the energy recoverable from the expander.
- More CO_2 in the gas turbine reduces the quantity of H_2O to be added to saturate the expander and, at the same time, contributes to NO_x reduction.

For Catalytic Conversion of CO to H_2 and CO_2 Synetix and Süd Chemie provided Shift Reactors data.

5.0 Acid Gas Removal

The removal of acid gases, H_2S and CO_2, is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the complex.
Several different technologies are commercially available for acid gas removal. They can be grouped in 3 categories. The physical solvents, which capture the acid gas in accordance with the Henry's law; the chemical solvents, which capture the acid gas with a chemical reaction with the solvent, and the mixed solvents, which display both types of capture, physical and chemical. The first group is obviously favoured by a high partial pressure of the acid gas in the syngas, while the second group is less sensitive to the acid gas partial pressure.
The selection of the acid gas removal process for each of the alternatives examined in the study was done with a dedicated optimization study reported in Section H of this report.

The process description of the AGR used in each of the alternative cases is given in Section D and Section G. This description is limited to the information which the Licensor (UOP and DOW) of the process has authorized for disclosure, without a secrecy agreement by IEA.

6.0 Sulphur Recovery Unit and Tail Gas Treatment

The Sulphur Recovery Unit (SRU) processes the main acid gas from the Acid Gas Removal, together with other small flash gas and ammonia containing offgas streams coming from other units. SRU consists of two Claus Units, each sized for approx. 100% of the max sulphur production in order to assure a satisfactory service factor. Low pressure oxygen from ASU may be used as oxidant of Claus reaction.
The required recovery of sulphur from the entering streams is 95% minimum @ EOR (End Of Run), (95.5% minimum @ SOR, Start Of Run); it is obtained by means of thermal reactor plus two Claus catalytic reactors.
Each train is equipped with its own liquid sulphur product degassing facilities whereby each train sulphur pit (48 h minimum total hold up) is divided into

separate zones for collection from condensers etc. in the unit and for degassing (24 h hold up) plus transfer to liquid sulphur storage.
The Tail Gas Treatment Unit (TGT), is designed as a single train, capable of processing 100% tail gas resulting from the possible SRU operating modes.
A complete hydrogenation of SO_2, residual COS, CS_2 and elemental sulphur is achieved. After quenching tail gas is recycled back to the Acid Gas Removal (Unit 2300) by means of two tail gas recycle compressors (one operating, one spare).
In case a small quantity of hydrogen is needed for tail gas hydrogenation, back-up hydrogen containing gas (syngas) is available at SRU/TGT battery limit.

The catalyst selection shall be adequate to convert HCN and COS, in order not to accumulate them through the tail gas recycle to the solvent wash unit.
Ammonia contained in the feed gas streams to the Unit shall be completely destroyed.

However, due to the recycle of tail gas to the Acid Gas Removal, the sulphur recovery achieved in the IGCC Complex is significantly higher (more than 99 %).

Product Characteristics

Liquid Sulphur

State		liquid	
Colour		bright yellow	(at ambient temperature)
Sulphur content	wt %	99.9	min. (dry basis)
H_2S content	wt ppm	10	max.
Ash content	wt %	0.05	max.
Carbonaceous material	wt %	0.05	max.

7.0 CO_2 Compression and Drying

CO_2 as produced by the AGR section is required to be compressed up to 110 barg prior to export for sequestration, as per the IEA battery limit definition. CO_2 at these conditions is a supercritical fluid.

The incoming streams to CO_2 Compression and Drying Unit are three, at different pressures of between 1 and 30 barg. All of these streams require treating, to remove water, and compression. These requirements therefore present some alternatives:

- Provide separate dryers and compress the streams either with individual machines or a single machine;
- Use a pass-out compression system where the drier is operated at the highest pressure of the streams, and the compressor passes-out the remaining streams at the required pressure for drying in a single drier;
- Let down the higher pressure streams to the lowest pressure, dry at the low pressure and compress the combined LP stream to 110 bar g;
- Dry after compression at 110barg.

The flow rates of the streams are approx. similar, making the letdown option expensive, as this would add nearly 10% to the total compression duty compared against the first option. For this reason, the flowscheme described below has been adopted, based on the relative costs of the equipment involved and metallurgy considerations.

The stream at lowest pressure is compressed to intermediate pressure and routed to the molecular sieve drier, together with the stream at intermediate pressure, and the higher pressure stream which has been letdown to intermediate pressure. The letdown duty is available for power generation or turbine duty, but has been used adiabatically to cool the combined drier outlet to reduce the compressor power. The total combined stream at intermediate pressure is then dried in the molecular sieve dryers to remove the water to ensure no free water in CO_2 service. The final CO_2 moisture content of the product stream is less than 1 ppm. The dryers are provided as 2x50% units, each with 2x100% absorption beds, which are electrically regenerated. Total quantities of water removed are small, and are of sufficient quality for recycle to the steam system after appropriate dissolved gas removal. A buffer drum is provided to smooth the returned water flow from the batch dryers. The main equipment of the Drying Unit are as follows:

- Feed Heater
- 3 x Absorption Beds
- Aftercooler
- Water KO Drum
- After Filter (cartridge type)
- Recycle Blower
- Regeneration Heater
- Moisture Analyser

The dry gas is cooled against the incoming letdown service and routed to the compressors as 2x50% streams. The study is based on compressor information provided by Nuovo Pignone.
The compressor system recommended is of the following type:

- 2x50% machines (API 617);
- Between bearing design (NP 2MCL526 + gearbox + BCL405/A or equivalent);
- Auto-transformer with appropriate taps for start-up operation;
- 2 casings, 3 stages, dry gas seals;
- Speed: 9600 rpm;
- intermediate pressure inlet (different depending on cases);
- 110 bar g outlet.

It is noted that for the CO_2 flow rate required for compression, these machines are currently available on the market.

8.0 Hydrogen production unit

The feed gas to the Hydrogen Production Plant is the purified Syngas from the Acid Gas Removal Unit.
The Plant consists of a hydrogen purification section based on Pressure Swing Adsorption (PSA) System.

The PSA system is based on the property of specific adsorbent materials to preferentially adsorb gaseous components different from hydrogen. The impurities are stopped by the adsorbents and rejected in the PSA Purge Offgas and are routed to burn in the HRSG postfiring system of the Power Island.

The process works on two pressure levels corresponding to two different phases: adsorption and regeneration.
Adsorption of impurities takes place in a HP environment (usually between 10 and 40 bars) in order to increase the partial pressure of the component in the mixture and correspondent loading of the impurities on the absorber material.
The regeneration phase consists of the regenerator adsorption of the impurities usually in a LP pressure environment and a consequent cleaning of the adsorption material.

Even if the plant has different stages, the plant is designed to work continuously because the different phases are cyclically alternated.

The PSA product gas is high purity hydrogen exported to battery limits at approximately 25 barg.

9.0 Power Island

The power island is based on different configurations in dependence of the considered case described in section G. For cases 1 and 2 it is based on two frame 9FA General Electric gas turbines, two Heat Recovery Steam Generators (HRSG) generating steam at 3 levels of pressure, and one steam turbine common to the two HRSGs. Case 3 is based on one 6FA General Electric gas turbine, one HRSG generating steam at 3 levels of pressure, and one steam turbine. Finally in cases 4 and 5 it is based on one 9FA General Electric gas turbine, one HRSG generating steam at 3 levels of pressure, and one steam turbine.

The power island is integrated with the other process units. The following interfaces generally exist, even if power island schemes may present some differences alternative by alternative:

- Compressed Air to Unit 2100 – Air Separation Unit (except case G3);
- HP steam generated in the gasification is superheated and processed in the steam turbine;
- MHP steam generated in the gasification and sent to the steam turbine (only for Shell gasification)
- Steam to moderate gasification temperature is supplied by the power island (for GEE alternatives only);
- MP and LP steam generated in the process unit are routed to the power island;
- BFW is supplied by the power island to the process units for steam generation;
- Process condensate recovered from the process units is recycled to the power island, after polishing.

The HRSG description provided below has to be considered as reference even if slightly variations may be present in any different alternatives. For each alternative in Section G, the eventual main differences of Power Island configuration with respect to the described case are listed.

During normal operation, the clean syngas, coming from Unit 2200 - Syngas Treatment and Conditioning Line, is heated up against MP BFW in a syngas final heater.
Before entering each machine the hot syngas goes through a dedicated final separator in order to protect the Gas Turbine from liquid entrainment, mainly during cold start-up.
Finally, the hot syngas is burnt inside the Gas Turbine to produce electric power; the resulting stream of hot exhaust gas is conveyed to the Heat Recovery Steam Generator located downstream each Gas Turbine.

In dependence of case by case, compressed air is extracted from the Gas Turbines and delivered to the ASU. MP nitrogen coming from the ASU is injected into the Gas Turbines for NOx abatement and power output augmentation.

The flue gas stream flows through superheaters, evaporators and economizers and then is discharged to the atmosphere with the stream coming from the other (if present) HRSG through a common stack at about 130 °C.

The condensate stream, extracted from the Steam Condenser by means of Condensate Pumps, is sent as Cold Condensate to the Polishing Unit, located in Unit 4200 – DM Water / Condensate Recovery System. Demineralised water makeup is mixed to the polished stream and finally is sent to the IGCC Process Units where it is heated up by recovering the low temperature heat available.

The Hot Condensate coming back from IGCC Process Units enters the VLP steam drum, which is equipped with the degassing tower operating at a temperature of 120 °C.
Degassed Boiler Feed Water for HP, MP, LP and VLP services is directly taken from deaerator and delivered to the relevant sections.

HP, MP and LP FW are delivered to the equivalent economizer by means of BFW pumps (two pumps for each pressure level, with one pump in operation and one in hot stand-by). Hot BFW for all the three pressure levels is then extracted at about 160 °C and sent to the IGCC Process Units. The three pressure level remaining BFW are then sent directly to dedicated evaporators or to an extra economizer coil and then to the evaporator.

The HP steam generated is then mixed with HP steam from the process, superheated in a dedicated coil and sent to the HP steam turbine. To control the maximum value of the HP superheated steam final temperature, a desuperheating station, located between the HP superheater coils, is provided.

The exhaust steam from the HP module of the steam turbine is sent back to the HRSG. Each stream feeds an MP header, and it is mixed with the MP generated steam from the relevant MP Evaporator coil, superheated and sent back to the steam turbine. To control the Reheated steam final temperature, a desuperheating station, located between Reheater coils, is provided.

The MHP steam from the process (only for Shell gasification) is processed into a dedicated steam turbine. Refer to the single cases for a precise description.

Finally The LP Steam generated is sent to the LP Steam distribution network as saturated steam.

The wet steam at the outlet of the LP module of the Steam Turbine is routed to the steam condenser. The cooling medium in the tube side of the surface condenser is seawater in once through circuit.
In section G, a detailed description of the steam turbine configuration is present.

Continuous HP, MP and LP blowdown flowrates from HRSGs are manually adjusted by means of dedicated angle valves; they are sent to the dedicated blowdown drum together with the possible overflows coming from HRSGs Steam Drums.
After flashing, recovered VLP steam is fed to the VLP steam drum while the remaining liquid is cooled down against cold condensate by means a dedicated Blowdown Cooler and delivered to the atmospheric blowdown drum.
Intermittent HP, MP and LP blowdown flowrates from HRSGs are manually adjusted by means of dedicated angle valves and sent to the dedicated atmospheric blow-down drum.

In case of Steam Turbine trip, live HP Steam is bypassed to MP manifold by means of dedicated letdown stations, while Reheated Steam and excess of LP steam are also let down and then sent directly into the condenser neck.

When the clean syngas production is not sufficient to satisfy the appetite of both Gas Turbines it is possible to co-fire natural gas or to switch to natural gas one or both Gas Turbines.
This could happen in case of partial or total failure of the Gasification/Gas Treatment units of the IGCC and during start-up.
The selected machines are suitable to co-fire syngas and natural gas from 20% to 100% load.

During Natural Gas Operation no air extraction is foreseen, while a stream of MP Steam has to be injected into the combustion chambers of the Gas Turbines to reduce the NO_x emissions.

During normal operation on Natural Gas, the Power Island does not export/import to/from IGCC Process Units any steam/water stream and no low temperature heat can be recovered in Process Units. Then all cold condensate coming from Steam Condenser can be directly sent to the deaerator after polishing.

In this situation, the degassing steam demand of the deaerator is very high, more than VLP steam produced by HRSG's that needs to be integrated with steam coming from LP and MP headers.

10. Utility and Offsite Units

Since the study is based on Shell gasification technology, the description of utility and offsite refers only to that kind of plants. Only the main units are described, as the other ones are typical units designed according to general standards.

10.1 Cooling Water/Fresh Cooling Water System (Unit 4100)

Unit 4100 includes the IGCC primary cooling system, sea water in once through circuit, and the IGCC secondary cooling system, fresh cooling water in closed circuit with relevant distribution system.

Five electric driven operating pumps are provided to pump sea water from the Sea Water Basin, located on the beach, to the IGCC site, and back to the sea. The sea water intake and the discharge to the sea, connected to the beach facilities by means of submarine lines, are located at a suitable distance in order not to mix the two streams, supply and return.

Inside the IGCC plant, sea water is used directly to condense steam in the steam turbine condenser, as cooling medium of the ASU and the CO_2 compression and drying Unit, and in a separate branch, after further pumping, to cool the Fresh Cooling Water. The machinery cooling water system produces fresh cooling water, circulating in a closed circuit, used as cooling medium for all IGCC users other than steam turbine condenser, CO_2 compression and ASU users.

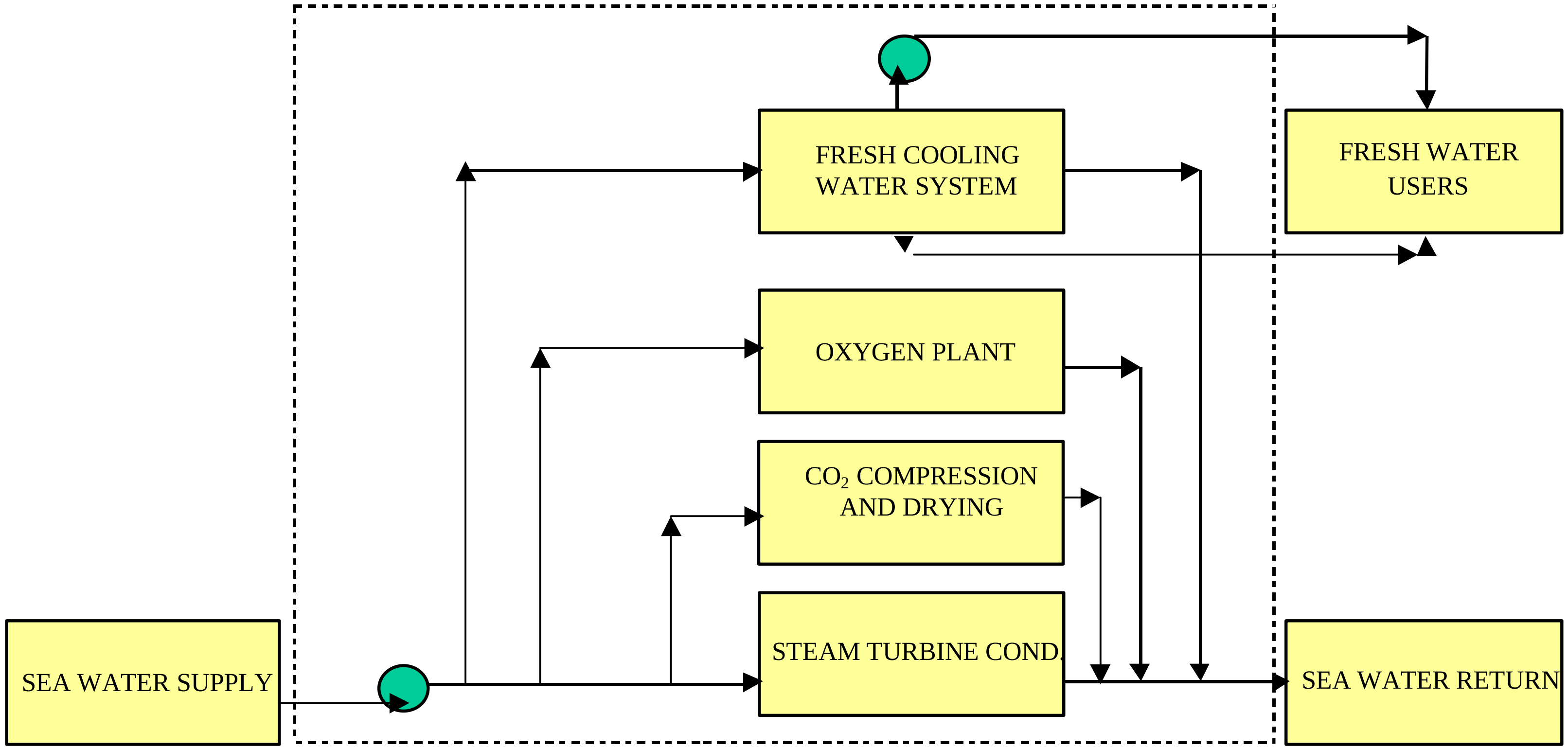
The max allowed sea water temperature increase is 7°C.

A plate heat exchanger type is selected to cool the machinery cooling water by means of sea water, in order to minimize the plot area, surface and pressure drop.

Self cleaning backflushing filters will be provided to protect plate exchangers from excessive sea water fouling.

A machinery cooling water expansion drum is installed to compensate the fluctuation of the water volume, due to the temperature variations.
Three electric driven pumps are provided to keep the machinery cooling water circulation, two operating and one spare.
Demineralized water is used as first filling of the machinery cooling water circuit and to compensate water losses.
A chemical injection system is provided in order to add the oxygen scavanger to the machinery cooling water circuit.

COOLING WATER/ FRESH COOLING WATER SYSTEM



10.2 Demi Water / Condensate recovery System (Unit 4200)

Raw water is used to produce Demineralized Water and as make-up water in the Gasification Island to close the Gasification water balance.

For the Shell cases with shift reaction, a large quantity of water is added to syngas to keep the reaction active. As a consequence, a large amount of condensate is recovered and sent to the Waste Water Treatment after stripping. Part of the effluent from the Waste Water Treatment (Unit 4600) is recovered and recycled back to the gasification island as process water, closing the Gasification water balance. The other part is sent to a dedicated treatment where the Reverse Osmosis process allows recovering almost 60% of the treated water. This recovered water is recycled back to the Demi Water System, Unit 4200, reducing the raw water to be fed to the Demineralized water plant. The remaining 40% of water is discharged together with the sea cooling water return stream.

Raw water flows through the Demineralized Water Plant, and is collected in the Demineralized Water Storage Tank. The Demi Water is pumped by the Demineralized Water Pump, taking suction from Demineralized Water Storage Tank and then fed to the combined cycle as make-up.

Condensate recovered from Process Units is collected in a Condensate Recovery Drum, where the condensate is cooled down with cold reflux. Output stream is then pumped by the Recovered Condensate Pump, cooled in the Condensate/Cold Condensate Exchanger, and divided into cold reflux and condensate streams. In the Condensate Recovery Drum temperature is controlled by the reflux steam flow and level is controlled by the condensate stream flow.
Condensate is cooled in the air cooler and then stored in the Condensate Storage Tank. After polishing in the condensate polishing Unit, this condensate is then pumped by the Condensate Pumps, taking suction from the tank, under level control, and fed to the Power Island via the Condensate/Cold Condensate Exchanger.
Cold condensate from Power Island (Steam Turbine condensate) enters Unit 2400 for polishing in the cold condensate polishing unit. Furtherly it flows to the Syngas Treatment and Conditioning Line for heating.

10.3 Waste Water Treatment (Unit 4600)

The Effluents from Unit 1000 - Gasification Island flow to the anaerobic section, where a phosphoric acid solution is added to the waste water to support the bacterial growth.

In the Anaerobic Reactor the organic pollutants are biodegraded with production of biological gas and biological sludge. The biogas produced in the reactor is routed to the local flare to be burned.

The biological mass exits the anaerobic reactor and enters the Anaerobic Clarifier where the biomass is separated by gravity from the supernatant.

Effluent from the anaerobic section is subject to a further aerobic treatment for the complete removal of ammonia and organic contaminants. The effluent from the anaerobic clarifier is pumped to the denitrification/oxidation tanks where it is mixed with the rainwater bleed-off and drainage coming from the deoiling section.

In this deoiling section, the oily drainage mixed with contaminated rainwater is fed by means of pumps from the oil water storage tank to the primary deoiling section, consisting of a Corrugate Plate Interceptor, which provides gravity separation of free oil and suspended solids carried in the waste water.

The effluent from the separator cells is dosed with polyelectrolyte and is routed by gravity to a secondary deoiling step, consisting of Induced Air Flotation. Air induced by a motor driven self aerating rotors mechanism removes the oil and suspended solids, which are collected in a dense froth to be recycled back to the CPI.
The deoiled water is then pumped to the denitrification/oxidation tanks, where it is mixed with the section from the anaerobic treatment effluent and where the organic contaminants are removed and ammonia is oxidized to nitrates which are further reduced to nitrogen gas in the denitrification section.

The effluent from the oxidation tank enters the aerobic clarifier, where the biomass separates by gravity from the supernatant. The sludge from the bottom of the clarifier is recycled to the anaerobic reactor by the Sludge Pump.

The supernatant from the clarifier is dosed with polyelectrolyte and pumped into Dual Media Filter, which uses sand and anthracite as filter media for the removal of residual hydrocarbons and suspended solids, and into Activated Carbon Filters, for the complete removal of organic contaminants.

From the filters the water is sent to the Reverse Osmosis process.

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME	:	HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME:		GASIFICATION TECHNOLOGY SELECTION - BASIC INFORMATION FOR EACH ALTERNATIVE

ISSUED BY	:	M. GALLIO
CHECKED BY	:	P. COTONE
APPROVED BY	:	S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	M. Gallio	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

SECTION D

GASIFICATION TECHNOLOGY SELECTION
BASIC INFORMATION FOR EACH ALTERNATIVE

INDEX

SECTION D

0.0 Introduction

Scope of this section is the technical description of the three gasification technologies considered in the first phase of the study.

The three gasification technologies are the following:
1. GEE gasification – Case 0A
2. Shell gasification – Case 0B
3. Siemens gasification – Case 0C

The comparison, both from technical and economical point of view, is carried on in section F.

For each case, the gasification island is sized in order to satisfy the appetite of two 9FA gas turbines in combined cycle. In Unit 2200 (Syngas Treatment and Conditioning Line) the syngas is split into two equal streams: half of the syngas generated is dedicated to the power generation in a combined cycle based on one Gas Turbine 9FA and the second half is dedicated to the hydrogen production. The offgas coming from the hydrogen production unit is routed to the post firing system of the HRSG.

For GEE case, reference is made to a previous study that FWI made for IEA GHG in 2003 (Gasification Power Generation Study). The study was performed on the same coal and a similar plant configuration (without Hydrogen production).

For Shell case, reference is made to a previous study that FWI made for IEA GHG in 2003 (Gasification Power Generation Study) and to a technical and economical upgrading of the offers that Shell communicated to FWI in 2005. The 2003 study was performed on the same coal and a similar plant configuration (without Hydrogen production) as in the present study.

For Siemens case, reference is made to in house data elaborated based on the FWI experience in gasification. No data are available from Siemens for this study.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 0.A - GEE GASIFICATION TECHNOLOGY

ISSUED BY : M. GALLIO
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	M. Gallio	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

SECTION D.1

BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION D.1 BASIC INFORMATION FOR EACH ALTERNATIVE

1.0 Case 0.A

1.1 Introduction

The main features of the Case 0.A configuration of the IGCC Complex are:

- High pressure (65 bar g) GEE Gasification;
- Coal Water Slurry Feed;
- Gasifier Quench Type;
- Single Stage Dirty Shift;
- Separate Removal of H_2S and CO_2;
- PSA Unit for Hydrogen Production.

The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The degree of integration between the Air Separation (ASU) and the Gas Turbines is 25%. Gas Turbine power augmentation and syngas dilution for NO_x control are achieved with injection of compressed N_2 from ASU to the Gas Turbines.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

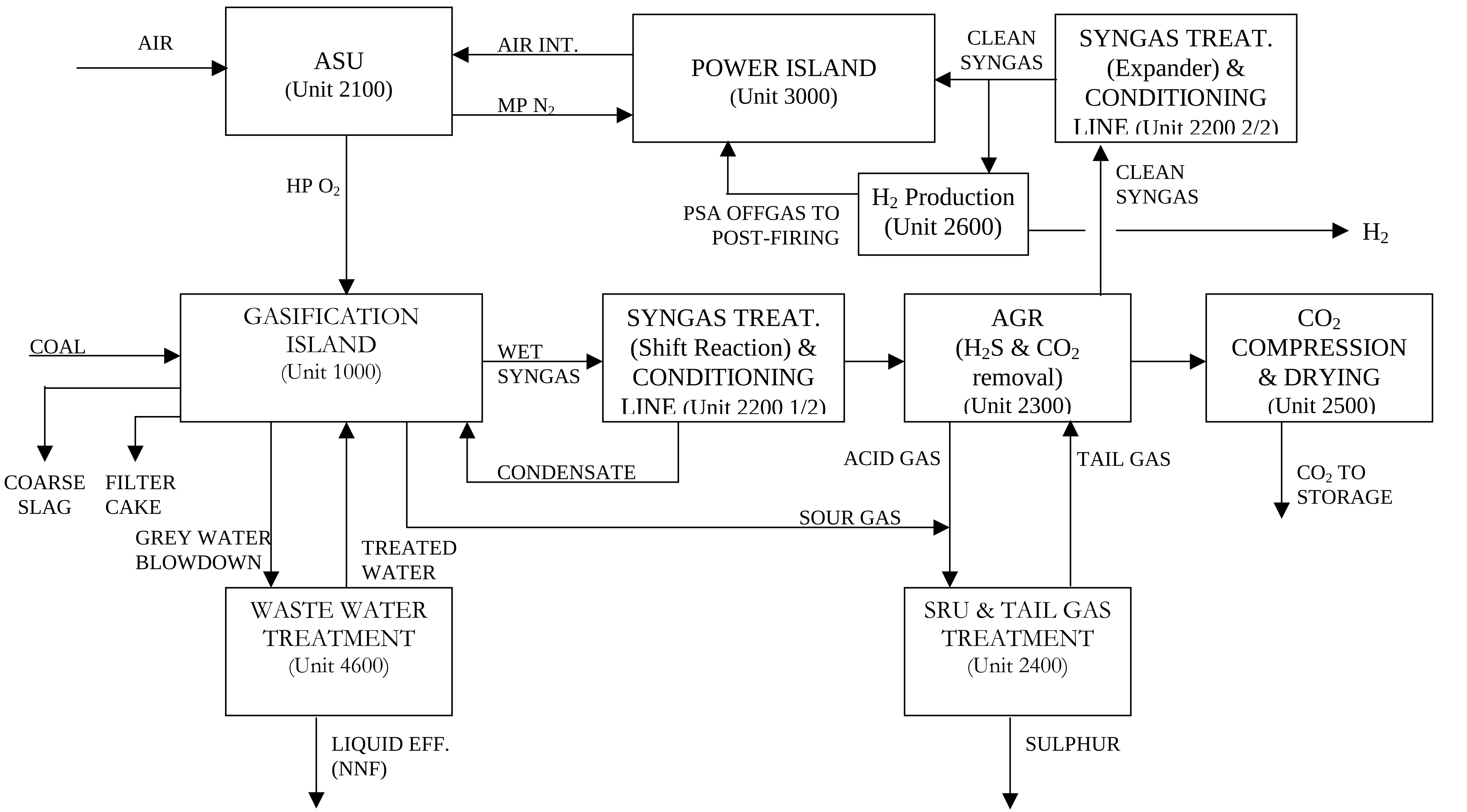
The arrangement of the process units is:

Unit		Trains
1000	Gasification	4 x 33 %
	Waste water pre-treatment	2 x 66 %
2100	ASU	2 x 50 %
2200	Syngas Treatment and Conditioning Line	2 x 50 %
	Syngas Expansion	1 x 100%
2300	AGR	1 x 100%
2400	SRU	2 x 100%
	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50 %

2600	H_2 Production	1 x 100%
3000	Gas Turbine (PG 9351-FA)	1 x 100%
	HRSG	1 x 100%
	Steam Turbines	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.

GEE 0.A – IGCC COMPLEX BLOCK FLOW DIAGRAM



1.2 Process Description

Unit 1000: Gasification Island

Information relevant to GEE Gasification Island are collected in para 1.2 of Section C.
The main process data of the Gasification Island relevant to this alternative are summarised in the following table:

STREAM	FUEL FEED (COAL)	HP OXYGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	149	243
Pressure (bar)	AMB.	80	63
TOTAL FLOW			
Mass flow (kg/h)	323,100	278,700	1,388,000
Molar flow (kmol/h)		8,650	72,260
Composition (%vol)			
H_2			15.1
CO			15.6
CO_2			7.3
N_2 + Ar		5	0.8
O_2		95	-
H_2S + COS			0.12
H_2O			61
Others			0.08

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package unit supplied by specialised Vendors. Reference is made to Section C, para. 2.0 for a general description of the Air Separation Unit.

The integration between ASU and Gas Turbine has been optimized considering a reference plant with two gas turbines in operation without hydrogen production as the gasification island is sized in order to satisfy the appetite of two gas turbines 9FA in combined cycle. In this configuration, the optimum integration between ASU and Gas Turbine is 50%. When the gasification operates at full load, 50% of the air required by the ASU to obtain the design oxygen production is derived from both gas turbine compressors; the integration between the gas turbines operation and the ASU is achieved at a level where 50% of the atmospheric air is compressed with selfstanding units and the difference comes already pressurized from the compressors of the gas turbines in the combined cycle.

For the gasification technology selection, only one gas turbine has been considered, as half of the clean syngas flowrate, coming from Unit 2200 is sent to Hydrogen production. In this configuration, the same fraction of air extraction from the gas turbine is considered and as a consequence the integration between ASU and Gas Turbine is half of the optimized figure for a power-only plant (25%).

The main process data and the main consumption of the ASU are summarised in following tables.

	Mass Flow (kg/h)
Air from ambient	930,000
Air from GT	310,000
Oxygen to gasifier (95% vol)	278,700
LP Nitrogen to Gasification Island (98% vol)	-
HP Nitrogen to Gasification Island (98% vol)	-
Nitrogen to Power Island (for syngas dilution)	438,300

Main air compressor	74,500	kW
Oxygen compressor	28,000	kW
Nitrogen compressor	20,400	kW
Miscellanea	1,600	kW
Total	124,500	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from Unit 1000, at approximately 240°C and 62 barg enters the Sour Shift section of Unit 2200. The syngas is first heated in a gas/gas exchanger by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2, and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 434°C.

A single stage shift, containing sulphur tolerant shift catalyst (dirty shift), is used, this being sufficient to meet the required degree of CO_2 removal.

The hot shifted syngas is cooled in a series of heat exchangers:

Shift feed product exchanger
HP Steam Generator
MP Steam Generator
LP Steam Generator
VLP Steam Generator

Process condensate collected in the cooling process of the syngas is accumulated and from there pumped back to the syngas scrubber of Unit 1000.
The final cooling step of the syngas takes place preheating the cold condensate from CCU. The process condensate separated after this step is routed to Unit 4000, Sour Water Stripper, being heavily contaminated, the remaining part is accumulated in a drum.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
Clean syngas is preheated with VLP steam and then reduced in pressure, down to 26 bar (g) in the Expander, generating electric energy.
The syngas is then split in two equal streams: one is fed to the hydrogen production unit; the remaining clean syngas is pre-heated with VLP steam and sent to the gas turbine (Unit 3000).

Unit 2300: Acid Gas Removal (AGR)

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a high syngas pressure (55 bar g) and an extremely high CO_2/H_2S ratio (183/1).

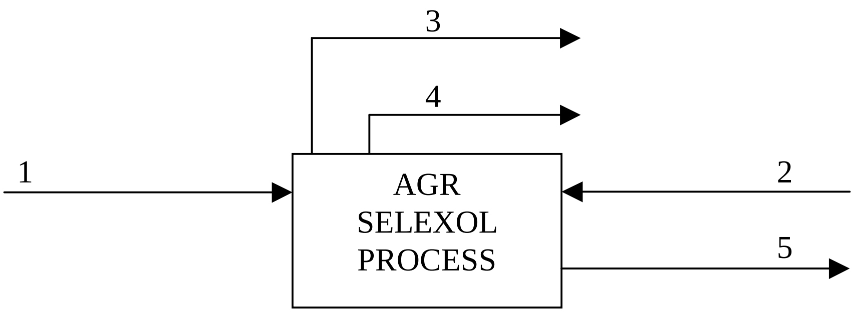
The interfaces of the process are the following, as shown in the scheme:

Entering Streams

1. Untreated Gas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit

Exit Streams

3. Treated Gas to Expander
4. CO_2 to compression.
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in the following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treatment	**Recycle Tail Gas from SRU**	**Treated Syngas to Expander**	**CO_2 to Compression**	**Acid Gas to SRU & TGT**
Temperature (°C)	38	38	30	-	49
Pressure (bar)	57.2	28.3	56.2	(1)	1.8
Mass flow (kg/h)	776000	25294	159700	626354	19573
Molar flow (kgmole/h)	38370	622	24060	14550	485
Composition (vol %)					
H_2	55.04	2.88	86.75	1.80	0.37
CO	2.84	0.03	4.43	0.17	0.04
CO_2	40.22	83.71	6.47	97.12	75.15
N_2	0.68	12.47	1.07	0.55	0.00
O_2	0.00	0.00	0.00	0.00	0.00
CH_4	0.02	0.00	0.03	0.00	0.00
H_2S + COS	0.22	0.52	0.00	0.01	17.94
Ar	0.79	0.13	1.23	0.05	0.01
H2O	0.19	0.26	0.02	0.30	6.49

Note: (1) CO_2 stream is the combination of three different streams at following pressure levels: 28 bara; 11 bara; 1.5 bara.

The Selexol solvent consumption, to make-up losses, is 120 m^3/year.

The proposed process matches the process specification with reference to concentration of the treated gas exiting the Unit. In fact the H_2S+COS concentration is 4 ppm. This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power consumption = 32% of the overall AGR power requirement) before flowing to the CO_2 absorber.
The CO_2 removal rate is more than 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 19% dry basis, more than suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 262 kmol/h of Hydrogen, corresponding to 1.8% vol and to an overall thermal power of 17.7 MWt, i.e. more than 5.8 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 92 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constant of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

Unit 2400: SRU and TGT

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The Sulphur Recovery Section consists of two trains each sized for a production of 66.8 t/day and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 28 bara.

Unit 2500: CO_2 Compression and Drying

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 27 barg
- LP stream : 10 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is composed of CO_2 and H_2+N_2 coabsorbed. The main properties of the stream are as follows:

- Product stream : 626 t/h.
- Product stream : 110 bar.
- Composition :

	%wt
CO_2	99.4
N_2	0.3
H_2	0.1
Others	0.2
TOTAL	100.0

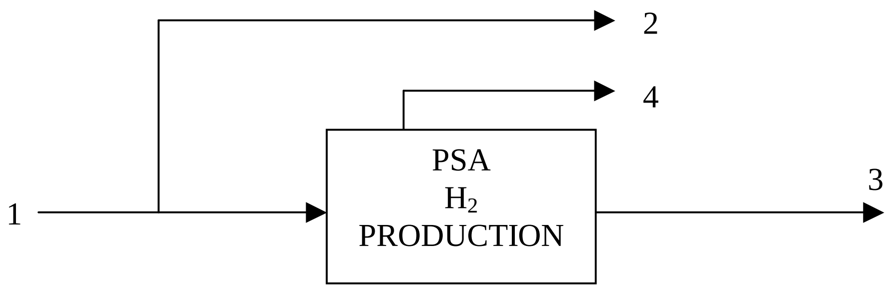
Unit 2600: H_2 Production

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 8.0 for the general information about the technology.

A small portion of the syngas entering the hydrogen production unit bypasses the PSA and is sent to the post-firing system of the HRSG together with the PSA offgas to make the burner flame stable.

The interfaces of the process are the following, as shown in the scheme:

1 – Total Clean Syngas from AGR Unit
2 – Bypass to post-firing
3 – Hydrogen
4 – Offgas to post-firing



The main process data of the hydrogen production unit are summarised in following table:

	1 Syngas	2 Bypass	3 H_2	4 Offgas
Hydrogen	86.75	86.75	99.50	46.70
Nitrogen	1.07	1.07	0.40	3.17
Argon	1.23	1.23	0.10	4.78
Carbon Monoxide	4.43	4.43		18.35
Carbon Dioxide	6.47	6.47		26.79
Methane	0.03	0.03		0.12
Water	0.02	0.02		0.08
Hydrogen Sulfide	0.00	0.00		0.00
	100.00	100.00	100.00	100.00
Flow (Nm^3/h)	269,646	5,302	200,510	63,834
Flow (kmol/h)	12,030	237	8,946	2,848
(kg/h)	79,845	1,570	19,270	59,004
p (barg)	26.0	26.0	25.2	0.7
T (°C)	34	34	39	26

<u>Unit 3000: Power Island</u>

Reference is made to Section C, para. 9.0 for the general information about the technology.
For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

- HP steam (160 barg) : steam imported from Syngas Treatment and Conditioning Line.
- HP steam (85 barg) : steam exported to the Gasification Island users.
- MP steam (40 barg) : steam imported from Syngas Treatment and Conditioning Line. A small quantity is also generated in the Sulphur Recovery Unit.
- LP steam (6.5 barg): steam imported from Syngas Treatment and Conditioning Line. A small quantity is also generated in the Sulphur Recovery Unit.
- VLP steam (3.2 barg): steam imported from Syngas Treatment and Conditioning Line.
- BFW : HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate : All the condensate recovered from the condensation of the steam utilised in the Process

Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.

- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Cooling and Conditioning Line and recycled back to the HRSG.

Flow rate of the above interfaces of the Plant are shown in table attached to para 1.3, Utility Consumption.

The net balance on each steam header inside the Power Island is positive, thus meaning that for all generation levels steam is imported from Process Units to the Power Island. Only steam at 85 bar g is exported to the Gasification Island. As a consequence, the generation levels of the Power Island are the same of the Process Units.

1.3 Utility Consumption

The utility consumption of the process / utility and offsite units are shown in the attached Tables.

FOSTER WHEELER

CLIENT: IEA GHG
PROJECT: H2 and Electricity co-production
LOCATION: the Netherlands

REVISION	Rev 1		
DATE	July 2007		
ISSUED BY	LV		
CHECKED BY	PC		
APPROVED BY	SA		

UTILITY CONSUMPTION SUMMARY - GEE - CASE 0A - HP with CO_2 capture, separate removal of H_2S and CO_2

UNIT	DESCRIPTION UNIT	HP Steam 160 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS										
1000	Gasification Section	5.1 (2)								5.1	
2100	Air Separation Unit			21.5						21.5	
2200	Syngas Treating and Conditioning Line	-52.6	-121.5	-528.3	-29.9	52.6	121.5	528.3	72.3	42.4	
2300	Acid Gas Removal			72.4						72.4	
2400	Sulphur Recovery (SRU)- Tail gas treatment (TGT)		-1.3	-1.2			4.4	1.2		3.0	
3000	**POWER ISLANDS UNITS**	47.5	122.8	423.6	29.9	-52.6	-125.9	-529.5	-72.3		
4000 to 5300	**UTILITY and OFFSITE UNITS**			12.0						12.0	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	156.4	0.0

Note: (1) Minus prior to figure means figure is generated
(2) Steam exported @ 85 barg

1.4 IGCC Overall Performance

The following Table shows the performance of the IGCC Complex.

GEE

Case 0A - High Pressure gasification with CO2 capture, separate removal of H_2S and CO_2

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	323.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWt**	**2321.8**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWt	1637.9
Gasification Efficiency (based on coal LHV)	%	70.5
Thermal Power of Clean Syngas (based on LHV)	MWt	1488.4
Syngas treatment efficiency	%	90.9
Hydrogen production (99.5% purity)	Nm^3/h	200,510
Hydrogen Thermal Power (E)	MWt	598.0
Equivalent H2 based combined cycle net efficiency	%	56.0
Gas turbines total power output	MWe	281.7
Steam turbine power output	MWe	332.2
Equivalent Electric Power from H_2	MWe	334.9
Expander turbine power output	MWe	11.2
ACTUAL GROSS ELECTRIC POWER	**MWe**	**625.1**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**960.0**
ASU power consumption	MWe	124.5
Process Units consumption	MWe	50.8
Utility Units consumption	MWe	1.9
Offsite Units consumption (including sea cooling water system)	MWe	10.0
Power Islands consumption	MWe	8.6
CO_2 compression and Drying	MWe	38.5
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**234.3**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**390.8**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**725.7**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**41.3**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**31.3**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**16.8**
Net H_2 output efficiency (E/A*100) (based on coal LHV)	**%**	**25.8**
H_2 thermal power Net Electric power generated ratio (E/B)		**1.53**

The following Table shows the overall CO_2 removal efficiency of the IGCC Complex.

	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	17393
Slag (Carbon =~4% wt)	708
Net Carbon flowing to Process Units (A)	16685
Liquid Storage	
CO	24
CO_2	14132
Total to storage (B)	14156.0
Emission	
CO_2	2523
CO	7
Total Emission	2530.0
Overall CO_2 removal efficiency, % (B/A)	84.8

1.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown at Section B - para 2.0, and to co-produce electric power and hydrogen. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

The gaseous emissions are not considered in this paragraph, as they do not affect the selection of the gasification technology. They are analysed in the development of the detailed cases for the selected technology.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 0.B – SHELL GASIFICATION TECHNOLOGY

ISSUED BY : M. GALLIO
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	M. Gallio	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

SECTION D.2 – Shell Gasification Technology

BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION D.2 GASIFICATION TECHNOLOGY SELECTION

SECTION D.2 BASIC INFORMATION FOR EACH ALTERNATIVE

2.0 Case 0.B

2.1 Introduction

The main features of the Case 0.B configuration of the IGCC Complex are:

- Low pressure (39 bar g) Shell Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Heat Recovery Type;
- Double Stage Dirty Shift;
- Separate Removal of H_2S and CO_2;
- PSA Unit for Hydrogen Production.

The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The degree of integration between the Air Separation Unit (ASU) and the Gas Turbines is 15%. Gas Turbine power augmentation and syngas dilution, for NO_x control, is achieved with injection of compressed N_2 from ASU to the gas turbines.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

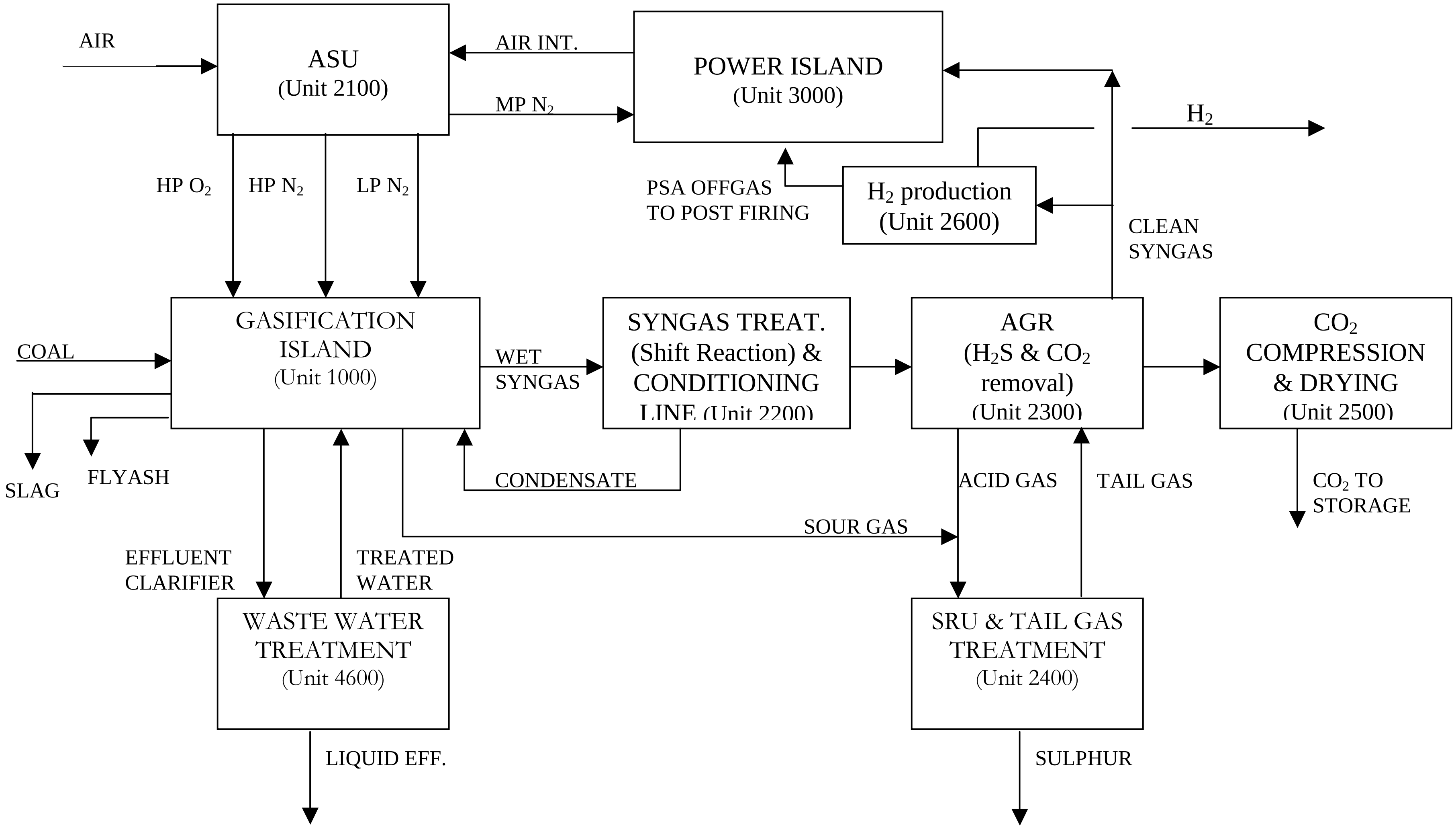
The arrangement of the process units is:

Unit		Trains
900	Coal milling and drying	4 x 33 %
1000	Coal pressurization/feeding	6 x 20 %
	Gasification heat recovery	2 x 50 %
	Slag removal	2 x 50 %
	Dry solids removal	2 x 50 %
	Wet scrubbing	2 x 50 %
	Sour slurry and sour water stripper	1 x 100 %
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%
2300	AGR	2 x 50%

2400	SRU	2 x 100%
	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50%
2600	H_2 production	1 x 100%
3000	Gas Turbine (PG 9351 – FA)	1 x 100%
	HRSG	1 x 100%
	Steam Turbine	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.

SHELL 0.B – IGCC COMPLEX BLOCK FLOW DIAGRAM



2.2 Process Description

Unit 1000: Gasification Island

Information relevant to Shell Gasification Island are collected in para 1.1 of Section C.
The main process data of the Gasification Island relevant to this alternative are summarised in the following table:

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	80	80	70	160
Pressure (bar)		40	69	7.5	37
TOTAL FLOW					
Mass flow (kg/h)	273,100	214,550	87,000	33,680	568,200
Molar flow (kmol/h)			3,100	1,200	28,850
Composition (% vol)					
H_2					26.25
CO					49.60
CO_2					1.24
N_2		3.5	99.88	99.88	4.00
Ar		1.5	0.08	0.08	0.62
O_2		95.0	0.04	0.04	0.00
H_2S + COS					0.23
H_2O					18.05
Others					0.01

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package unit supplied by specialised Vendors. Reference is made to Section C, para. 2.0 for a general description of the Air Separation Unit.

The integration between ASU and Gas Turbine has been optimized considering a reference plant with two gas turbines in operation without hydrogen production as the gasification island is sized in order to satisfy the appetite of two gas turbines 9FA in combined cycle. In this configuration, the optimum integration between ASU and Gas Turbine is 30%. When the gasification operates at full load, 30% of the air required by the ASU to obtain the design oxygen production is derived from both gas turbine compressors; the integration between the gas turbines operation and the ASU is achieved at a

level where 70% of the atmospheric air is compressed with selfstanding units and the difference comes already pressurized from the compressors of the gas turbines in the combined cycle.
For the gasification technology selection, only one gas turbine has been considered, as half of the clean syngas flowrate, coming from Unit 2200 is sent to Hydrogen production. In this configuration, the same fraction of air extraction from the gas turbine is considered and as a consequence the integration between ASU and Gas Turbine is half of the optimized figure for a power-only plant (15%).
The main process data and the main consumption of the ASU are summarised in following tables.

	Mass Flow (kg/h)
Air from ambient	804,300
Air from GT	141,900
Oxygen to gasifier (95% vol)	214,550
LP Nitrogen to Gasification Island (98% vol)	33,700
HP Nitrogen to Gasification Island (98% vol)	87,000
Nitrogen to Power Island (for syngas dilution)	304,350

Main air compressor	64,500	kW
Oxygen compressor	11,000	kW
Nitrogen compressor	22,200	kW
Miscellanea	1,400	kW
Total	99,100	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 36 barg and 160°C, enters the Sour Shift section of Unit 2200. The syngas is first heated in a gas/gas exchanger by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2 and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 451°C. Due to the low water content of the syngas, the injection of MP steam to the syngas is required before entering the shift reactor. In order to meet the required degree of CO_2 removal, a double stage shift containing sulphur tolerant shift catalyst (dirty shift) is used. The hot shifted syngas outlet from the first stage is cooled in a series of heat exchangers:

Shift feed product exchanger
HP Steam Generator

MP Steam Generator

Inlet temperature to the second stage shift is controlled to 250°C. Outlet temperature from second shift is 331°C. The hot shifted syngas outlet from the second stage is cooled in a series of heat exchangers:

MP Steam Generator
LP Steam Generator
VLP Steam Generator
Condensate fro CCU Preheater

The final cooling step of the syngas takes place in a cooling water cooler, where syngas is cooled with cooling water. Process condensate separated in syngas cooling is recycled back to the Sour Water Stripper of the Gasification Island.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
The syngas is then split in two equal streams: one is fed to the hydrogen production unit; the remaining clean syngas is preheated with VLP steam and sent to the gas turbine (Unit 3000).

Unit 2300: Acid Gas Removal (AGR)

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (27 bar g) and an extremely high CO_2/H_2S ratio (205/1).

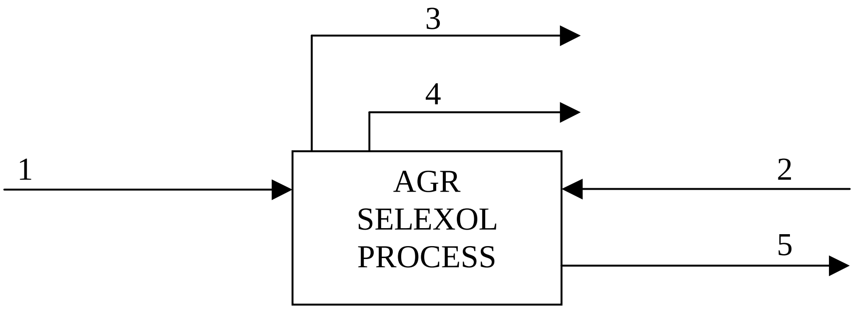
The interfaces of the process are the following, as shown in the scheme:

Entering Streams

1. Raw syngas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit.

Exit Streams

3. Treated Gas
4. CO_2 to compression
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in the following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treament	**Recycle Gas (tail gas) from SRU**	**Treated gas**	**CO_2 to compression**	**Acid gas to SRU**
Temperature (℃)	38	38	34	(1)	49
Pressure (bar)	27.8	27.0	27.0	(1)	1.8
Mass flow (kg/h)	714433	13011	164839	549273	13419
Molar flow (kgmole/h)	37113	332	24480	12728	336
Composition (vol %)					
H_2	56.51	4.10	85.35	1.74	0.28
CO	2.51	0.15	3.74	0.19	0.03
CO_2	36.91	76.63	5.24	97.69	72.41
N_2	3.10	17.78	4.93	0.06	0.01
CH_4	0.00	0.00	0.00	0.00	0.00
H_2S	0.18	0.72	0.00	0.01	20.25
COS	0.00	0.01	0.00	0.00	0.02
Ar	0.48	0.19	0.72	0.03	0.01
H2O	0.31	0.42	0.03	0.28	6.46

Note (1): CO_2 stream is the combination of three different streams at following pressure levels 26.0, 3.5 and 0.5 barg;

The Selexol solvent consumption, to make-up losses, is 120 m^3/year.

The proposed process matches the process specification with reference to H_2S+COS concentration of the treated gas exiting the Unit (H_2S+COS concentration is 3 ppm). This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power Consumption = 41% of the overall AGR Power requirement) before flowing to the CO_2 absorber.

The CO_2 removal rate is 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 22% dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1.7% vol and to an overall thermal power of 14.9 MWt, i.e. almost 5 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 100 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

<u>Unit 2400: SRU and TGT</u>

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 5.0 for the general information about the technology.

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of 56.4 t/day, and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 27 barg.

Unit 2500: CO_2 Compression and Drying

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 26.0 barg
- LP stream : 3.5 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is mainly composed of CO_2 and CO. The main properties of the stream are as follows:

- Product stream : 550 t/h.
- Product stream : 110 bar.
- Composition :

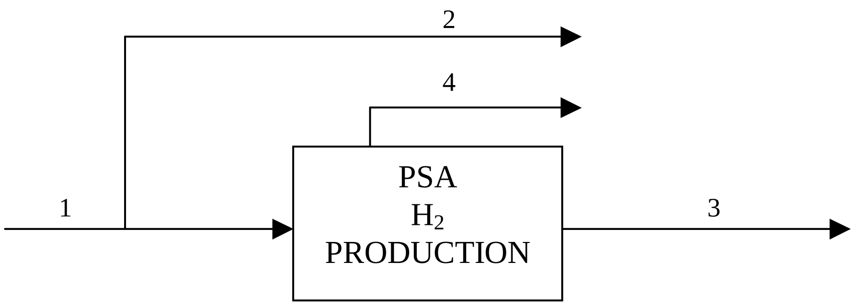
	%wt
CO_2	99.8
CO	0.1
Others	0.1
TOTAL	100.0

Unit 2600: H_2 Production

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 8.0 for the general information about the technology.
A small portion of the syngas entering the unit bypasses the PSA and is sent to the post firing system of the HRSG together with the PSA off gas to make the burners flame stable.

The interfaces of the process are the following, as shown in the scheme:

1. Total clean syngas from AGR
2. Bypass to post firing
3. Hydrogen
4. Offgas to post firing



The main process data of the hydrogen production unit are summarised in following table:

		1 Syngas	2 By pass	3 H_2	4 Offgas
Hydrogen		85.35	85.35	99.50	43.73
Nitrogen		4.93	4.93	0.40	18.25
Argon		0.72	0.72	0.10	2.54
Carbon Monoxide		3.74	3.74		14.74
Carbon Dioxide		5.24	5.24		20.65
Methane		0.00	0.00		0.00
Water		0.02	0.02		0.08
Hydrogen Sulfide		0.00	0.00		0.00
		100.00	100.00	100.00	100.00
Flow	(Nm^3/h)	274,296	5,149	200,858	68,289
Flow	(kmol/h)	12,238	230	8,961	3,047
	(kg/h)	82,571	1,550	19,303	61,717
p	(barg)	26.0	26.0	25.2	0.7
T	(°C)	34	34	39	26

Unit 3000: Power Island

Reference is made to Section C, para. 9.0 for the general information about the technology.

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

- HP steam (160 barg) : steam imported from Syngas Treatment and Conditioning Line.
- MHP steam (70 barg) : steam imported from Gasification section.

- MP steam (40 barg) : steam exported to Syngas Treatment and Conditioning Line to meet the water requirement of the shift reaction. A small quantity of steam is also generated in the Gasification Island and in the Sulphur Recovery Unit.
- LP steam (6.5 barg): steam exported to the following Process Units: AGR, ASU, Utility and Offsite Unit. LP steam is also generated in the Syngas Treatment and Conditioning Line.
- VLP steam (3.2 barg): steam imported from Syngas Treatment and Conditioning Line.
- BFW : HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate : All the condensate recovered from the condensation of the steam utilised in the Process Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

The MHP saturated steam at 70 bar from the gasification island, is superheated in a dedicated coil and sent to a dedicated ST section where is expanded. The exhaust steam is mixed with the exhaust steam from the ST IP section and flows to the ST LP main section. This steam turbine is coupled to the same generator of the main steam turbine. A dedicated clutch allows isolating the smaller steam turbine during the start-up of the plant.

Flow rate of the above interfaces of the Plant are shown in table attached to para 2.3, Utility Consumption.

2.3 Utility Consumption

The utility consumption of the process / utility and offsite units are shown in the attached Tables.

FOSTER WHEELER	CLIENT:	IEA GHG	
	PROJECT:	Hydrogen and Electricity co-production	
	LOCATION:	the Netherlands	

REVISION	Rev 1			
DATE	July 2007			
ISSUED BY	LV			
CHECKED BY	PC			
APPROVED BY	SA			

UTILITY CONSUMPTION SUMMARY - SHELL - CASE 0B - LP with CO_2 capture, separate removal of H_2S and CO_2

UNIT	DESCRIPTION UNIT	HP Steam 160 barg [t/h]	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS											
1000	Gasification Section		-317.4				390.9				41.3	32.2
2100	Air Separation Unit				16.8						16.8	
2200	Syngas Treatment and Conditioning line	-40.6		267.3	-75.5	-103.4	40.6	150.5	75.5	120.3	16.9	417.8
2300	Acid Gas Removal				82.4						82.4	
2400	Sulphur Recovery (SRU) - Tail gas treatment (TGT)			-0.7	-1.0			4.3	1.0		3.6	
3000	POWER ISLANDS UNITS	40.6	317.4	-266.6	-32.1	103.4	-431.5	-154.8	-76.5	-120.3		
4100 to 5300	**UTILITY and OFFSITE UNITS**				9.4						9.4	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	170.4	450.0

Note: Minus prior to figure means figure is generated

2.4 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex.

SHELL

Case 0B - Low Pressure gasification with CO2 capture, separate removal of H_2S and CO_2

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	273.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWt**	**1962.5**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWt	1638.2
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWt	1467.2
Syngas treatment efficiency	%	89.6
Hydrogen production (99.5% purity)	Nm^3/h	200,858
Hydrogen Thermal Power (E)	MWt	599.0
Equivalent H_2 based combined cycle net efficiency	%	56.0
Gas turbines total power output	MWe	286.0
Steam turbine power output	MWe	232.1
Equivalent Electric Power from H_2	MWe	335.4
ACTUAL GROSS ELECTRIC POWER	**MWe**	**518.1**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**853.5**
ASU power consumption	MWe	99.1
Process Units consumption	MWe	48.0
Utility Units consumption	MWe	2.5
Offsite Units consumption (including sea cooling water system)	MWe	7.5
Power Islands consumption	MWe	11.3
CO_2 compression and Drying	MWe	32.6
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**201.0**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**317.1**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**652.5**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**43.5**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**33.3**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**16.2**
Net H_2 output efficiency (E/A*100) (based on coal LHV)	**%**	**30.5**
H_2 thermal power Net Electric power generated ratio (E/B)		**1.89**

The following Table shows the overall CO_2 removal efficiency of the IGCC Complex.

	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	14,701
Slag (Carbon =~0.4% wt) *	61
Net Carbon flowing to Process Units (A)	14,640
Liquid Storage	
CO	24
CO_2	12,434
Total to storage (B)	12,458
Emission	
CO_2	2,177
CO	6
Total Emission	2,183
Overall CO_2 removal efficiency, % (B/A)	85.1

* The percentage of unreacted C stated by Shell is 0.2%. However, the carbon mass balance of the whole IGCC results in a 0.4% carbon less. This value is conservatively assumed.

2.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown at Section B - para 2.0, and co-produce electric power and hydrogen. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

The gaseous emissions are not considered in this paragraph as they do not affect the selection of the gasification technology. They are analysed in the development of the detailed cases for the selected technology.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 0.C - SIEMENS GASIFICATION TECHNOLOGY

ISSUED BY : M. GALLIO
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007 July 2007	Draft Rev 1	M. Gallio L. Valota	P. Cotone P. Cotone	S. Arienti S. Arienti

SECTION D.3

BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION D.3 BASIC INFORMATION FOR EACH ALTERNATIVE

SECTION D.3 BASIC INFORMATION FOR EACH ALTERNATIVE

3.0 Case 0.C

3.1 Introduction

The main features of the Case 0.C configuration of the IGCC Complex are:

- Low pressure (38 barg) Siemens Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Quench Type;
- Double Stage Dirty Shift;
- Separate Removal of H_2S and CO_2;
- PSA Unit for Hydrogen Production.

The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The degree of integration between the Air Separation (ASU) and the Gas Turbines is 15%. Gas Turbine power augmentation and syngas dilution for NO_x control are achieved with injection of compressed N_2 from ASU to the Gas Turbines.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

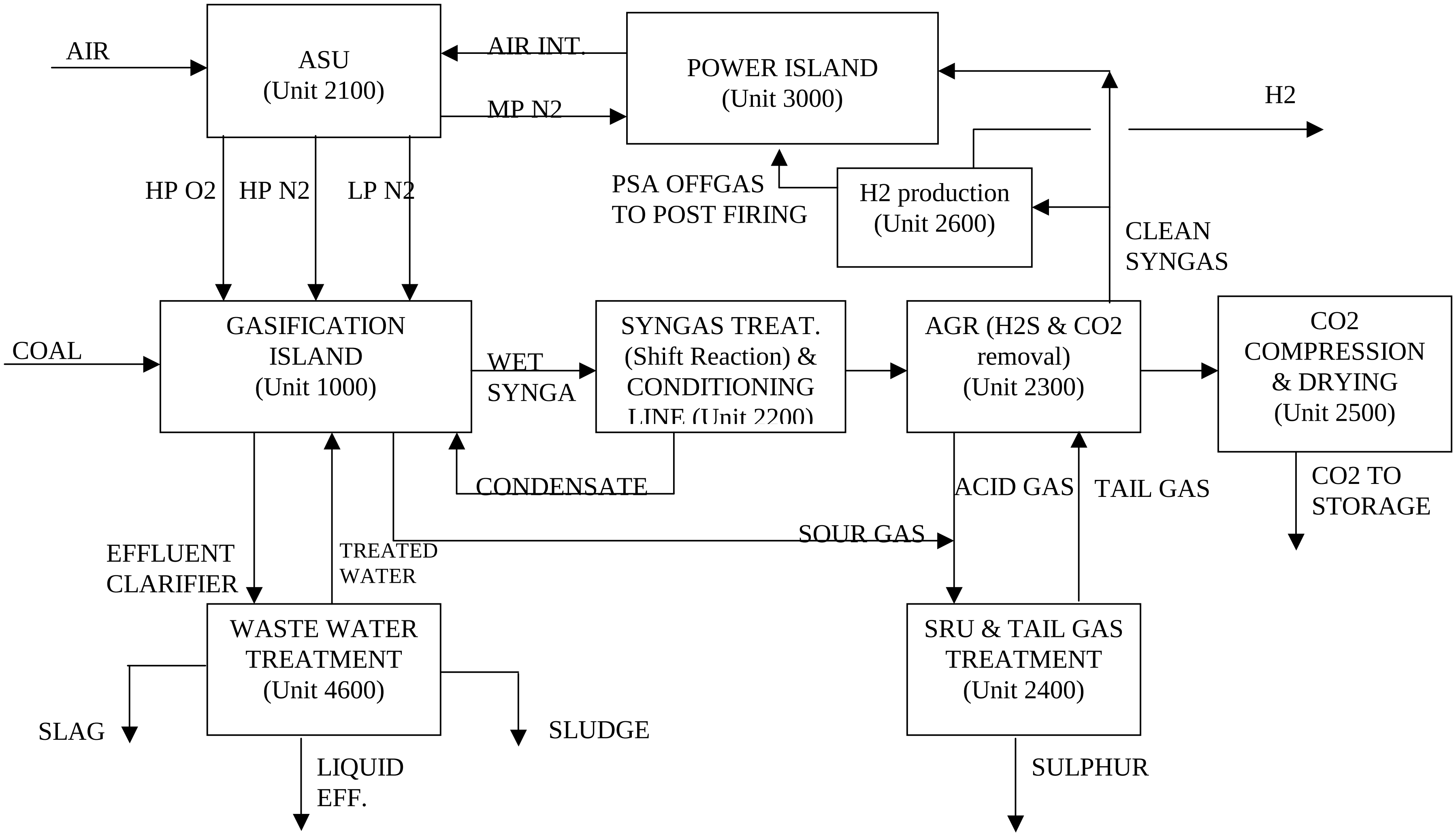
The arrangement of the main process units is:

Unit		Trains
900	Coal milling and drying	4 x 33 %
1000	Gasification	4 x 25%
	Waste Water Pre-treatment	1 x 100%
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%
2300	AGR	2 x 50%
2400	SRU	2 x 100%
	TGT	1 x 100%

2500	CO_2 Compression and Drying	2 x 50%
2600	H_2 Production	1 x 100%
3000	Gas Turbine (PG 9351 – FA)	1 x 100%
	HRSG	1 x 100%
	Steam Turbine	1 x 100%

Reference is made to the attached overall Block Flow Diagram of the IGCC Complex.

SIEMENS 0.C – IGCC COMPLEX BLOCK FLOW DIAGRAM



3.2 Process Description

Unit 1000: Gasification Island

Information relevant to Siemens Gasification Island are collected in para 1.4 of Section C.

The main process streams of the Gasification Island relevant to this alternative are summarised in the following table:

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	120	70	15	216
Pressure (bar)		48	55	7.5	37
TOTAL FLOW					
Mass flow (kg/h)	295,300	233,000	72,000	19,000	1,075,630
Molar flow (kmol/h)					53,870
Composition (% vol)					
H_2					11.3
CO					29.2
CO_2					1.9
N_2		3.5	99.88	99.88	3.0
O_2		95.0	0.04	0.04	0.0
H_2S + COS					0.2
H_2O					54.0
Ar		1.5	0.08	0.08	0.4

Note: Figures referred to the total flowrates

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package unit supplied by specialised Vendors. Reference is made to Section C, para. 2.0 for a general description of the Air Separation Unit.

The integration between ASU and Gas Turbine has been optimized considering a reference plant with two gas turbines in operation without hydrogen production as the gasification island is sized in order to satisfy the appetite of two gas turbines 9FA in combined cycle. In this configuration, the optimum

integration between ASU and Gas Turbine is 30%. When the gasification operates at full load, 30% of the air required by the ASU to obtain the design oxygen production is derived from both gas turbine compressors; the integration between the gas turbines operation and the ASU is achieved at a level where 70% of the atmospheric air is compressed with self-standing units and the difference comes already pressurized from the compressors of the gas turbines in the combined cycle.
For the gasification technology selection, only one gas turbine has been considered, as half of the clean syngas flowrate coming from Unit 2200 is sent to Hydrogen production. In this configuration, the same fraction of air extraction from the gas turbine is considered and as a consequence the integration between ASU and Gas Turbine is half of the optimized figure for a power-only plant (15%).

The main process data and the main consumption of the ASU are summarised in following tables.

	Mass Flow (kg/h)
Air from ambient	890,000
Air from GT	157,000
Oxygen to gasifier (95% vol)	233,000
LP Nitrogen to Gasification Island (98% vol)	19,000
HP Nitrogen to Gasification Island (98% vol)	72,000
Nitrogen to Power Island (for syngas dilution)	295,000

Main air compressor	72,300	kW
Oxygen compressor	14,300	kW
Nitrogen compressor	21,550	kW
Miscellanea	1,450	kW
Total	109,600	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 36 barg and 216°C, enters the Sour shift section of Unit 2200. The syngas is first heated in a gas/gas exchanger by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2 and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 460°C.

In order to meet the required degree of CO_2 removal, a double stage shift containing sulphur tolerant shift catalyst (dirty shift) is used. The hot shifted syngas outlet from the first stage is cooled in a series of heat exchangers:

Shift feed product exchanger
HP Steam Generator
MP Steam Generator

Inlet temperature to the second stage shift is controlled to 250°C. Outlet temperature from the second shift is 330°C. The hot shifted syngas outlet from the second stage is cooled in a series of heat exchangers:

MP Steam Generator
LP Steam Generator
VLP Steam Generator
Condensate from CCU Preheater

The final cooling step of the syngas takes place in a cooling water cooler, where syngas is cooled with cooling water. Process condensate separated in syngas cooling is recycled back to the Sour Water Stripper of the Gasification Island.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
The syngas is then split in two equal streams: one is fed to the hydrogen production unit; the remaining clean syngas is preheated with VLP steam and sent to the gas turbine (Unit 3000).

Unit 2300: Acid Gas Removal (AGR)

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (27 bar g) and an extremely high CO_2/H_2S ratio (204/1).

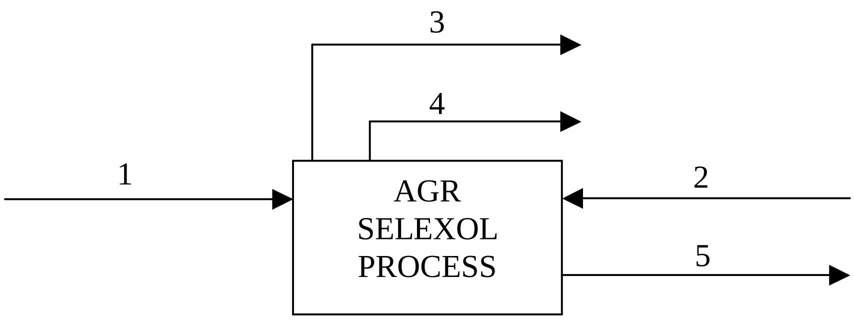
The interfaces of the process are the following, as shown in the scheme:

Entering Streams

1. Raw syngas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit.

Exit Streams

3. Treated Gas
4. CO_2 to compression
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treament	Recycle Gas (tail gas) from SRU	Treated gas	CO_2 to compression	Acid gas to SRU
Temperature (°C)	38	38	34	(1)	49
Pressure (bar)	27.8	27.0	27.0	(1)	1.8
Mass flow (kg/h)	817963	25627	196006	628340	5
Molar flow (kgmole/h)	39620	631	25199	14525	441
Composition (vol %)					
H2	52.7	2.2	82.0	1.5	0.2
CO	2.5	0.0	3.8	0.2	0.0
CO2	39.8	82.2	6.9	98.0	76.7
N2	3.9	14.7	6.1	0.1	0.0
CH4	0.0	0.0	0.0	0.0	0.0
H2S	0.2	0.5	0.0	0.0	18.3
COS	0.0	0.0	0.0	0.0	0.0
Ar	0.7	0.2	1.1	0.0	0.0
H2O	0.3	0.2	0.0	0.2	4.8

Note: (1) CO_2 stream is the combination of three different streams at following pressure levels: 27 bar; 4.5 bar; 1.5 bar.

The Selexol solvent consumption, to make-up losses, is 126 m^3/year.

The proposed process matches the process specification with reference to H_2S+COS concentration of the treated gas exiting the Unit (H_2S+COS concentration is 4 ppm). This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power Consumption = 41% of the overall AGR Power requirement) before flowing to the CO_2 absorber.

The CO_2 removal rate is around 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 22% dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1.5% vol and to an overall thermal power of 14.9 MWt, i.e. almost 5 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 100 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

Unit 2400: SRU and TGT

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 5.0 for the general information about the technology.

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of around 57 t/day, and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 27 bar.

Unit 2500: CO_2 Compression and Drying

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 26.0 barg
- LP stream : 3.5 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is mainly composed of CO_2 and CO. The main properties of the stream are as follows:

- Product stream : 628 t/h.
- Product stream : 110 bar.
- Composition :

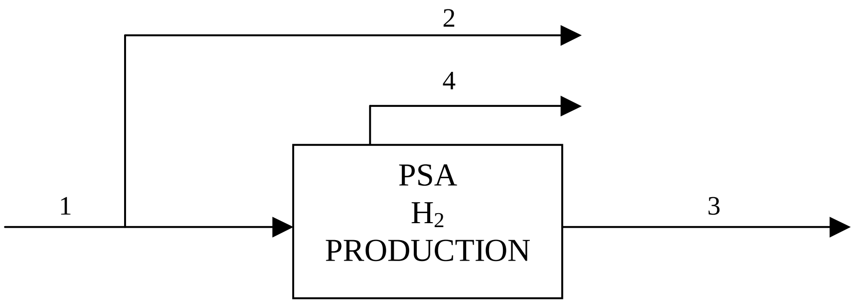
	%wt
CO_2	99.8
CO	0.1
Others	0.1
TOTAL	100.0

Unit 2600: H_2 Production

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 8.0 for the general information about the technology.
A small portion of the syngas entering the unit bypasses the PSA and is sent to the post firing system of the HRSG together with the PSA off gas to make the burners flame stable.

The interfaces of the process are the following, as shown in the scheme:

1. Total clean syngas from AGR
2. Bypass to post firing
3. Hydrogen
4. Offgas to post firing



The main process data of the hydrogen production unit are summarised in following table:

	1 Syngas	2 By pass	3 Hydrogen	4 Tail gas
%mol - kmol/h				
H2/CO				
Hydrogen	82.0	82.0	99.5	37.7
Nitrogen	6.1	6.1	0.5	20.4
Argon	1.1	1.1	0.0	3.9
Carbon Monoxide	3.8	3.8		13.4
Carbon Dioxide	6.9	6.9		24.5
Methane	0.0	0.0		0.0
Water	0.0	0.0		0.1
Hydrogen Sulfide	0.0	0.0		0.0
	100.0	100.0	100.0	100.0
Flow (Nm^3/h)	282,416	5,536	198,519	78,361
Flow (kmol/h)	12,600	247	8,857	3,496
(kg/h)	99,987	1,960	18,972	79,056
p (barg)	26.0	26.0	25.2	0.7
T (°C)	34	34	39	26

Unit 3000: Power Island

Reference is made to Section C, para. 9.0 for the general information about the technology.

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

- HP steam (160 barg) : steam imported from Syngas Treatment and Conditioning Line.
- MP steam (40 barg) : steam imported from Syngas Treatment and Conditioning Line.
- MP steam (40 barg) : steam exported to the Gasification Island users
- LP steam (6.5 barg) : steam imported from Syngas Treatment and Conditioning Line. The steam is also exported to the following Process Units: AGR, ASU, Utility and Offsite Unit.
- VLP steam (3.2 barg) : steam imported from Syngas Treatment and Conditioning Line.
- BFW : HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate : All the condensate recovered from the condensation of the steam utilised in the Process Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

Flow rate of the above interfaces of the Plant are shown in table attached to para 3.3, Utility Consumption.

3.3 Utility Consumption

The utility consumption of the process / utility and offsite units are shown in the attached Tables.

FOSTER WHEELER

CLIENT: IEA GHG
PROJECT: H2 and Electricity co-production
LOCATION: the Netherlands

REVISION	Rev 1
DATE	July 2007
ISSUED BY	LV
CHECKED BY	PC
APPROVED BY	SA

UTILITY CONSUMPTION SUMMARY - Siemens - CASE 0C - LP with CO_2 capture, separate removal of H_2S and CO_2

UNIT	DESCRIPTION UNIT	HP Steam 160 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS										
1000	Gasification Section			-28.0		19.0		99.0		65.0	25.0
2100	Air Separation Unit			20.0						20.0	
2200	Syngas Treating and Conditioning Line	-71.7	-150.9	-77.7	-56.1	71.7	150.9	77.7	70.0	13.9	
2300	Acid Gas Removal			89.2						89.2	
2400	Sulphur Recovery (SRU)- Tail gas treatment (TGT)		-1.2	-1.1			3.9	1.1		2.7	
3000	**POWER ISLANDS UNITS**	71.7	152.1	-13.9	56.1	-90.7	-154.8	-177.8	-70.0		
4000 to 5300	**UTILITY and OFFSITE UNITS**			11.5						11.5	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	202.3	25.0

Note: (1) Minus prior to figure means figure is generated

3.4 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex.

Siemens

Case 0C - Low Pressure gasification with CO2 capture, separate removal of H_2S and CO_2

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	295.3
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWt**	**2122.0**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWt	1659.3
Gasification Efficiency (based on coal LHV)	%	78.2
Thermal Power of Clean Syngas (based on LHV)	MWt	1467.0
Syngas treatment efficiency	%	88.4
Hydrogen production (99.5% purity)	Nm^3/h	198,500
Hydrogen Thermal Power (E)	MWt	591.8
Equivalent H2 based combined cycle net efficiency	%	56.0
Gas turbines total power output	MWe	286.0
Steam turbine power output	MWe	252.5
Equivalent Electric Power from H2	MWe	331.4
ACTUAL GROSS ELECTRIC POWER	**MWe**	**538.5**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**869.9**
ASU power consumption	MWe	109.6
Process Units consumption	MWe	46.5
Utility Units consumption	MWe	2.4
Offsite Units consumption (including sea cooling water system)	MWe	9.0
Power Islands consumption	MWe	8.0
CO2 compression and Drying	MWe	36.2
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**211.7**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**326.8**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**658.2**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**41.0**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**31.0**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**15.4**
Net H2 output efficiency (E/A*100) (based on coal LHV)	**%**	**27.9**
H2 thermal power Net Electric power generated ratio (E/B)		**1.81**

The following Table shows the overall CO_2 removal efficiency of the IGCC Complex.

	Equivalent flow of CO_2, kmol/h
Carbon incoming (Coal carbon = 82.5%wt) Carbon incoming (Natural gas)	16,754 153
Slag	119
Net Carbon Flowing to Process Units (A)	16,788
Liquid Storage CO CO_2 Total to storage (B)	 25 14,236 14,261
Emission CO CO_2 Total Emission	 6 2,521 2,527
Overall CO_2 removal efficiency, % (B/A)	84.9

3.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown at Section B - para 2.0, and co-produce electric power and hydrogen. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

The gaseous emissions are not considered in this paragraph as they do not affect the selection of the gasification technology. They are analysed in the development of the detailed cases for the selected technology.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : GASIFICATION TECHNOLOGY SELECTION ECONOMICS

ISSUED BY : P. COTONE
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	P. Cotone	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

SECTION E

GASIFICATION TECHNOLOGY SELECTION ECONOMICS

I N D E X

1.0 Introduction

This section summarises the economic data evaluated for each alternative of the study, including:

a. Investment cost;
b. Operation & Maintenance costs;
c. Electric power production cost.

2.0 Basis of Investment Cost Evaluation

2.1 Basis of the Estimate

The basis of the estimate for each alternative is the technical documentation collected in Sections C and D of this Study.
In particular the investment cost of the following Units or blocks of Units is detailed:

Unit 900	:	Coal Handling and Storage
Unit 1000	:	Gasification Section
Unit 2100	:	Air Separation Unit
Unit 2200	:	Syngas Treatment and Conditioning Line
Unit 2300	:	Acid Gas Removal
Unit 2400	:	Sulphur Recovery Unit and Tail Gas Treatment
Unit 2500	:	CO_2 Compression and Drying
Unit 2600	:	H_2 Production Unit
Unit 3000	:	Power Island
Units 4000 to 5200	:	Utilities and Offsites

The overall investment cost of each Unit or block of Units is split into the following items:

- Direct Materials, including equipment and bulk materials;
- Construction, including mechanical erection, instrument and electrical installation, civil works and, where applicable, buildings and site preparation;
- Other Costs, including temporary facilities, solvents, catalysts, chemicals, training, commissioning and start-up costs, spare parts etc.;
- EPC Services including Contractor’s home office services and construction supervision.

2.2 Estimate Methodology and Cost Basis

2.2.1 Direct Materials

The direct materials cost estimate of the main Units or Blocks of Units listed at para. 2.1 is developed according to the following general criteria:

Unit 900 (Coal Handling and Storage)

The cost of equipment delivered and erected is based on a budget quotation received from a qualified Vendor, detailing direct materials and construction costs.
The investment cost of the unit is calculated on the basis of the capacity of each alternative, as detailed in Section D. The unit includes, for Shell and Siemens cases, the drying section. For Shell and Siemens gasification systems, coal milling is included in Unit 900, while for GEE it's included in unit 1000.

Unit 1000 (Gasification)

Shell provided investment cost data of the main equipment in a study made in year 2003 with IEA GHG and FWI, based on same coal and gasification configuration as in the present study. In 2005 Shell provided updated technical and economical information for a second study that FWI performed for IEA GHG, based on lignite feedstock.
In the second study, Shell proposed steam generation in the WHB at much lower pressure than in the first study (70 barg vs. 130 barg), requiring lower investment cost. The investment cost of the gasification island in the present study is derived from this second study.
The figures have then been adjusted based on the actual syngas and coal flowrate resulting from finalization of the IGCC performances taking into account the different LHV.
After this adjustment the investment cost has been increased by a factor in order to consider the escalation and update the costs to today figures.
The resulting figure is the direct materials cost.

GEE provided the cost of all the equipment, bulk materials and labour for reference cases in a study made in year 2003 with IEA GHG and FWI, based on the same coal and gasification configuration as in the present study.
The direct materials cost was taken out and used as the basis for FWI's estimate of overall investment cost.
As per Shell cases, the direct materials have been adjusted based on the actual coal flowrate resulting from finalization of the IGCC performances.

Also for GEE case an escalation factor has been applied. Nonetheless, the offer received from GEE is less updated with respect to the Shell one and as a consequence a higher escalation factor has been applied.
For Siemens case, FWI used investment cost data of the main equipment for reference cases provided by Siemens in a study made in year 2005 with IEA GHG and FWI, based on different coal. The figure has been adjusted to the specific case by FWI based on in house data and with the support of FWI estimate department.

Besides all this consideration, the basic cost data, both requested in this report and older ones, have been provided directly by the vendors and they not have been commented by FWI.

Process Packages: Unit 2100 (Air Separation Unit), Unit 2400 (Sulphur Recovery and Tail Gas Treatment) and Unit 2600 (H_2 Production)

Unit 2100 (Air Separation Unit), Unit 2400 (Sulphur Recovery and Tail Gas Treatment) and Unit 2600 (H_2 Production) are Process Packages. The investment cost is derived from competitive bids received and technically evaluated by FW in the past for similar projects.
For each alternative the figure taken as a reference has been adjusted on the basis of electric power consumption and Oxygen production (for ASU), syngas feed and sulphur production (for SRU & TGT) and H_2 production (for Unit 2600).

Unit 2200 (Syngas Cooling and Conditioning Line) and 2300 (Acid Gas Removal)

Investment costs for these units are derived from previous studies that Foster Wheeler made for IEA GHG, by using suitable parameters like syngas flowrate and characteristics.

Unit 2500 (CO_2 Compression and Drying)

Direct materials cost of CO_2 compressors and drivers is based on a budget quotation received from qualified Vendors. Costs of other equipment are derived from in house data.

Unit 3000 (Power Island)

The direct materials cost is based on competitive bids received in the recent past for similar equipment (gas turbine, HRSG, steam turbine) and on proprietary software output for other equipment and bulk materials.

Unit 4000 to 5200 (Utilities and Offsite)

Cost of each Unit is evaluated based on in house data for similar Units and adjusted on the basis of Unit Capacity.
These units also include DCS, ESD, EMS, Electrical Systems and HV substation.

2.2.2 Construction, Other Costs and EPC Services

Per each Unit (if necessary, for each Technology), or block of Units, the remaining costs (i.e. Construction, Other Costs and EPC Services) are calculated multiplying the cost of direct materials by factors, built up by FW from statistics based on cost estimates of similar plants.

2.2.3 Contingencies

The estimating contingency is a provisional sum that will give to an estimate equal chance of overrun or underrun within certain limits and it is meant to cover:
- estimating errors
- estimating omissions

Contingency is included in the estimate as a percentage of the estimated costs on the basis of:
- definition of the technical documentation in term of quality and completeness;
- estimate quality;
- methodology adopted to develop the estimate.

Different percentages of contingency are applied to the different sections on the basis of historical data.

2.2.4 Estimate Currencies

The estimate was developed in Euro.
The following exchange Euro to US $ rate has been used:

1.25 US $ equivalent to 1 Euro.

2.2.5 Inflation

No escalation is applied to the estimated installed cost.

2.2.6 Miscellanea Costs

Land purchase, surveys and general site preparation are taken into account at a cost equal to 5% of the installed plant cost.
Additional costs for process/patent fees, fees for agents and consultants, legal and planning activities, are taken into account at a cost equal to 2% of the installed plant cost.

The sum of the installed plant cost plus the miscellanea costs is the Total Investment Cost.

2.3 Estimate Accuracy

The estimate accuracy is within the range +/- 30%.

3.0 <u>Investment Cost of the Alternatives</u>

3.1 GEE alternative (Case 0.A)

The following Table E.3.1 shows the investment break down and the total figures for the GEE alternative.

FOSTER WHEELER

Table E.3.1 - ESTIMATE SUMMARY

GEE CASE 0A

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		UNIT 900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	2600 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
1	DIRECT MATERIALS		9,950,850	184,602,600	126,487,998	45,144,918	41,574,078	29,484,000	28,386,540	8,000,000	213,595,000	161,263,788	848,489,772	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		1,505,400	62,835,900	28,703,046	16,939,600	14,693,300	10,035,900	5,368,600	4,000,000	40,391,700	67,090,500	251,563,946	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		727,800	20,235,300	3,697,341	9,040,900	14,265,800	3,016,400	1,037,800	480,000	15,609,000	11,784,000	79,894,341	
4	EPC SERVICES		1,090,500	47,215,700	13,865,031	10,149,400	7,178,400	3,231,900	1,452,000	3,520,000	12,487,000	23,570,000	123,759,931	
A	**Installed Costs** (Contingency excluded)		**13,274,550**	**314,889,500**	**172,753,416**	**81,274,818**	**77,711,578**	**45,768,200**	**36,244,940**	**16,000,000**	**282,082,700**	**263,708,288**	**1,303,707,990**	
B	Contingency	%	7	7	5	7	7	7	5	7	7	5	6.3	
		Euro	929,219	22,042,265	8,637,671	5,689,237	5,439,810	3,203,774	1,812,247	1,120,000	19,745,789	13,185,414	81,805,426	
C	Fees (2% of A)		265,491	6,297,790	3,455,068	1,625,496	1,554,232	915,364	724,899	320,000	5,641,654	5,274,166	26,074,160	
D	Land Purchases; surveys (5% of A)		663,728	15,744,475	8,637,671	4,063,741	3,885,579	2,288,410	1,812,247	800,000	14,104,135	13,185,414	65,185,399	
	TOTAL INVESTMENT COST		**15,132,987**	**358,974,030**	**193,483,826**	**92,653,293**	**88,591,199**	**52,175,748**	**40,594,333**	**18,240,000**	**321,574,278**	**295,353,283**	**1,476,772,976**	

900 Coal Handling & Storage
1000 Gasification Section
2100 Air Separation Unit
2200 Syngas Treat.&Condt. Line
2300 Acid Gas Removal
2400 SRU & TGT
2500 CO2 Compression&Drying
2600 Hydrogen production unit
3000 Power Island
4000+ Utilities&Offsites

3.2 Shell alternative (Case 0.B)

The following Table E.3.2 shows the investment break down and the total figures for the Shell alternative.

FOSTER WHEELER

Table E.3.2 - ESTIMATE SUMMARY

SHELL CASE 0B

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		UNIT 900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	2600 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
1	DIRECT MATERIALS		40,041,100	137,377,000	102,096,540	29,827,980	56,040,894	24,727,248	24,871,392	8,009,500	169,134,000	146,343,325	738,468,979	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		12,913,100	62,118,825	23,168,061	10,893,300	26,118,100	8,416,500	4,703,200	4,004,750	38,238,800	60,883,100	251,457,736	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		2,256,900	7,454,047	2,984,360	15,242,600	24,007,000	2,529,300	908,500	480,570	15,295,000	10,694,000	81,852,278	
4	EPC SERVICES		6,008,900	31,058,883	11,191,352	6,524,800	12,181,900	2,710,000	1,272,400	3,524,180	12,237,000	21,389,000	108,098,414	
A	**Installed costs** (contingency excluded)		**61,220,000**	**238,008,755**	**139,440,313**	**62,488,680**	**118,347,894**	**38,383,048**	**31,755,492**	**16,019,000**	**234,904,800**	**239,309,425**	**1,179,877,407**	
B	Contingency	%	7	7	5	7	7	7	5	7	7	5	6.3	
		Euro	4,285,400	16,660,613	6,972,016	4,374,208	8,284,353	2,686,813	1,587,775	1,121,330	16,443,336	11,965,471	74,381,314	
C	Fees (2% of A)		1,224,400	4,760,175	2,788,806	1,249,774	2,366,958	767,661	635,110	320,380	4,698,096	4,786,189	23,597,548	
D	Land Purchases; surveys (5% of A)		3,061,000	11,900,438	6,972,016	3,124,434	5,917,395	1,919,152	1,587,775	800,950	11,745,240	11,965,471	58,993,870	
	TOTAL INVESTMENT COST		**69,790,800**	**271,329,981**	**156,173,150**	**71,237,095**	**134,916,599**	**43,756,675**	**35,566,151**	**18,261,660**	**267,791,472**	**268,026,556**	**1,336,850,140**	

900 Coal Handling & Storage
1000 Gasification Section
2100 Air Separation Unit
2200 Syngas Treat.&Condt. Line
2300 Acid Gas Removal
2400 SRU & TGT
2500 CO2 Compression&Drying
2600 Hydrogen production unit
3000 Power Island
4000+ Utilities&Offsites

3.3 Siemens alternative (Case 0.C)

The following Table E.3.3 shows the investment break down and the total figures for the Siemens alternative.

FOSTER WHEELER

Table E.3.3 - ESTIMATE SUMMARY

SIEMENS CASE 0C

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		UNIT 900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	2600 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
1	DIRECT MATERIALS		47,465,015	86,393,000	107,914,000	30,532,320	58,467,000	26,222,000	26,346,000	7,962,000	201,671,000	143,876,061	736,848,396	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		18,185,566	31,861,479	25,084,304	11,150,131	27,248,797	8,925,274	4,982,050	3,981,000	38,136,822	59,856,700	229,412,122	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		3,116,400	10,523,057	3,231,198	15,602,203	25,046,304	2,682,195	962,364	477,720	14,737,623	10,513,000	86,892,065	
4	EPC SERVICES		8,645,952	24,553,800	12,116,994	6,678,868	12,709,275	2,873,818	1,347,840	3,503,280	11,789,910	21,028,400	105,248,137	
A	**Installed Costs** (Contingency excluded)		**77,412,933**	**153,331,336**	**148,346,497**	**63,963,521**	**123,471,376**	**40,703,288**	**33,638,254**	**15,924,000**	**266,335,355**	**235,274,161**	**1,158,400,720**	
B	Contingency	%	7	7	5	7	7	7	5	7	7	5	6.3	
		Euro	5,418,905	10,733,194	7,417,325	4,477,446	8,642,996	2,849,230	1,681,913	1,114,680	18,643,475	11,763,708	72,742,872	
C	Fees (2% of A)		1,548,259	3,066,627	2,966,930	1,279,270	2,469,428	814,066	672,765	318,480	5,326,707	4,705,483	23,168,014	
D	Land Purchases; surveys (5% of A)		3,870,647	7,666,567	7,417,325	3,198,176	6,173,569	2,035,164	1,681,913	796,200	13,316,768	11,763,708	57,920,036	
	TOTAL INVESTMENT COST		**88,250,743**	**174,797,723**	**166,148,076**	**72,918,414**	**140,757,369**	**46,401,748**	**37,674,844**	**18,153,360**	**303,622,305**	**263,507,061**	**1,312,231,643**	

900	**Coal Handling & Storage**
1000	**Gasification Section**
2100	**Air Separation Unit**
2200	**Syngas Treat.&Condt. Line**
2300	**Acid Gas Removal**
2400	**SRU & TGT**
2500	**CO2 Compression&Drying**
2600	**Hydrogen production unit**
3000	**Power Island**
4000+	**Utilities&Offsites**

4.0 Operation and Maintenance Cost of the Alternatives

Operating and Maintenance (O&M) costs include:

- Feedstock
- Chemicals
- Catalysts
- Solvents
- Raw Water make-up
- Direct Operating labour
- Maintenance
- Overhead Charges

O&M costs are generally allocated as variable and fixed costs.
Variable operating costs are directly proportional to the amount of kilowatt-hours and Hydrogen produced and are referred as incremental costs.
Fixed operating costs are essentially independent of the amount of products.
However, accurately distinguishing the variable and fixed operating costs is not always simple. Certain cost items may have both, variable and fixed, components; for instance the planned maintenance and inspection of the gas turbine, that are known to occur based on number of running hours.
In this study these costs have been considered fixed, assuming that the complex operates at design capacity and with the expected design service factor.

4.1 Variable Costs

The consumption of the various items and the corresponding costs are yearly, based on the expected equivalent availability of 7446 equivalent hours of operation in one year with syngas. Another 554 equivalent hours of operation of the power plant in one year with natural gas as back-up fuel is expected, provided the resulting greenhouse gas emissions are acceptable but conservatively this has not been considered in the economical analysis.

The following Table E.4.1/2/3 show the total yearly operating costs for the three alternatives.

FOSTER WHEELER

Client : IEA GHG

Date July 2007 Rev.1

Table E.4.1 - GEE Case 0A Yearly Variable Costs

Yearly Operating hours =	7446	GEE - Case 0A		
Consumables	**Unit Cost**	**Consumption**		**Oper. Costs**
		Hourly	Yearly	**(yearly basis)**
	Euro/t	kg/h	t/y	
Feedstock				
Coal	31.0	323,100	2,405,803	74,579,881
Auxiliary feedstock				
Natural Gas (Flare)	113.0	80	595.7	67,312
Make-up water	0.100	315,000	2,345,490	234,549
Solvents				
Selexol	6500	16.76	124.8	811,200
Catalyst				998,119
Chemicals				2,046,515
Waste Disposal	7.0	101,400	755,024	5,285,171
TOTAL YEARLY OPERATING COSTS, Euro/year				**84,371,963**

FOSTER WHEELER		Client	: IEA GHG	
		Date	July 2007 Rev.1	

Table E.4.2 - Shell Case 0B Yearly Variable Costs

Yearly Operating hours =	7446	Shell - Case 0B		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	 Yearly t/y	**Oper. Costs (yearly basis)**
Feedstock				
Coal	31.0	273,100	2,033,503	63,038,581
Flux	15.0	8,340	62,097	931,461
Auxiliary feedstock				
Natural Gas (Flare)	113.0	75	558	63,105
Make-up water	0.1	406,000	3,023,076	302,308
Solvents				
Selexol	6500	16.76	124.8	811,200
Catalyst				1,683,899
Chemicals				1,315,364
Waste Disposal	7.0	40,500	301,563	2,110,941
TOTAL YEARLY OPERATING COSTS, Euro/year				**70,256,858**

FOSTER WHEELER		Client : IEA GHG Date July 2007 Rev.1		
Table E.4.3 - Siemens Case 0C Yearly Variable Costs				
Yearly Operating hours =	7446	**Siemens - Case 0C**		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	 Yearly t/y	**Oper. Costs (yearly basis)**
Feedstock				
Coal	31.0	295,500	2,200,293	68,209,083
Auxiliary feedstock				
Natural Gas (1)	113.0	2,420	18,019	2,036,183
Make-up water	0.1	315,000	2,345,490	234,549
Solvents				
Selexol	6500.0	17.60	131.0	851,760
Catalyst				1,702,595
Chemicals				2,006,519
Waste Disposal	7.0	58,410	434,921	3,044,446
TOTAL YEARLY OPERATING COSTS, Euro/year				**78,085,135**

4.2 Fixed Costs

The fixed costs of the different Power Plants operation include the following items:
- Direct labour.
- Administrative and general overhead.
- Maintenance.

For maintenance, variable elements of the cost, such as gas turbine inspections, have been treated as part of the fixed costs, on the assumption that the Complex operates at the design capacity and with the expected design service factor.

4.2.1 Direct Labour

The yearly cost of the direct labour is calculated assuming, for each individual, an average cost equal to 50,000 Euro/year. The number of personnel engaged for the different alternatives is hereinafter.

The Owner's personnel engaged in the Operation and Maintenance of the IGCC Complex is shown in Table E.4.4. The Complex has been divided into 3 areas of operation: Air Separation Unit, Gasification, including syngas processing and CO_2 capture plant, and Power Island with common Utilities. The same division will be reflected in the design of the centralized Control Room, which will have, correspondingly, 3 main DCS control groups, each one equipped with a number of control stations, from where the operation of the units of each of the three areas will be controlled.

The Area Responsible and his Assistant will supervise each area of operation; both are daily positions. The Shift Superintendent and the Electrical Assistant are common for the 3 areas; both are shift positions. The rest of the Operation staff is structured around the standard positions: shift supervisors, control room operators and field operators.

The maintenance personnel are based on large use of external subcontractors for all medium-major type of maintenance work. Maintenance costs described at para. 4.2.3 take into account the service outsourcing. Plant Maintenance personnel, like the instrument specialists, perform routine maintenance and resolve emergency problems.

Table E.4.4 – IGCC personnel.

OPERATION	ASU	GASIFICATION	CCU & UTILITIES	TOTAL	NOTES
Area Responsible	1	1	1	3	daily position
Assistant Area Responsible	1	1	1	3	daily position
Shift Superintendent		5		5	1 shift position
Electrical Assistant		5		5	1 shift position
Shift Supervisor	5	5	5	15	3 shift position
Control Room Operator	5	10	10	25	5 shift position
Field Operator	5	25	20	50	10 shift position
Subtotal				**106**	
MAINTENANCE					
Mechanical group		4		4	daily position
Instrument group		7		7	daily position
Electrical group		5		5	daily position
Subtotal				**16**	
LABORATORY					
Superintendent + Analysts		6		**6**	daily position
TOTAL				**128**	

4.2.2 Administrative and General Overheads

All other Company services not directly involved in the operation of the Complex fall in this category, such as:

- Management.
- Administration.
- Personnel services.
- Technical services.
- Clerical staff.

These services vary widely from company to company and are also dependent on the type and complexity of the operation.
Based on EPRI, Technical Assessment Guide for the Power Industry, an amount equal to 30% of the direct labour cost has been considered.

4.2.3 Maintenance

A precise evaluation of the cost of maintenance would require a breakdown of the costs amongst the numerous components and packages of the Complex. Since these costs are all strongly dependent on the type of equipment selected and statistical

maintenance data provided by the selected Vendors, this type of evaluation of the maintenance cost is premature at this stage of the study.
For this reason the annual maintenance cost of the Complex has been estimated, as suggested by EPRI Technical Evaluation Guide, as a percentage of the installed capital cost of the facilities.
In accordance with EPRI recommendations the Complex has been divided into four major sections, applying to each section different percentages of the capital cost of the section to determine the relative cost of maintenance, as shown in the attached tables.
The percentage applied to the Power Island has been adjusted to take into account the gas turbine maintenance cost based on the assumption of a Long Term Service Agreement (LTSA) with the gas turbine manufacturer.
The total yearly maintenance cost of the Complex is assumed to be subcontracted to external firms under the supervision of the maintenance staff of the Owner, included in the fixed cost as direct labour.
The overall cost of maintenance can be statistically split as follows:

- maintenance materials : 60% of total maintenance cost;
- maintenance labour : 40% of total maintenance cost.

The attached table E.4.5 shows the total maintenance costs for the three alternatives.

FOSTER WHEELER	E.4.5 - Maintenance Costs						Client : IEA GHG Date : July 2007	
		GEE - Case 0A		Shell - Case 0B		Siemens - Case 0C		
Complex section	**Maintenance** %	**Capital Cost** Euro x 10^3	**Maintenance** 10^3 Euro/Year	**Capital Cost** Euro x 10^3	**Maintenance** 10^3 Euro/Year	**Capital Cost** Euro x 10^3	**Maintenance** 10^3 Euro/Year	
ASU, AGR, SRU & TGT, CO_2 Comp., Coal St, H2 prod (Units: 900, 2100, 2300, 2400, 2500, 2600)	2.5	361,753	9,044	405,166	10,130	439,496	10,987	
Gasification, Syngas Treat., (Units:1000,2200)	4.0	396,164	15,847	300,497	12,020	217,295	8,692	
Power Island (Unit: 3000)	5.0 (1)	282,083	14,104	234,905	11,745	266,335	13,317	
Common facilities (Utilities, Offsite, etc.)	1.7	263,708	4,483	239,309	4,068	235,274	4,000	
TOTAL		1,303,708	43,477	1,179,877	37,964	1,158,401	36,995	
		Maint. % =	3.3	Maint. % =	3.2	Maint. % =	3.2	

NOTES: (1) Including the Gas Turbine Long Term Service Agreement.

4.3 Summary

The following table summarizes the total Operating and Maintenance Costs on a yearly basis for all the alternatives.

Table E.4.6 – Total O&M Costs

		GEE Case 0A Euro/year	Shell Case 0B Euro/year	Siemens Case 0C Euro/year
Fixed Costs	direct labour	6,400,000	6,400,000	6,400,000
	adm./gen overheads	1,920,000	1,920,000	1,920,000
	maintenance	43,477,000	37,964,000	36,995,000
	Subtotal	51,797,000	46,284,000	45,315,000
Variable Costs		84,371,963	70,256,858	78,085,000
TOTAL O&M COSTS		136,168,963	116,540,858	123,400,000

5.0 Evaluation of the Electric Power Cost of the Alternatives

5.1 Electric Power Cost

The following Tables summarize the economic analyses performed on each alternative in order to evaluate the electric power production cost, based on the following main assumptions:

- 7446 equivalent operating hours in normal conditions at 100% capacity;
- Total investment cost as evaluated in para.3.0 of this Section;
- O&M costs as evaluated in para 4.0;
- 10% discount rate on the investment cost over 25 operating years;
- No selling price is attributed to CO_2;
- Other financial parameters as per Project Design Basis, Section B.

Table E.5.1 summarizes the electric power cost for the three alternatives, with 10% discount rate applied on the Total Investment Cost.

Tables E.5.2/3/4 show the cash flow detailed calculation.

Table E.5.1– Electric Power Cost

ALTERNATIVE		**0A GEE**	**0B Shell**	**0C Siemens**
Coal Flow Rate	t/h	323.1	273.1	295.3
Net Power Output	MWe	390.8	317.1	326.8
Hydrogen Production	MWe equiv	334.9	333.3	331.4
Total Investment Cost	MM Euro	1476.8	1336.9	1312.2
Revenues /year	MM Euro/year	350.9	310.9	314.3
Electricity prod Cost	Euro/kWh	0.071	0.071	0.071

Table E.5.2 - GEE CASE 0A - Cost Evaluation - Discount Rate = 10%

Production		
Coal Florate	323.1	t/h
Net Power Output	390.8	MW
Sold Sulphur	2.78	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	200,510	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1303.7
Land purchase; surveys	5%	65.2
Fees	2%	26.1
Average Contingencies	6.3%	81.8
Total Investment Cost		1476.8

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	74.6
Maintenance	43.5
Waste Disposal (7€/t)	5.6
Chemicals + Consumable	4.2
Insurance and local taxes	26.1

Working Capital	**MM Euro**
30 days Chemical Storage	0.4
30 days Coal Storage	7.2
Total Working capital	7.6

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.071	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	350.9	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electric Energy				109.5	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9
Sulphur				1.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1
Hydrogen				75	142	142	142	142	142	142	142	142	142	142
Operating Costs														
Fuel Cost				-39.5	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6
Maintenance				-29.0	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-2.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2
Waste Disposal				-3.0	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6
Insurance				-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1
Working Capital Cost				-7.6										
Fixed Capital Expenditures	-295.4	-664.5	-516.9											
Total Cash flow (yearly)	-295.4	-664.5	-516.9	70.1	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6
Total Cash flow (cumulated)	-295.4	-959.9	-1476.8	-1406.7	-1218.0	-1029.4	-840.7	-652.1	-463.4	-274.8	-86.1	102.5	291.2	479.8
Discounted Cash Flow (Yearly)	-268.5	-549.2	-388.3	47.9	117.1	106.5	96.8	88.0	80.0	72.7	66.1	60.1	54.6	49.7
Discounted Cash Flow (Cumul.)	-268.5	-817.7	-1206.0	-1158.2	-1041.0	-934.5	-837.7	-749.7	-669.7	-597.0	-530.9	-470.8	-416.1	-366.4

CASH FLOW ANALYSYS Millions Euro	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**
	12	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electric Energy	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	206.9	
Sulphur	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	2.1	
Hydrogen	142	142	142	142	142	142	142	142	142	142	142	142	142	142	
Operating Costs															
Fuel Cost	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	-74.6	
Maintenance	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	-43.5	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	-4.2	
Waste Disposal	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	-5.6	
Insurance	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	
Working Capital Cost															7.6
Fixed Capital Expenditures															
Total Cash flow (yearly)	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	188.6	7.6
Total Cash flow (cumulated)	668.5	857.1	1045.8	1234.4	1423.1	1611.7	1800.4	1989.0	2177.7	2366.3	2555.0	2743.6	2932.3	3120.9	3128.5
Discounted Cash Flow (Yearly)	45.2	41.1	37.3	33.9	30.8	28.0	25.5	23.2	21.1	19.2	17.4	15.8	14.4	13.1	0.5
Discounted Cash Flow (Cumul.)	-321.3	-280.2	-242.9	-209.0	-178.1	-150.1	-124.6	-101.4	-80.3	-61.2	-43.8	-28.0	-13.6	-0.5	0.0

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Table E.5.3 - SHELL CASE 0B - Cost Evaluation - Discount Rate = 10%

Production		
Coal Florate	273.1	t/h
Net Power Output	317.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	200,860	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1179.9
Land purchase; surveys	5%	59.0
Fees	2%	23.6
Average Contingencies	6.3%	74.4
Total Investment Cost		1336.85

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	63.0
Maintenance	38.0
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	23.6

Working Capital	**MM Euro**	
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6
Labour Cost	**MM Euro/year**	
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.071	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	310.9	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				88.4	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen				75	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	
Operating Costs																													
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance				-25.3	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance				-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	
Working Capital Cost				-6.6																									6.6
Fixed Capital Expenditures	-267.4	-601.6	-467.9																										
Total Cash flow (yearly)	-267.4	-601.6	-467.9	63.6	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	6.6
Total Cash flow (cumulated)	-267.4	-869.0	-1336.9	-1273.3	-1102.5	-931.7	-761.0	-590.2	-419.4	-248.7	-77.9	92.9	263.6	434.4	605.1	775.9	946.7	1117.4	1288.2	1459.0	1629.7	1800.5	1971.3	2142.0	2312.8	2483.6	2654.3	2825.1	2831.7
Discounted Cash Flow (Yearly)	-243.1	-497.2	-351.5	43.4	106.0	96.4	87.6	79.7	72.4	65.8	59.9	54.4	49.5	45.0	40.9	37.2	33.8	30.7	27.9	25.4	23.1	21.0	19.1	17.3	15.8	14.3	13.0	11.8	0.4
Discounted Cash Flow (Cumul.)	-243.1	-740.2	-1091.8	-1048.3	-942.3	-845.9	-758.3	-678.6	-606.2	-540.4	-480.5	-426.1	-376.6	-331.7	-290.8	-253.6	-219.9	-189.1	-161.2	-135.8	-112.8	-91.8	-72.7	-55.4	-39.6	-25.3	-12.3	-0.4	0.0

Table E.5.4 - Siemens CASE 0C - Cost Evaluation - Discount Rate = 10%

Production		
Coal Florate	295.3	t/h
Net Power Output	326.8	MW
Sold Sulphur	2.25	t/h
Fuel Price	31.0	USD/t (*)
Insurance and local taxes	2%	Installed cost
Hydrogen production	198,500	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures	**MM Euro**	
Installed Costs		1158.4
Land purchase; surveys	5%	57.9
Fees	2%	23.2
Average Contingencies	6.3%	72.7
Total Investment Cost		1312.2

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	68.2
Maintenance	37.0
Waste Disposal (7€/t)	3.0
Chemicals + Consumable	6.9
Insurance and local taxes	23.2

Working Capital	**MM Euro**	
30 days Chemical Storage		0.7
30 days Coal Storage		6.6
Total Working capital		7.3
Labour Cost	**MM Euro/year**	
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.071	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	314.3	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electric Energy				91.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1
Sulphur				0.9	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
Hydrogen				74	140	140	140	140	140	140	140	140	140	140
Operating Costs														
Fuel Cost				-36.1	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2
Maintenance				-24.7	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-3.6	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9
Waste Disposal				-1.6	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0
Insurance				-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2
Working Capital Cost				-7.3										
Fixed Capital Expenditures	-262.4	-590.5	-459.3											
Total Cash flow (yearly)	-262.4	-590.5	-459.3	61.6	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7
Total Cash flow (cumulated)	-262.4	-853.0	-1312.2	-1250.6	-1082.9	-915.2	-747.5	-579.8	-412.1	-244.4	-76.7	91.0	258.7	426.4
Discounted Cash Flow (Yearly)	-238.6	-488.0	-345.1	42.1	104.1	94.7	86.1	78.2	71.1	64.7	58.8	53.4	48.6	44.2
Discounted Cash Flow (Cumul.)	-238.6	-726.6	-1071.7	-1029.6	-925.4	-830.8	-744.7	-666.5	-595.4	-530.7	-471.9	-418.5	-369.9	-325.8

CASH FLOW ANALYSYS Millions Euro	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**
	12	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electric Energy	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	172.1	
Sulphur	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	
Hydrogen	140	140	140	140	140	140	140	140	140	140	140	140	140	140	
Operating Costs															
Fuel Cost	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	-68.2	
Maintenance	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	-6.9	
Waste Disposal	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	
Insurance	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	-23.2	
Working Capital Cost															7.3
Fixed Capital Expenditures															
Total Cash flow (yearly)	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	167.7	7.3
Total Cash flow (cumulated)	594.1	761.8	929.5	1097.2	1264.9	1432.6	1600.3	1768.0	1935.7	2103.4	2271.1	2438.8	2606.5	2774.2	2781.4
Discounted Cash Flow (Yearly)	40.1	36.5	33.2	30.2	27.4	24.9	22.7	20.6	18.7	17.0	15.5	14.1	12.8	11.6	0.5
Discounted Cash Flow (Cumul.)	-285.6	-249.1	-216.0	-185.8	-158.4	-133.4	-110.8	-90.2	-71.5	-54.4	-38.9	-24.9	-12.1	-0.5	0.0

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME	:	HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME	:	COMPARISON OF ALTERNATIVES AND SELECTION OF THE MOST PROMISING TECHNOLOGY

ISSUED BY	:	P. COTONE
CHECKED BY	:	P. COTONE
APPROVED BY	:	S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	P. Cotone	P. Cotone	S. Arienti
July 2007	Rev 1	L.Valota	P. Cotone	S. Arienti

SECTION F

COMPARISON OF ALTERNATIVES AND SELECTION OF THE BEST TECHNOLOGY

I N D E X

1.0 Introduction

The purpose of this section F is to present the performance and cost data developed for the alternatives studied in the previous sections, in order to show the major features and merits of each alternative.

From the first analysis of the table F.3.1, it is evident that the alternatives have approximately a similar net electrical efficiency, despite the differences of the various technologies involved. With reference to the production costs, the range of variation falls in a very tight range, although, there are differences in single factors (investment cost, operating costs, electric power output etc.).

The following paragraph presents a more detailed analysis of the different alternatives.

2.0 Alternatives comparison

This comparison is mainly aimed at evaluating the effect of the pre-combustion CO_2 capture and hydrogen co-production on different gasification technologies, by examining plant performances and investment/production cost data. The different gasification technologies are: GEE (Case 0A), Shell (Case 0B), Siemens (0C).

Table F.3.1 summarises the most important data of the alternatives.

Table F.3.1 – Performance data.

		Case 0A GEE Gasifier	Case 0B Shell Gasifier	Case 0C Siemens Gasifier
ACID GAS REMOVAL TECHNOLOGY		Selexol	Selexol	Selexol
CO_2 Capture Efficiency	**%**	84.8	85.1	84.9
OVERALL PERFORMANCES				
Coal Flow Rate A.R.	**t/h**	323.1	273.1	295.3
Coal LHV	**kJ/kg**	25,869.5	25,869.5	25869.5
Thermal Energy of Feedstock	**MWth**	2321.8	1962.5	2122.0
Actual Gross Electric power output	**MWe**	625.1	518.1	538.5
H_2 produced	**MWth**	598	599	591.8
Auxiliary Consumption	**MWe**	234.3	201	211.7
Actual Net Electric power output	**MWe**	390.8	317.1	326.8
Net Equivalent Electric Power Output	**MWe**	725.7	652.5	658.2
Gross Equivalent Electrical Efficiency	**%**	41.3	43.5	41.0
Hydrogen Equivalent electric power	**MWe**	334.9	335.4	331.4
Gross Equivalent Electric Power Output	**MWe**	960	853.5	869.9
Net Equivalent Electrical Efficiency	**%**	31.3	33.3	31.0
(H_2/effective EE) ratio	**MWt/MWe**	1.5	1.9	1.8
INVESTMENT COST DATA				
Total Investment	**10^6 €**	1476.8	1336.9	1312.2
Equivalent Specific Net Investment Cost	**€/kW**	2035	2049	1994
O&M Costs	**10^6 €/y**	136.2	116.5	123.4
PRODUCTION COST DATA				
C.O.E (DCF=10%)	**c€/kWh**	0.071	0.071	0.071

The main common features of the alternatives are a gasification pressure suitable to feed the gas turbines and the use of a Selexol scrubbing for the acid gas washing, with a separated removal of CO_2 and H_2S. For GEE case, the gasification pressure is higher (approx 65 barg) allowing electric power generation by an expander on the syngas line, downstream of the Acid Gas Removal unit.

3.0 Selection of the best technology

Shell has the higher equivalent net efficiency, resulting in lower coal consumption at the same nominal plant capacity.
This is mainly due to the following reasons:

- Gasifier efficiency of the Shell Technology is higher with respect both to GEE and Siemens, due to the different gasification technology: dry feed and WHB for Shell with respect to GEE gasification (based on wet feed and quench gasifier) and only the WHB with respect to Siemens gasification (dry feed and quench gasifier);
- Auxiliary power consumption of the Shell technology is lower than those of GEE and Siemens: the lower coal flowrate corresponds to lower oxygen consumption and therefore to lower ASU electric power consumption.

Siemens has the worst efficiency mainly because of the syngas composition which, having a higher CO/H_2 ratio than other technologies, requires heavy CO shift reaction with deterioration of syngas quality. Moreover, due to low pressure and composition of the syngas, the condensation of the water vapour content in the syngas flow occurs at low temperature (at VLP generator). For this reason in the syngas treatment a large amount of heat (latent heat) is available at low temperature that can be only partially recovered and used in the combined cycle, while the most part of it is discharge to the sea water cooling system. As a consequence, less steam generated at higher pressure with respect to the GEE and Shell Gasification technologies.

The O&M costs are affected by the efficiency (variable costs) and by the investment cost (for maintenance); the best mingling of the two components is for Shell technology, having the lowest O&M costs.

The investment cost of the equivalent kWh produced is in favour of Siemens (thanks to the lowest investment cost) followed by GEE and Shell.

All these parameters concur to the evaluation of the cost of electricity (COE, €/kWh), the figure taken to compare economically the three alternatives, at a fixed H_2 selling price (9.5 €cent/Nm^3) and 10% discount rate. The calculated COE for Shell, GEE and Siemens are the same (0.071 €/kWh).

One of the main parameters affecting the COE is the investment cost. FWI has derived the cost for the Shell gasification unit from the previous 2005 study, while for GEE from the previous 2003 study. Siemens costs have been derived by FWI based on data provided by Siemens in similar study and finally approved by the supplier. These aspects make FWI more confident of evaluation on Shell.

In the attached table F.3.1 is shown also the ratio H_2 production and Electric Energy production. Shell and Siemens technologies appear the most suitable to match the Netherlands ratio evaluated in Section J (Attachment A), reflecting the future hypothetical hydrogen based economy in Europe. GEE would be more suitable for the USA.

Moreover it can be noted that Shell and GEE gasification technologies have more operating plants than Siemens.

Finally the Shell gasifier presents a higher efficiency and as consequence lower CO_2 production and a lower CO_2 storage cost.

These considerations lead to a slight preference for Shell gasification. Therefore, the study of Hydrogen and Electricity co-production will be performed based on Shell gasification technology.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME: HYDROGEN AND ELECTRICITY COPRODUCTION - BASIC INFORMATION FOR EACH ALTERNATIVE

ISSUED BY : L. VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L.Valota	P. Cotone	S. Arienti

SECTION G

HYDROGEN AND ELECTRICITY COPRODUCTION BASIC INFORMATION FOR EACH ALTERNATIVE

INDEX

SECTION G

0.0 Introduction

The scope of this section is the technical description of five different co-production plants. All the plants are based on Shell gasification technology described in Section D.2.

The five co-production plants are the following:

Case 1: w/o CO_2 capture, w/o H_2 production
Case 2: with CO_2 capture, w/o H_2 production
Case 3: with CO_2 capture, with maximum H_2 production
Case 4: with CO_2 capture, with H_2 production, with optimum H_2/Electric Energy ratio
Case 5: with CO_2 capture, with H_2 production, with flexible H_2/EE ratio

The economical comparison is carried out in section H.

Case 1 is taken for reference and consists of an only electric energy production plant, without hydrogen production and without CO_2 capture (Section G1).

Case 2 consists of a co-production plant with the maximum electric energy production, without hydrogen production, with CO_2 capture (Section G2).

Case 3 consists of a co-production plant with the maximum hydrogen production and electric energy production only for internal electrical consumption (Section G3).

Case 4 consists of a co-production plant, with electricity and hydrogen production at a specific ratio and with CO_2 capture (Section G4). The plant has the same configuration as case D2. This is due to the fact that case G4 has to meet the same H_2/EE ratio as required (as an average) by the Netherlands and such value is approximately the same as shown in section D2.

Case 5 consists of a flexible coproduction plant with electricity and hydrogen production with CO_2 capture (Section G5).

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME	:	HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME	:	CASE 1: PLANT W/O H_2 PRODUCTION, W/O CO_2 CAPTURE

ISSUED BY	:	L.VALOTA
CHECKED BY	:	P.COTONE
APPROVED BY	:	S. ARIENTI

Date	**Revised Pages**	**Issued by**	**Checked by**	**Approved by**
April 2007 July 2007	Draft Rev 1	L. Valota L.Valota	P. Cotone P. Cotone	S. Arienti S. Arienti

SECTION G.1 HYDROGEN AND ELECTRICITY COPRODUTION BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION G.1 BASIC INFORMATION FOR EACH ALTERNATIVE

1.0 Case 1

1.1 Introduction

The main features of the Case 1 configuration of the IGCC Complex are:

- Low pressure (36 bar g) Shell Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Heat Recovery Type;
- No Shift and CO_2 removal.

The removal of acid gas (AGR) is based on DOW-UCARSOL process (activated MDEA solvent).
The degree of integration between the Air Separation Unit (ASU) and the Gas Turbines is 50%. Gas Turbine power augmentation and syngas dilution, for NO_x control, is achieved with injection of compressed moisturised N_2 from the ASU to the gas turbines.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

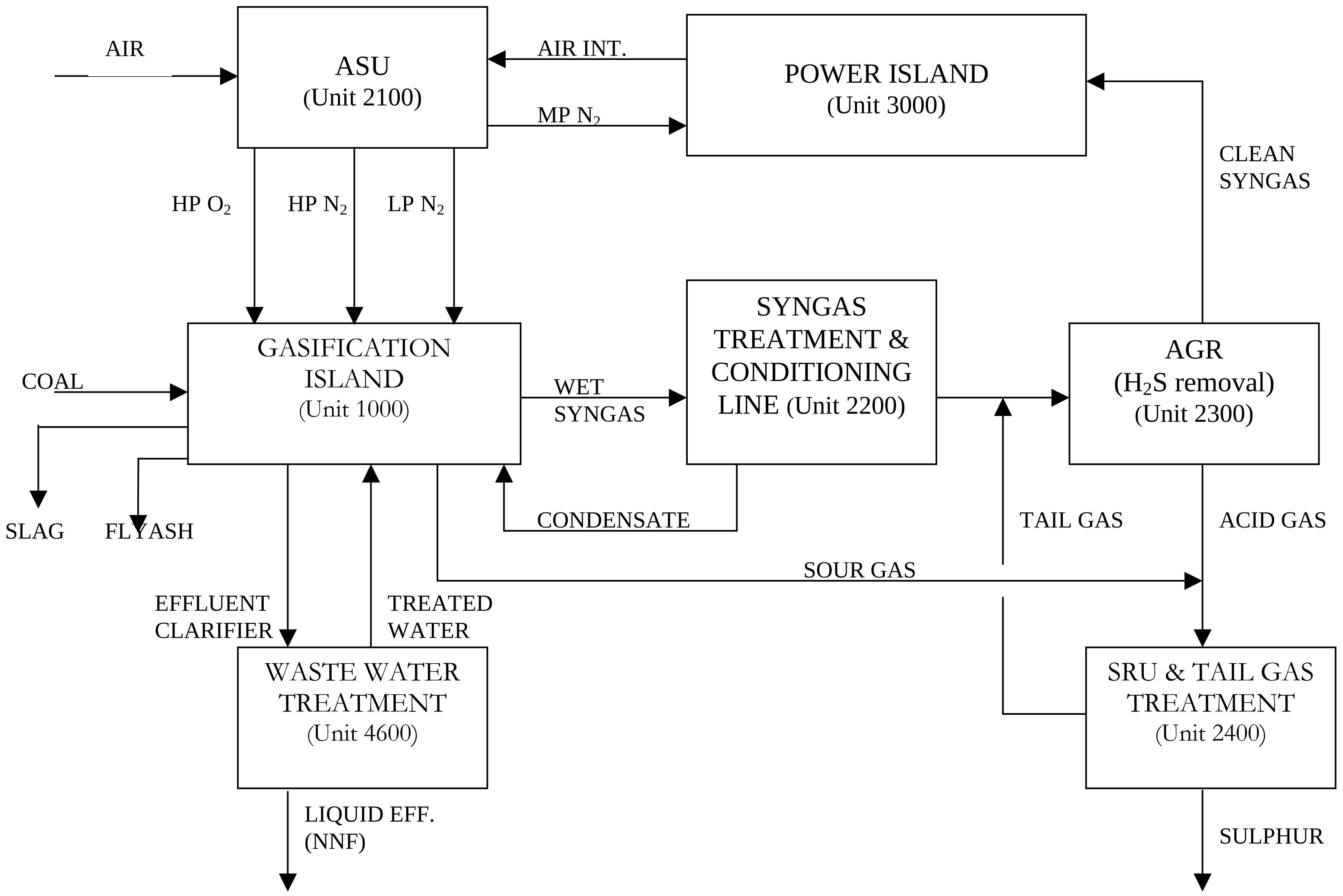
The arrangement of the process units is :

Unit		Trains
1000	Coal milling and drying	4 x 33 %
	Coal pressurization/feeding	6 x 20 %
	Gasification heat recovery	2 x 50 %
	Slag removal	2 x 50 %
	Dry solids removal	2 x 50 %
	Wet scrubbing	2 x 50 %
	Sour slurry and sour water stripper	1 x 100 %
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%
2300	AGR	1 x 100%
2400	SRU	2 x 100%
	TGT	1 x 100%

3000	Gas Turbine (PG – 9351 – FA)	2 x 50%
	HRSG	2 x 50%
	Steam Turbine	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.

CASE 1 – SHELL - IGCC COMPLEX BLOCK FLOW DIAGRAM



1.2 Process Description

Unit 1000: Gasification Island

Information relevant to the Shell Gasification Island are collected in para 1.1 of Section C.
The main process data of the Gasification Island relevant to this alternative are summarised in the following table:

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	80	80	70	126
Pressure (bar)		40	69	7.5	34
TOTAL FLOW					
Mass flow (kg/h)	250,600	196,980	82,000	31,800	463,500
Molar flow (kmol/h)			2,920	1,132	23,260
Composition (% vol)					
H_2					29.70
CO					56.40
CO_2					1.40
N_2		3.5	99.88	99.88	4.53
Ar		1.5	0.08	0.08	0.70
O_2		95	0.04	0.04	0.00
H_2S + COS					0.26
H_2O					7.00
Others					0.01

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package unit supplied by specialised Vendors. Reference is made to Section C, para. 3.0 for a general description of the Air Separation Unit.

The degree of integration with the gas turbines is 50% and the N_2 used to augment the power of the gas turbine and control the NOx is moisturised by direct contact with hot water in order to increase the syngas diluent mass flow.

The main process data and the main consumption of the ASU are summarised in following tables.

Process Data	Mass Flow (kg/h)	
Air from ambient	450,000	
Air from GT	450,000	
Oxygen to gasifier (95% vol)	1 97,000	
LP Nitrogen to Gasification Island (98% vol)	32,000	
HP Nitrogen to Gasification Island (98% vol)	82,000	
Nitrogen to Power Island (for syngas dilution)	575,000	
Consumption		
Main air compressor	40,000	kW
Oxygen compressor	10,800	kW
Nitrogen compressor	42,700	kW
Miscellanea	1,000	kW
Total	94,500	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 33 barg and 126°C enters Unit 2200. The syngas is first preheated, with the hydrolysis effluent, and then with MP steam, before entering the hydrolysis reactor, which converts COS to H_2S. The effluent is cooled against cold condensate. Process condensate separated is recycled to Unit 1000 Gasification while cold syngas is sent to Unit 2300 AGR.
Up to this point Unit 2200 is split in two parallel lines, each sized for 50% capacity.
Clean syngas, returning from Unit 2300, after removal of H_2S, is preheated with LP steam in E-2204 and sent to the gas turbines of Unit 3000.

Unit 2300: Acid Gas Removal (AGR)

Unit 2300 utilises the DOW-UCARSOL solvent (activated MDEA) as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (29 barg), and a low CO_2/H_2S ratio (5.5/1). As UOP/DOW see this separation as relatively easy, only an UCARSOL chemical wash has been proposed.

A single-stage absorption is suitable to accomplish all objectives, i.e. no acid gas enrichment is required. Therefore the tail gas coming from the Sulphur Recovery Unit is mixed with the raw syngas before entering the AGR section.

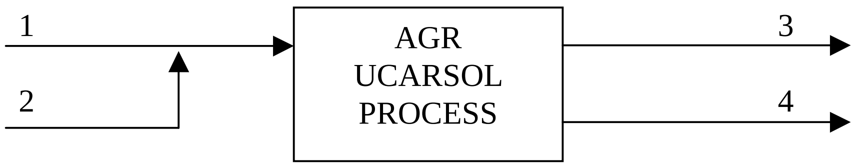
The interfaces of the Ucarsol process with the other Units are the following:

<u>Entering Streams</u>

1. Untreated Gas from Syngas Treatment & Conditioning Unit
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit

<u>Exit Streams</u>

3. Treated Gas to Gas Turbines
4. Acid Gas to Sulphur Recovery Unit



The MDEA solvent consumption, to make-up losses, is 60 m^3/year.
The proposed process matches the process specifications with reference to H_2S+COS concentration of the treated gas exiting the unit and fed to the Combined Cycle Unit. The treated gas feeding the gas turbines has an H_2S+COS concentration of 18 ppm.

CO_2 slippage with respect to expansion through the gas turbine is virtually 100% and even CO_2 derived from the other minor acid streams fed to the SRU is recovered.

The acid gas H_2S concentration is 49% dry basis, more than suitable to feed the oxygen blown Claus process.

Unit 2400: SRU and TGT

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of 51.5 t/d normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 30 barg.

Unit 3000: Power Island

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

Imported

- MHP steam (70 barg) : Steam imported from Gasification section.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

Exported

- MP steam (40 barg): Steam exported to Syngas Treatment and Conditioning Line. Part of the required steam is also generated in the Sulphur Recovery Unit and in the Gasification Island.
- LP steam(6.5 barg): Steam exported to the following Process Units: Syngas Treatment and Conditioning Line, AGR, ASU, Utility and Offsite Unit. Most of the steam is used to heat the recirculation of the Saturator Tower to moisturise the nitrogen fed to the gas turbine.
- BFW: HP, MP, LP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate: All the condensate recovered from the condensation of the steam utilised in the Process

Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.

Flow rates of the above interfaces of the Plant are shown in the table attached to para 1.3, Utilities Consumption.

The HP saturated steam from the Syngas Treatment and Conditioning line (Unit 2200) is mixed with the HP steam generated in the coil, superheated and expanded in HP ST down to condenser pressure including one stage of reheating.

The MHP saturated steam at 70 bar from the gasification island, is superheated in a dedicated coil and sent to the MHP ST where it is expanded down to 5.7 barg and then sent to the low pressure section of the other turbine.

Steam imported to the Power Island is only HP; all other streams are exported. As a consequence, the generated steam pressure levels are the same as those of the Process Units.

1.3 Utility Consumption

The utility consumptions of the process / utility and offsite units are shown in the attached Table.

CLIENT:	IEA GHG	REVISION	Rev.0		
PROJECT:	Hydrogen and Electricity co-production	DATE	July 2007		
LOCATION:	the Netherlands	ISSUED BY	LV		
		CHECKED BY	PC		
		APPROVED BY	SA		

UTILITY CONSUMPTION SUMMARY - SHELL - CASE 1 - Shell gasification w/o CO_2 capture, w/o H_2 production

UNIT	DESCRIPTION UNIT	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS										
1000	Gasification Section	-219.4		-57.0		403.7		11.3		37.6	**29.0**
2100	Air Separation Unit			124.4			113.2			124.4	113.2
2200	Syngas Treatment and Conditioning line		15.1	35.6						50.7	
2300	Acid Gas Removal			12.5						12.5	
2400	Sulphur Recovery (SRU) - Tail gas treatment (TGT)		-0.7	-0.9			4.1	0.9		3.4	
3000	**POWER ISLANDS UNITS**	219.4	-14.4	-123.6	0.0	-403.7	-117.3	-12.2	0.0		
4100 to 5300	**UTILITY and OFFSITE UNITS**			9.0						9.0	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	237.6	142.2

Note: Minus prior to figure means figure is generated

1.4 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex.

CASE 1

Shell gasification, w/o CO_2 capture, w/o H_2 production

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	250.6
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWt**	**1800.8**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWt	1504.4
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWt	1496.6
Syngas treatment efficiency	%	99.5
Gas turbines total power output	MWe	553.6
Steam turbine power output	MWe	338.3
GROSS ELECTRIC POWER (C)	**MWe**	**891.9**
ASU power consumption	MWe	94.5
Process Units consumption	MWe	13.0
Utility Units consumption	MWe	1.6
Offsite Units consumption (including sea cooling water system)	MWe	7.2
Power Islands consumption	MWe	13.3
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**129.6**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**762.3**
Equivalent Gross electrical efficiency (C/A *100) (based on coal LHV)	**%**	**49.5**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**42.3**

1.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown at Section B - para 2.0, and produce electric power. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

The gaseous emissions, liquid effluents and solid wastes from the IGCC Complex are summarised in this section.

1.5.1 Gaseous Emissions

Main Emissions

In normal operation at full load, the main continuous emissions are the combustion flue gases of the two trains of the Power Island, proceeding from the combustion of the Syngas in the two gas turbines, and emission from the coal Drying process.

Table 1.1 summarises expected flow rate and concentration of the combustion flue gas from one train of the Power Island. Both the Combined Cycle Units have the same flue gas composition and flow rate. The total gaseous emissions of the Power Island are given in Table 1.2

Table 1.1 – Expected gaseous emissions from two trains of the Power Island.

	Normal Operation	
Wet gas flow rate, kg/s	1,490	
Flow, Nm^3/h(1)	5,670,140	
Temperature, °C	129	
Composition	(%vol)	
Ar	0.82	
N_2	74.23	
O_2	11.48	
CO_2	7.30	
H_2O	6.17	
Emissions	mg/Nm^3 (1)	kg/h
NOx	80	453.6
SOx	5	28.3
CO	31	176.0
Particulate	5	28.0

(1) Dry gas, O_2 content 15%vol.

In normal operation at full load, the following emission to the atmosphere is foreseen from the Coal Drying Process:

Flow rate	:	35	t/h
N2	:	80	% vol.
$H_2O+O_2+CO_2$	:	20	% vol.
Particulate	:	<10	mg/Nm^3, wet basis.

Minor Emissions

The remainder of the gaseous emissions within the IGCC Complex are created by process vents and fugitive emissions.
Some of the vent points emit continuously; others during process upsets or emergency conditions only. All vent streams containing, potentially, undesirable gaseous components are sent to a flare system. Venting via the flare will be minimal during normal operation, but will be significant during emergencies, process upsets, start up and shutdown.

A small continuous emission is generated in the Waste Water Treatment plant; in fact a small burner is installed to destroy the biogas stream coming from the anaerobic section of the plant.

Fugitive emissions are those emissions caused by storage and handling of materials (solids transfer, leakage, etc.). They are prevented by proper design and operation.

1.5.2 Liquid Effluent

The effluent from the Waste Water Treatment (Unit 4600) is recovered and recycled back to the gasification island.

Sea water in open circuit is used for cooling.
The return stream water is treated with meta-bisulphite in the Dechlorination System to reduce the Cl_2 concentration. Main characteristics of the water are listed in the following:

- Maximum flow rate : 87.800 m^3/h
- Temperature : 19 °C
- Cl_2 : <0.05 ppm

1.5.3 Solid Effluent

The process does not produce any solid waste, except for typical industrial plant waste e.g. (sludge from Waste Water Treatment etc.). In any case, the waste water sludge (expected flow rate: 2 m^3/h) can be recovered, recycled back to the Gasification Island and burned into the Gasifier.

In addition, the Gasification Island is expected to produce the following solid by-products:

<u>Slag from Slag Removal Unit</u>

Flow rate	:	37.2	t/h
Water content	:	10	%wt

Slag product can be sold to be commercially used as major components in concrete mixtures to make road, pads, storage bins.

<u>Flyash from Dry Solids Removal Unit</u>

Flow rate	:	1.2	t/h

Flyash can be dispatched to cement industries.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 2 – PLANT W/O H_2 PRODUCTION, WITH CO_2 CAPTURE

ISSUED BY : L.VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007 July 2007	Draft Rev 1	L. Valota L.Valota	P. Cotone P. Cotone	S. Arienti S. Arienti

SECTION G.2 HYDROGEN AND ELECTRICITY COPRODUTION BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION G.2 BASIC INFORMATION FOR EACH ALTERNATIVE

2.0 Case 2

2.1 Introduction

The main features of the Case 2 configuration of the IGCC Complex are:

- Low pressure (39 bar g) Shell Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Heat Recovery Type;
- Double stage dirty shift;
- Separate removal of H_2S and CO_2.

The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The degree of integration between the Air Separation Unit (ASU) and the Gas Turbines is 30%. Gas Turbine power augmentation and syngas dilution, for NO_x control, is achieved with injection of compressed N_2 from ASU to the gas turbines.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

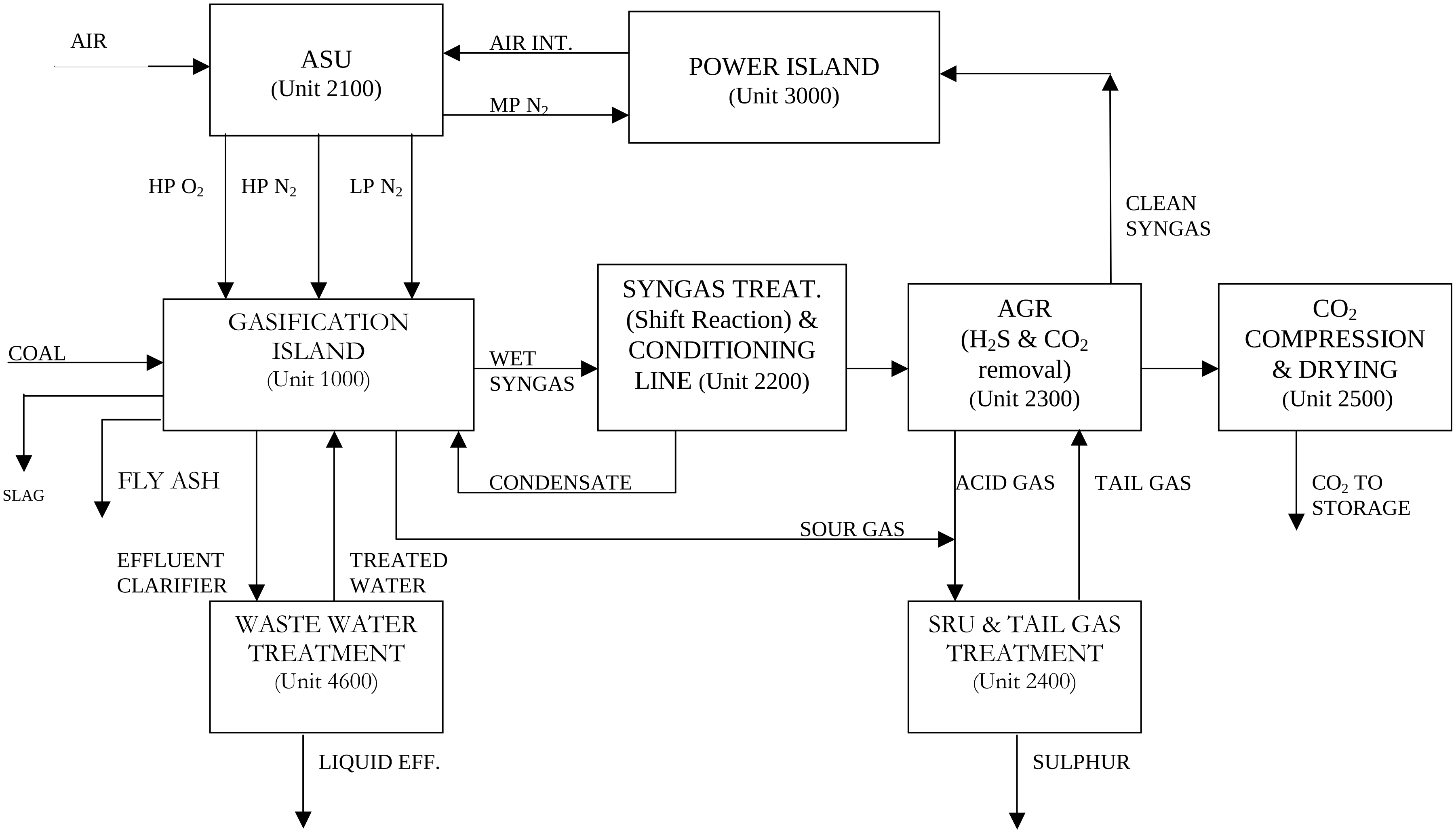
The arrangement of the process units is:

Unit		Trains
900	Coal milling and drying	4 x 33 %
1000	Coal pressurization/feeding	6 x 20 %
	Gasification heat recovery	2 x 50 %
	Slag removal	2 x 50 %
	Dry solids removal	2 x 50 %
	Wet scrubbing	2 x 50 %
	Sour slurry and sour water stripper	1 x 100 %
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%
2300	AGR	2 x 50%
2400	SRU	2 x 100%

	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50%
3000	Gas Turbine (PG 9351 – FA)	2 x 50%
	HRSG	2 x 50%
	Steam Turbine	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.

CASE 2 – IGCC COMPLEX BLOCK FLOW DIAGRAM



2.2 Process Description

Unit 1000: Gasification Island

Information relevant to Shell Gasification Island are collected in para 1.1 of Section C.
The main process data of the Gasification Island relevant to this alternative are summarised in following table:

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	80	80	70	160
Pressure (bar)		40	69	7.5	37
TOTAL FLOW					
Mass flow (kg/h)	273,100	214,550	87,000	33,680	568,200
Molar flow (kmol/h)			3,100	1,200	28,850
Composition (% vol)					
H_2					26.25
CO					49.60
CO_2					1.24
N_2		3.5	99.88	99.88	4.00
Ar		1.5	0.08	0.08	0.62
O_2		95.0	0.04	0.04	0.00
H_2S + COS					0.23
H_2O					18.05
Others					0.01

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package unit supplied by specialised Vendors. Reference is made to Section C, para. 2.0 for a general description of the Air Separation Unit.

The integration value between ASU and Gas Turbine is the percentage of the air extracted from a Gas Turbine sent to ASU over the total air required by ASU. It has been optimized and the optimum arrangement presents an integration of 30%.

The main process data and the main consumption of the ASU are summarised in the following tables.

Process Data	Mass Flow (kg/h)	
Air from ambient	656,570	
Air from GT	281,400	
Oxygen to gasifier (95% vol)	214,550	
LP Nitrogen to Gasification Island (98% vol)	33,700	
HP Nitrogen to Gasification Island (98% vol)	87,000	
Nitrogen to Power Island (for syngas dilution)	608,700	
Consumption		
Main air compressor	56,200	kW
Oxygen compressor	11,000	kW
Nitrogen compressor	43,800	kW
Miscellanea	1,500	kW
Total	112,500	kW

<u>Unit 2200: Syngas Treatment and Conditioning Line</u>

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 36 barg and 160°C, enters Unit 2200. The syngas is first heated by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2 and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 451°C. Due to the low water content of the syngas, the injection of MP steam to the syngas is required before entering the shift reactor. In order to meet the required degree of CO_2 removal, a double stage shift containing sulphur tolerant shift catalyst (dirty shift) is used. The hot shifted syngas outlet from the first stage is cooled in a series of heat exchangers:

Shift feed product exchanger
HP Steam Generator
MP Steam Generator

Inlet temperature to the second stage shift is controlled to 250 °C. Outlet temperature from second shift is 331°C. The hot shifted syngas outlet from the second stage is cooled in a series of heat exchangers:

MP Steam Generator
LP Steam Generator
VLP Steam Generator
Condensate Preheater

A final syngas cooling step with cooling water is present. Process condensate separated in Separator Drums is recycled back to the Sour Water Stripper of the Gasification Island.
The first stage of the shift reactor is split into three parallel trains. Downstream this point, Unit 2200 is split into two parallel lines, each sized for 50% capacity of the total syngas flow because of the size limitation of the exchangers involved.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
Clean syngas is then preheated with VLP steam and then sent to the gas turbines, Unit 3000.

Unit 2300: Acid Gas Removal (AGR)

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the entire complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (26 bar g) and an extremely high CO_2/H_2S ratio (205/1).

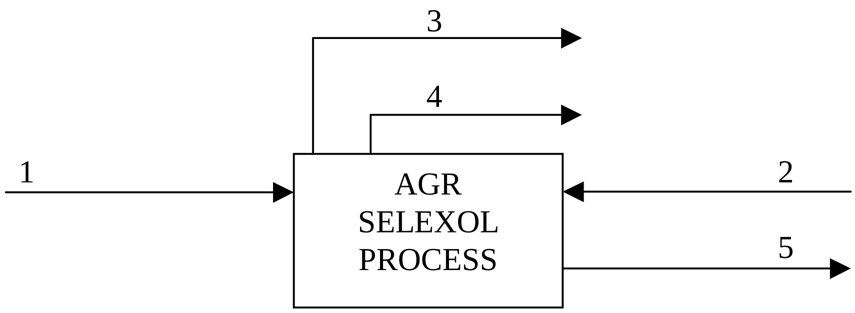
The interfaces of the process are the following, as shown in the scheme:

Entering Streams

1. Untreated Gas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit.

Exit Streams

3. Treated Gas to Gas Turbines
4. CO_2 to Compression
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treament	**Recycle Gas (tail gas) from SRU**	**Treated gas**	**CO_2 to compression**	**Acid gas to SRU**
Temperature (°C)	38	38	34	(1)	49
Pressure (bar)	27.8	27.0	27.0	(1)	1.8
Mass flow (kg/h)	714433	13011	164839	549273	13419
Molar flow (kgmole/h)	37113	332	24480	12728	336
Composition (vol %)					
H_2	56.51	4.10	85.35	1.74	0.28
CO	2.51	0.15	3.74	0.19	0.03
CO_2	36.91	76.63	5.24	97.69	72.41
N_2	3.10	17.78	4.93	0.06	0.01
CH_4	0.00	0.00	0.00	0.00	0.00
H_2S	0.18	0.72	0.00	0.01	20.25
COS	0.00	0.01	0.00	0.00	0.02
Ar	0.48	0.19	0.72	0.03	0.01
H2O	0.31	0.42	0.03	0.28	6.46

Note (1): CO_2 stream is the combination of three different streams at following pressure levels 26.0, 3.5 and 0.5 barg.

The Selexol solvent consumption, to make-up losses, is 120 m^3/year.

The proposed process matches the process specification with reference to H_2S+COS concentration of the treated gas exiting the Unit (H_2S+COS concentration is 3 ppm). This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power Consumption = 41% of the overall AGR Power requirement) before flowing to the CO_2 absorber.

The CO_2 removal rate is 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 22% dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1.7% vol and to an overall thermal power of 14.9 MWth, i.e. almost 5 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 100 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

Unit 2400: SRU and TGT

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 5.0 for the general information about the technology.

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of 56.4 t/day, and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 28 barg.

Unit 2500: CO_2 Compression and Drying

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 26.0 barg
- LP stream : 3.5 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is mainly composed of CO_2 and CO. The main properties of the stream are as follows:

- Product stream : 550 t/h.
- Product stream : 110 bar.
- Composition :

	%wt
CO_2	99.8
CO	0.1
Others	0.1
TOTAL	100.0

Unit 3000: Power Island

For general information about the Power Island technology refer to Section C, para. 9.0

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

Imported

- HP steam (160 barg) : Steam imported from Syngas Treatment and Conditioning Line.
- MHP steam (70 barg) : Steam imported from the Gasification.
- VLP steam (3.2 barg): Steam imported from Syngas Treatment and Conditioning Line.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

Exported

- MP steam (40 barg): Steam exported to Syngas Treatment and Conditioning Line to meet the water requirement of the shift reaction. A small quantity of steam is also generated in the Sulphur Recovery Unit and in the Tail Gas Treatment Unit.
- LP steam (6.5 barg): Steam exported to the following Process Units: AGR, ASU, Utility and Offsite Unit. LP steam is also generated in the Syngas Treatment and Conditioning Line.
- BFW: HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam.
- Process Condensate: All the condensate recovered from the condensation of the steam utilised in the Process Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.

The steam turbine in the Power Island consists of two sections: One High Pressure Steam turbine (HP ST) and one Medium High Pressure Steam turbine (MHP ST).

The HP saturated steam at 160 bar from the Syngas Treatment and Conditioning line (Unit 2200) is mixed with the HP steam generated in the coil, superheated and expanded in HP ST down to condenser pressure.

The MHP saturated steam at 70 bar from the gasification island, is superheated in a dedicated coil and sent to the MHP ST where is expanded down to 5.7 barg and then sent to the low pressure section of the other turbine.

MP steam coming from HP section is reheated in the HRSG and then sent to MP section of the steam turbine.
The total steam coming from MP ST outlet, MHP ST outlet and Superheated LP steam from HRSG is sent to LP ST section where is expanded to condensation.

Steam imported to the Power Island is HP and VLP steam; all other streams are exported. As a consequence, the generated steam pressure levels are the same as those of the Process Units.

2.5 Utility Consumption

The utility consumption of the process / utility and offsite units are shown in the attached Table.

FOSTER WHEELER

CLIENT: IEA GHG
PROJECT: Hydrogen and Electricity co-production
LOCATION:

REVISION	Rev.0			
DATE	July 2007			
ISSUED BY	LV			
CHECKED BY	PC			
APPROVED BY	SA			

UTILITY CONSUMPTION SUMMARY - CASE 2 - Shell gassification, with CO_2 capture, w/o H_2 production

UNIT	DESCRIPTION UNIT	HP Steam 160 barg [t/h]	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS											
1000	**Gasification Section**		-317.4				390.9				41.3	32.2
2100	**Air Separation Unit**				16.8						16.8	
2200	**Syngas Treatment and Conditioning line**	-40.6		267.3	-75.5	-85.9	40.6	150.5	75.5	120.3	34.4	417.8
2300	**Acid Gas Removal**				82.4						82.4	
2400	**Sulphur Recovery (SRU) - Tail gas treatment (TGT)**			-0.7	-1.0			4.3	1.0		3.6	
3000	**POWER ISLANDS UNITS**	40.6	317.4	-266.6	-32.1	85.9	-431.5	-154.8	-76.5	-120.3		
4100 to 5300	**UTILITY and OFFSITE UNITS**				9.4						9.4	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	187.9	450.0

Note: Minus prior to figure means figure is generated

2.6 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex.

CASE 2

Shell gasification, with CO2 capture, w/o H2 production

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	273.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWth**	**1962.5**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWth	1638.2
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWth	1467.2
Syngas treatment efficiency	%	89.6
Gas turbines total power output	MWe	572.0
Steam turbine power output	MWe	303.0
GROSS ELECTRIC POWER (C)	**MWe**	**875.0**
ASU power consumption	MWe	112.5
Process Units consumption	MWe	48.0
Utility Units consumption	MWe	2.6
Offsite Units consumption (including sea cooling water system)	MWe	8.9
Power Islands consumption	MWe	14.6
CO_2 compression and Drying	MWe	32.6
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**219.2**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**655.8**
Equivalent Gross electrical efficiency (C/A *100) (based on coal LHV)	**%**	**44.6**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**33.4**

The following Table shows the overall CO_2 removal efficiency of the IGCC Complex.

	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	14701
Slag (Carbon =~0.4% wt) *	61
Net Carbon flowing to Process Units (A)	14640
Liquid Storage	
CO	24,0
CO_2	12434.0
Total to storage (B)	12458.0
Emission	
CO_2	2177.4
CO	5.6
Total Emission	2183.0
Overall CO_2 removal efficiency, % (B/A)	85.1

* The percentage of unreacted C stated by Shell is 0.2%. However, the carbon mass balance of the whole IGCC results in a 0.4% carbon less. This value is conservatively assumed.

2.7 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristic is shown at Section B - para 2.0, and produce electric power. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

The gaseous emissions, liquid effluents and solid wastes from the IGCC Complex are summarised in this section.

2.7.1 Gaseous Emissions

<u>Main Emissions</u>
In normal operation at full load, the main continuous emissions are the combustion flue gases of the two trains of the Power Island, proceeding from the combustion of the Syngas in the two gas turbines, and emission from the coal Drying process.

Table 2.1 summarises expected flow rate and concentration of the combustion flue gas from one train of the Power Island.

Table 2.1 – Expected gaseous emissions from one train of the Power Island.

	Normal Operation
Wet gas flow rate, kg/s	697.6
Flow, $Nm^3/h^{(1)}$	2,507,890
Temperature, °C	129
Composition	(%vol)
Ar	0.91
N_2	74.95
O_2	11.17
CO_2	1.20
H_2O	11.77
Emissions	$mg/Nm^{3\,(1)}$
NOx	74
SOx	1
CO	31
Particulate	5

(1) Dry gas, O_2 content 15%vol

Both the Combined Cycle Units have the same flue gas composition and flow rate. The total gaseous emissions of the Power Island are given in Table 2.2.

Table 2.2 – Expected total gaseous emissions of the Power Island.

	Normal Operation
Wet gas flow rate, kg/s	1395.2
Flow, $Nm^3/h^{(1)}$	5,015,780
Temperature, °C	129
Emissions	kg/h
NOx	371.2
SOx	5.0
CO	155.5
Particulate	25.1

(1) Dry gas, O_2 content 15%vol

In normal operation at full load, the following emission to the atmosphere is foreseen from the Coal Drying Process:

Flow rate	39	t/h
N_2	80	% vol.
$H_2O+O_2+CO_2$	20	% vol.
Particulate	<10	mg/Nm^3, wet basis.

Minor Emissions

The remainder of the gaseous emissions within the IGCC Complex are created by process vents and fugitive emissions.
Some of the vent points emit continuously; others during process upsets or emergency conditions only. All vent streams containing, potentially, undesirable gaseous components are sent to a flare system. Venting via the flare will be minimal during normal operation, but will be significant during emergencies, process upsets, start up and shutdown.

A small continuous emission is generated in the Waste Water Treatment plant; in fact a small burner is installed to destroy the biogas stream coming from the anaerobic section of the plant.

Fugitive emissions are those emissions caused by storage and handling of materials (solids transfer, leakage, etc.). Proper design and operation prevent them.

2.7.2 Liquid Effluent

Waste Water Treatment (Unit 4600)
Part of the effluent from the Waste Water Treatment (Unit 4600) is recovered and recycled back to the gasification island as process water, closing the Gasification water balance. The other part is sent to a dedicated treatment where the reverse osmosis process allows recovering almost 60% of the treated water. This recovered water is recycled back to the Demi Water System, Unit 4200, and used as raw water for the Demineralized water plant. The remaining 40% of water is discharged together with the sea cooling water return stream. The expected flow rate of this stream is as follows:

- Flow rate : 46 m^3/h

Sea Water System (Unit 4100)
Sea water in open circuit is used for cooling.
The return stream water is treated with meta-bisulphite in the Dechlorination System to reduce the Cl_2 concentration. Main characteristics of the water are listed in the following:

- Maximum flow rate : 93,160 m^3/h
- Temperature : 19 °C
- Cl_2 : <0.05 ppm

2.7.3 Solid Effluent

The process does not produce any solid waste, except for typical industrial plant waste e.g. (sludge from Waste Water Treatment etc.). In any case, the waste water sludge (expected flow rate: 2 m^3/h) can be recovered, recycled back to the Gasification Island and burned into the Gasifier.

In addition, the Gasification Island is expected to produce the following solid by-products:

Slag from Slag Removal Unit

Flow rate	:	40.5	t/h
Water content	:	10	%wt

Slag product can be sold to be commercially used as major components in concrete mixtures to make road, pads, storage bins.

Flyash from Dry Solids Removal Unit

Flow rate : 1.3 t/h

Fly ash can be dispatched to cement industries.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 3 – H_2 PRODUCTION PLANT

ISSUED BY : L.VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007 July 2007	Draft Rev 1	L. Valota L.Valota	P. Cotone P. Cotone	S. Arienti S. Arienti

SECTION G.3 HYDROGEN AND ELECTRICITY COPRODUTION BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION G.3 BASIC INFORMATION FOR EACH ALTERNATIVE

3.0 Case G.3

3.1 Introduction

The main features of the Case 3 configuration of the IGCC Complex are:

- Low pressure (39 bar g) Shell Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Heat Recovery Type;
- Double stage dirty shift;
- Separate removal of H_2S and CO_2;
- PSA unit for Hydrogen production with Off-Gas Compression
- Gas Turbine (General Electric 6FA)

The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The Air Separation Unit (ASU) and the Gas Turbine are not integrated. Gas Turbine NO_x emission reduction is achieved diluting the syngas with compressed N_2 from ASU.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

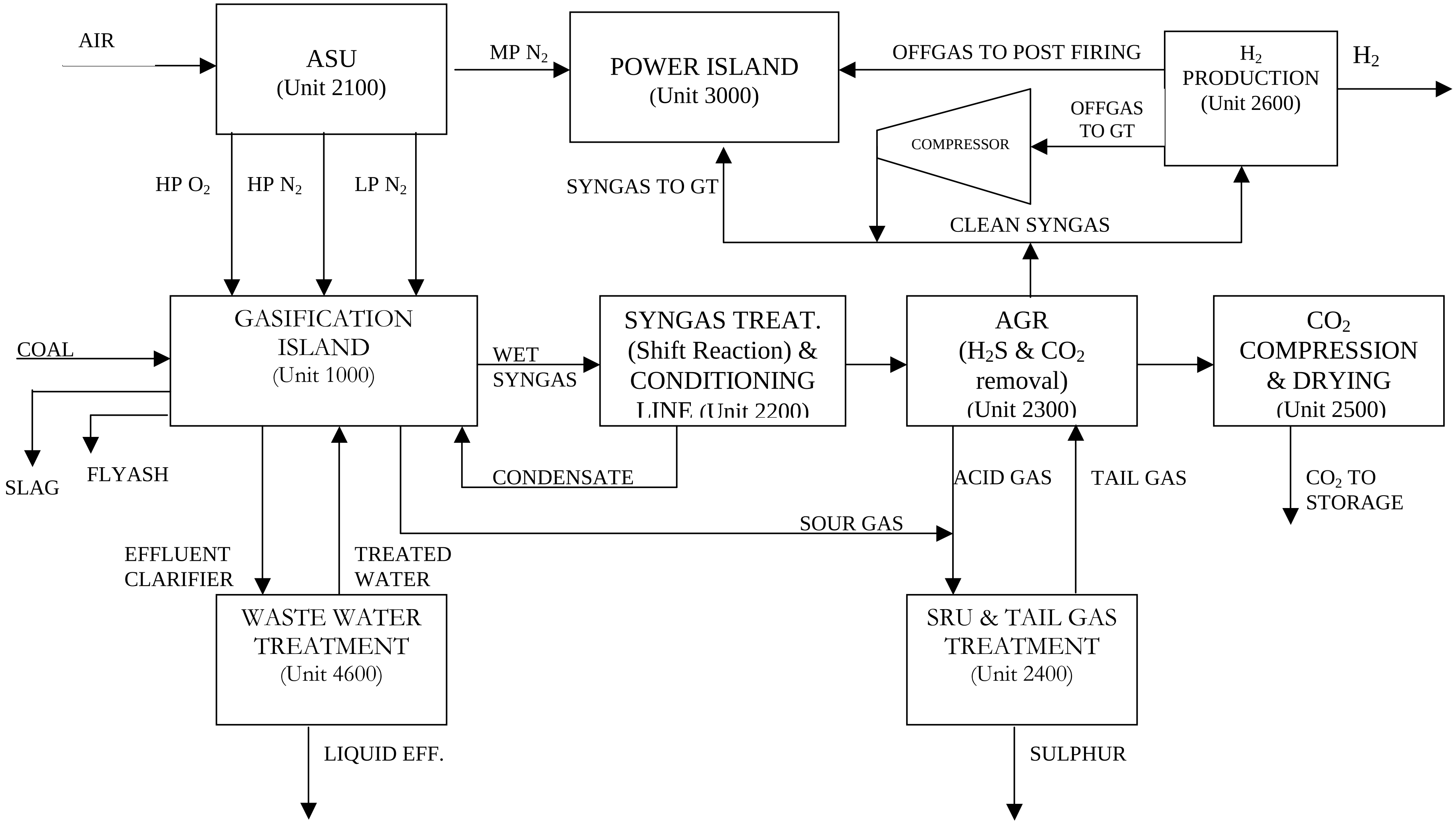
The arrangement of the process units is:

Unit		**Trains**
900	Coal milling and drying	4 x 33 %
1000	Coal pressurization/feeding	6 x 20 %
	Gasification heat recovery	2 x 50 %
	Slag removal	2 x 50 %
	Dry solids removal	2 x 50 %
	Wet scrubbing	2 x 50 %
	Sour slurry and sour water stripper	1 x 100 %
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%
2300	AGR	2 x 50%

2400	SRU	2 x 100%
	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50%
2600	H_2 production	1 x 100%
3000	Gas Turbine (PG6111-6FA)	1 x 100%
	HRSG	1 x 100%
	Steam Turbine	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.

CASE 3 – IGCC COMPLEX BLOCK FLOW DIAGRAM



3.2 Process Description

Unit 1000: Gasification Island

Shell Gasification Island relevant in information are collected in para. 1.1 of Section C.
The following table summarised the main process data of the Gasification Island for this alternative.

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	80	80	70	160
Pressure (bar)		40	69	7.5	37
TOTAL FLOW					
Mass flow (kg/h)	273,100	214,550	87,000	33,680	568,200
Molar flow (kmol/h)			3,100	1,200	28,850
Composition (% vol)					
H_2					26.25
CO					49.60
CO_2					1.24
N_2		3.5	99.88	99.88	4.00
Ar		1.5	0.08	0.08	0.62
O_2		95.0	0.04	0.04	0.00
H_2S + COS					0.23
H_2O					18.05
Others					0.01

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package supplied by specialised Vendors. For a general description of the Air Separation Unit refer to Section C, para. 3.0

The integration between ASU and Gas Turbine has been optimized considering a plant with production of hydrogen and co-production of the minimum amount of electricity to compensate the complex internal electrical consumption.
In the optimum arrangement there is no integration between ASU and Gas Turbine. In fact the maximum flowrate that can be extracted from one gas turbine is a small fraction in comparison to the total ASU air intake; therefore the integration between ASU and Power Island would not lead to an optimized configuration. Thus, when the gasification operates, the air required by the

ASU to obtain the design oxygen production is entirely derived from self-standing compressor units.

The main process data and the main consumption of the ASU are summarised in following tables.

Process Data	Mass Flow (kg/h)	
Air from ambient	951,900	
Oxygen to gasifier (95% vol)	214,550	
LP Nitrogen to Gasification Island (98% vol)	33,700	
HP Nitrogen to Gasification Island (98% vol)	87,000	
Nitrogen to Power Island (for syngas dilution)	58,000	
Consumption		
Main air compressor	76,300	kW
Oxygen compressor	11,000	kW
Nitrogen compressor	13,500	kW
Miscellanea	1,300	kW
Total	102,100	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 36 barg and 160°C, enters the Sour Shift section of Unit 2200. The syngas is first heated in a gas/gas exchanger by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2 and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 451°C. Due to the low water content of the syngas, the injection of MP steam to the syngas is required before entering the shift reactor. In order to meet the required degree of CO_2 removal, a double stage shift containing sulphur tolerant shift catalyst (dirty shift) is used. The hot shifted syngas outlet from the first stage is cooled in a series of heat exchangers:

Shift feed product exchanger
HP Steam Generator
MP Steam Generator

Inlet temperature to the second stage shift is controlled to 250°C. Outlet temperature from second shift is 331°C. The hot shifted syngas outlet from the second stage is cooled in a series of heat exchangers:

MP Steam Generator
LP Steam Generator
VLP Steam Generator
Condensate from CCU Preheater

The final cooling step of the syngas takes place in a cooling water cooler, where syngas is cooled with cooling water. Process condensate separated in syngas cooling is recycled back to the Sour Water Stripper of the Gasification Island.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
The syngas is then spit in two streams. The first consists of around 90% of the total syngas and is fed to the hydrogen production unit. The remaining clean syngas is preheated with VLP steam, diluted with nitrogen and sent to the Gas Turbine (Unit 3000).

<u>Unit 2300: Acid Gas Removal (AGR)</u>

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the entire complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (27 bar g) and an extremely high CO_2/H_2S ratio (205/1).

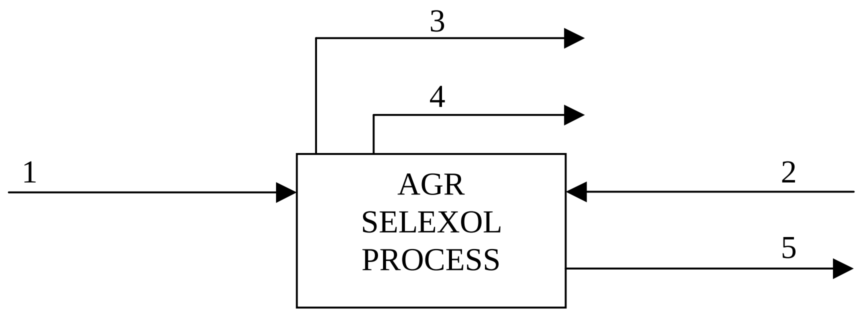
The interfaces of the process are the following, as shown in the scheme:

<u>Entering Streams</u>

1. Raw syngas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit.

Exit Streams

3. Treated Gas
4. CO_2 to compression
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treament	**Recycle Gas (tail gas) from SRU**	**Treated gas**	**CO_2 to compression**	**Acid gas to SRU**
Temperature (°C)	38	38	34	(1)	49
Pressure (bar)	27.8	27.0	27.0	(1)	1.8
Mass flow (kg/h)	714433	13011	164839	549273	13419
Molar flow (kgmole/h)	37113	332	24480	12728	336
Composition (vol %)					
H_2	56.51	4.10	85.35	1.74	0.28
CO	2.51	0.15	3.74	0.19	0.03
CO_2	36.91	76.63	5.24	97.69	72.41
N_2	3.10	17.78	4.93	0.06	0.01
CH_4	0.00	0.00	0.00	0.00	0.00
H_2S	0.18	0.72	0.00	0.01	20.25
COS	0.00	0.01	0.00	0.00	0.02
Ar	0.48	0.19	0.72	0.03	0.01
H2O	0.31	0.42	0.03	0.28	6.46

Note (1): CO_2 stream is the combination of three different streams at following pressure levels 26.0, 3.5 and 0.5 barg.

The Selexol solvent consumption, to make-up losses, is 120 m^3/year.

The proposed process matches the process specification with reference to H_2S+COS concentration of the treated gas exiting the Unit (H_2S+COS concentration is 3 ppm). This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power Consumption = 41% of the overall AGR Power requirement) before flowing to the CO_2 absorber.

The CO_2 removal rate is 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 22% dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1.7% vol and to an overall thermal power of 14.9 MWt, i.e. almost 5 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 100 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

Unit 2400: SRU and TGT

This Unit is treated as a package supplied by specialised Vendors. For general information about the technology refer to Section C, para. 6.0

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of 56.4 t/day, and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 27 barg.

Unit 2500: CO_2 Compression and Drying

This Unit is treated as a package supplied by specialised Vendors. For general information about the technology refer to Section C, para. 7.0

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 26.0 barg
- LP stream : 3.5 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is mainly composed of CO_2 and CO. The main properties of the stream are as follows:

- Product stream : 550 t/h.
- Product stream : 110 bar.

- Composition :

	%wt
CO_2	99.8
CO	0.1
Others	0.1
TOTAL	100.0

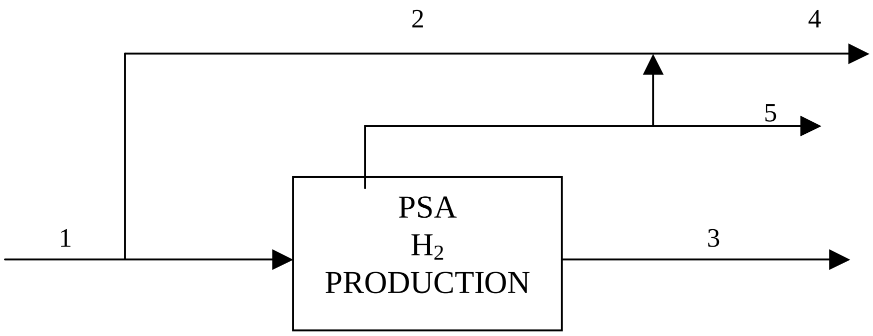
Unit 2600: H_2 Production

This Unit is treated as a package supplied by specialised Vendors. For general information about the technology refer to Section C, para. 8.0

A small portion of the syngas entering the unit bypasses the PSA and is sent to the post firing system of the HRSG together with the PSA off gas to make the burners flame stable.

The interfaces of the process are the following, as shown in the scheme:

1. Total clean syngas from AGR
2. Bypass to post firing
3. Hydrogen
4. Offgas to post firing
5. Offgas to GT



The main process data of the hydrogen production unit are summarised in following table:

	1 Syngas	2 By pass	3 H_2	4 Offgas to PF	5 Offgas to GT
Hydrogen	85.35	85.35	99.50	46.66	43.73
Nitrogen	4.93	4.93	0.40	17.32	18.25
Argon	0.72	0.72	0.10	2.42	2.54
Carbon Monoxide	3.74	3.74		13.97	14.74
Carbon Dioxide	5.24	5.24		19.57	20.65
Methane	0.00	0.00		0.00	0.00
Water	0.02	0.02		0.07	0.08
Hydrogen Sulfide	0.00	0.00		0.00	0.00
Flow (Nm^3/h)	503,867	4,817	372,429	131,438	62,934
Flow (kmol/h)	22,480	215	16,616	3,056	2,808
Flow (kg/h)	151,678	1,450	35,792	59,009	56,877
p (barg)	26.0	26.0	25.2	0.7	0.7
T (°C)	34	34	39	26	26

Off-gas is equally split in two streams: the first is mixed with the bypass and sent to the Post Firing (Unit 3000) while the second is compressed in an external compressor, mixed with the clean syngas from AGR and sent to the Gas Turbine (Unit 3000).

Unit 3000: Power Island

For general information about the Power Island technology refer to Section C, para. 9.0

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

Imported

- HP steam (160 barg) : steam imported from Syngas Treatment and Conditioning Line.
- MHP steam (70 barg) : steam imported from Gasification section.
- VLP steam (3.2 barg): steam imported from Syngas Treatment and Conditioning Line.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

Exported

- MP steam (40 barg): steam exported to Syngas Treatment and Conditioning Line to meet the water requirement of the shift reaction. A small quantity of steam is also generated in the Sulphur Recovery Unit and in the Tail Gas Treatment Unit.
- LP steam(6.5 barg): steam exported to the following Process Units: AGR, ASU, Utility and Offsite Unit. LP steam is also generated in the Syngas Treatment and Conditioning Line.
- BFW: HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate: All the condensate recovered from the condensation of the steam utilised in the Process Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.

Two Post Firing sections are present in the configuration with a total thermal power delivered of 130 MWth.

The steam turbine in the Power Island consists of two sections: One High Pressure Steam turbine (HP ST) and One Medium High Pressure Steam turbine (MHP ST).

The HP saturated steam at 160 bar from the Syngas Treatment and Conditioning line (Unit 2200) is mixed with the HP steam generated in the coil, superheated and expanded in HP ST down to 40 barg. This stream is then mixed with the steam generated at 40 barg and with an extraction from MHP ST, and finally sent to the Syngas Treatment and Conditioning Line (Unit 2200).

The MHP saturated steam at 70 bar from the gasification island, is superheated in a dedicated coil and sent to the MHP ST where is expanded down to condenser pressure.

Flow rate of the above interfaces of the Plant are shown in table attached to para 1.3, Utility Consumption.

3.3 Utility Consumption

The utility consumption of the process / utility and offsite units are shown in the attached Tables.

FOSTER WHEELER

CLIENT: IEA GHG
PROJECT: Hydrogen and Electricity co-production
LOCATION:

REVISION	Rev.0				
DATE	July 2007				
ISSUED BY	LV				
CHECKED BY	PC				
APPROVED BY	SA				

UTILITY CONSUMPTION SUMMARY - CASE 3 - H_2 Production Plant

UNIT	DESCRIPTION UNIT	HP Steam 160 barg [t/h]	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS											
1000	Gasification Section		-317.4				390.9				41.3	32.2
2100	Air Separation Unit				16.8						16.8	
2200	Syngas Treatment and Conditioning line	-40.8		270.7	-75.0	-58.5	53.5	147.1	75.0	65.7	7.2	430.5
2300	Acid Gas Removal				82.4						82.4	
2400	Sulphur Recovery (SRU) - Tail gas treatment (TGT)			-0.7	-1.0			4.3	1.0		3.6	
3000	POWER ISLANDS UNITS	40.8	317.4	-269.9	-32.6	58.5	-444.4	-151.4	-76.0	-65.7		
4100 to 5300	**UTILITY and OFFSITE UNITS**				9.4						9.4	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	160.7	462.7

Note: Minus prior to figure means figure is generated

3.4 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex.

CASE 3		
Shell gasification, with CO_2 capture, with maximum H_2 production		
OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	273.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWt**	**1962.5**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWt	1638.2
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWt	1467.2
Syngas treatment efficiency	%	89.6
Hydrogen production (99.5% purity)	Nm^3/h	372,400
Hydrogen Thermal Power (E)	MWt	1110.7
Equivalent H_2 based combined cycle net efficiency	%	51.4
Gas turbines total power output	MWe	87.6
Steam turbine power output	MWe	121.0
Equivalent Electric Power from H_2	MWe	570.9
ACTUAL GROSS ELECTRIC POWER OUTPUT	**MWe**	**208.6**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**779.5**
ASU power consumption	MWe	102.1
Process Units consumption	MWe	58.6
Utility Units consumption	MWe	2.4
Offsite Units consumption (including sea cooling water system)	MWe	6.2
Power Islands consumption	MWe	6.6
CO_2 compression and Drying	MWe	32.6
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**208.5**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**0.1**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**571.0**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**39.7**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**29.1**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**0.0**
Net thermal H_2 output efficiency (E/A*100) (based on coal LHV)	**%**	**56.6**

The following Table shows the overall CO_2 removal efficiency of the IGCC Complex.

	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	14701
Slag (Carbon =~0.4% wt) *	61
Net Carbon flowing to Process Units (A)	14640
Liquid Storage	
CO	24
CO_2	12434
Total to storage (B)	12458
Emission	
CO_2	2181
CO	2
Total Emission	2183
Overall CO_2 removal efficiency, % (B/A)	85.1

* The percentage of unreacted C stated by Shell is 0.2%. However, the carbon mass balance of the whole IGCC results in a 0.4% carbon less. This value is conservatively assumed.

3.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown in Section B - para 2.0, and co-produce electric power and hydrogen. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

3.5.1 Gaseous Emissions

Main Emissions

In normal operation at full load, the main continuous emissions are the combustion flue gas of the Power Island, proceeding from the combustion of the Syngas in one gas turbine, and the emission from the coal Drying process.

The next table summarises expected flow rate and concentration of the combustion flue gas from the Power Island.

	Normal Operation	
Wet gas flow rate, kg/s	244.9	
Flow, Nm^3/h(1)	1,270,000	
Temperature, °C	130	
Composition	(%vol)	
Ar	1.13	
N_2	73.8	
O_2	7.86	
CO_2	7.07	
H_2O	10.14	
Emissions	mg/Nm^3 (1)	kg/h
NOx	66	83.6
SOx	4	5
CO	28	36
Particulate	5	6.3

(1) Dry gas, O_2 content 15%vol

In normal operation at full load, the following emission to the atmosphere is foreseen from the Coal Drying Process:

Flow rate	39	t/h
N_2	80	% vol.
$H_2O+O_2+CO_2$	20	% vol.
Particulate	<10	mg/Nm^3, wet basis.

Minor Emissions

The remainder gaseous emissions within the IGCC Complex are created by process vents and fugitive emissions.
Some of the vent points emit continuously; others during process upsets or emergency conditions only. All vent streams containing, potentially, undesirable gaseous components are sent to a flare system. Venting via the flare will be minimal during normal operation, but will be significant during emergencies, process upsets, start up and shutdown.

A small continuous emission is generated in the Waste Water Treatment plant; in fact a small burner is installed to destroy the biogas stream coming from the anaerobic section of the plant.

Fugitive emissions are those emissions caused by storage and handling of materials (solids transfer, leakage, etc.). Proper design and operation prevent them.

3.5.2 Liquid Effluent

Waste Water Treatment (Unit 4600)
Part of the effluent from the Waste Water Treatment (Unit 4600) is recovered and recycled back to the gasification island as process water, closing the Gasification water balance. The other part is sent to a dedicated treatment where the reverse osmosis process allows recovering almost 60% of the treated water. This recovered water is recycled back to the Demi Water System, Unit 4200, and used as raw water for the Demineralised water plant. The remaining 40% of water is discharged together with the sea cooling water return stream. The expected flow rate of this stream is as follows:

- Flow rate : 46 m^3/h

Sea Water System (Unit 4100)
Sea water in open circuit is used for cooling.
The return stream Water is treated with meta-bisulphite in the Dechlorination System to reduce the Cl_2 concentration. Main characteristics of the water are listed in the following:

- Maximum flow rate : 75,230 m^3/h

- Temperature : 19 °C
- Cl_2 : <0.05 ppm

3.5.3 Solid Effluent

The process does not produce any solid waste, except for typical industrial plant waste e.g. (sludge from Waste Water Treatment etc.). In any case, the waste water sludge (expected flow rate: 2 m^3/h) can be recovered, recycled back to the Gasification Island and burned into the Gasifier.

In addition, the Gasification Island is expected to produce the following solid by-products:

<u>Slag from Slag Removal Unit</u>

Flow rate	:	40.5	t/h
Water content	:	10	%wt

Slag product can be sold to be commercially used as major components in concrete mixtures to make road, pads, storage bins.

<u>Flyash from Dry Solids Removal Unit</u>

Flow rate	:	1.3	t/h

Flyash can be dispatched to cement industries.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 4 – OPTIMUM H_2/ELECTRIC ENERGY PRODUCTION PLANT

ISSUED BY : L.VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L.Valota	P. Cotone	S. Arienti

SECTION G.4 HYDROGEN AND ELECTRICITY COPRODUTION BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION G.4 BASIC INFORMATION FOR EACH ALTERNATIVE

4.0 Case 4

4.1 Introduction

The main features of the Case 4 configuration of the IGCC Complex are:

- Low pressure (39 bar g) Shell Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Heat Recovery Type;
- Double stage dirty shift;
- Separate removal of H_2S and CO_2;
- PSA unit for Hydrogen production.

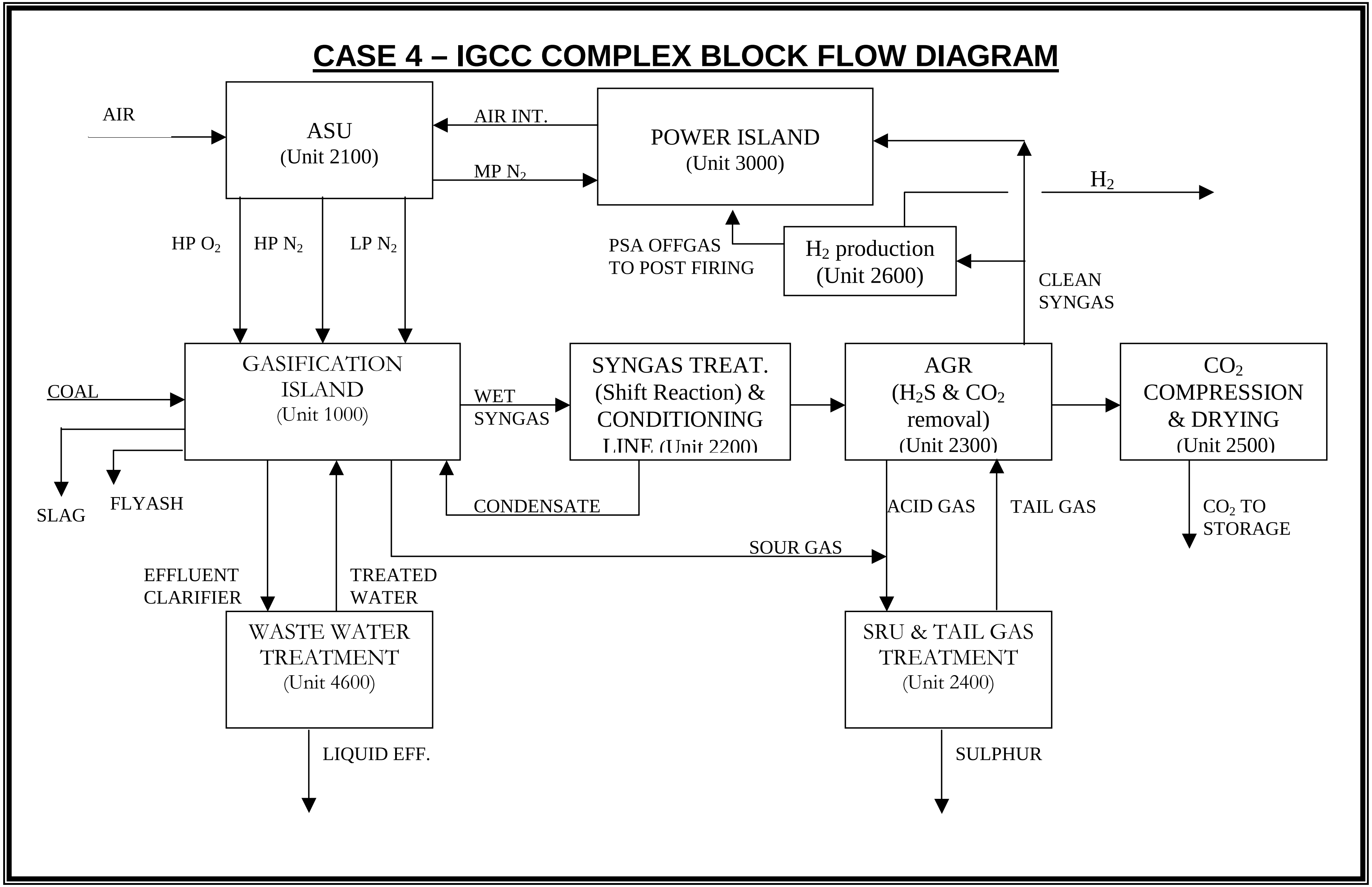
The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The degree of integration between the Air Separation Unit (ASU) and the Gas Turbines is 15%. Gas Turbine power augmentation and syngas dilution, for NO_x control, is achieved with injection of compressed N_2 from ASU to the gas turbines.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.

The arrangement of the process units is:

Unit		Trains
900	Coal milling and drying	4 x 33 %
1000	Coal pressurization/feeding	6 x 20 %
	Gasification heat recovery	2 x 50 %
	Slag removal	2 x 50 %
	Dry solids removal	2 x 50 %
	Wet scrubbing	2 x 50 %
	Sour slurry and sour water stripper	1 x 100 %
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%
2300	AGR	2 x 50%

2400	SRU	2 x 100%
	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50%
2600	H_2 production	1 x 100%
3000	Gas Turbine (PG 9351 – FA)	1 x 100%
	HRSG	1 x 100%
	Steam Turbine	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.


CASE 4 – IGCC COMPLEX BLOCK FLOW DIAGRAM
AIR
ASU
(Unit 2100)
AIR INT.
MP N_2
POWER ISLAND
(Unit 3000)
H_2
HP O_2
HP N_2
LP N_2
PSA OFFGAS
TO POST FIRING
H_2 production
(Unit 2600)
CLEAN
SYNGAS
GASIFICATION
ISLAND
(Unit 1000)
COAL
WET
SYNGAS
SYNGAS TREAT.
(Shift Reaction) &
CONDITIONING
LINE (Unit 2200)
AGR
(H_2S & CO_2
removal)
(Unit 2300)
CO_2
COMPRESSION
& DRYING
(Unit 2500)
SLAG
FLYASH
CONDENSATE
ACID GAS
TAIL GAS
CO_2 TO
STORAGE
SOUR GAS
EFFLUENT
CLARIFIER
TREATED
WATER
WASTE WATER
TREATMENT
(Unit 4600)
SRU & TAIL GAS
TREATMENT
(Unit 2400)
LIQUID EFF.
SULPHUR

4.2 Process Description

Unit 1000: Gasification Island

Information relevant to Shell Gasification Island are collected in para 1.1 of Section C.
The main process data of the Gasification Island relevant to this alternative are summarised in following table:

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	80	80	70	160
Pressure (bar)		40	69	7.5	37
TOTAL FLOW					
Mass flow (kg/h)	273,100	214,550	87,000	33,680	568,200
Molar flow (kmol/h)			3,100	1,200	28,850
Composition (% vol)					
H_2					26.25
CO					49.60
CO_2					1.24
N_2		3.5	99.88	99.88	4.00
Ar		1.5	0.08	0.08	0.62
O_2		95.0	0.04	0.04	0.00
H_2S + COS					0.23
H_2O					18.05
Others					0.01

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package unit supplied by specialised Vendors. Reference is made to Section C, para. 2.0 for a general description of the Air Separation Unit.

The integration between ASU and Gas Turbine has been optimized considering a reference plant with two gas turbines in operation without hydrogen production as the gasification island is sized in order to satisfy the appetite of two gas turbines 9FA in combined cycle. In this configuration, the optimum integration between ASU and Gas Turbine is 30%. When the gasification operates at full load, 30% of the air required by the ASU to obtain the design oxygen production is derived from both gas turbine compressors; the

integration between the gas turbines operation and the ASU is achieved at a level where 70% of the atmospheric air is compressed with selfstanding units and the difference comes already pressurized from the compressors of the gas turbines in the combined cycle.
For the gasification technology selection, only one gas turbine has been considered, as half of clean syngas flowrate, coming from Unit 2200 is sent to Hydrogen production. In this configuration, it is considered the same air extraction from gas turbine and as a consequence the integration between ASU and Gas Turbine is half of the optimized figure (15%).
The main process data and the main consumption of the ASU are summarised in following tables.

	Mass Flow (kg/h)
Air from ambient	804,300
Air from GT	141,900
Oxygen to gasifier (95% vol)	214,550
LP Nitrogen to Gasification Island (98% vol)	33,700
HP Nitrogen to Gasification Island (98% vol)	87,000
Nitrogen to Power Island (for syngas dilution)	304,350

Main air compressor	64,500	kW
Oxygen compressor	11,000	kW
Nitrogen compressor	22,200	kW
Miscellanea	1,400	kW
Total	99,100	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 36 barg and 160°C, enters the Sour Shift section of Unit 2200. The syngas is first heated in a gas/gas exchanger by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2 and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 451°C. Due to the low water content of the syngas, the injection of MP steam to the syngas is required before entering the shift reactor. In order to meet the required degree of CO_2 removal, a double stage shift containing sulphur tolerant shift catalyst (dirty shift) is used. The hot shifted syngas outlet from the first stage is cooled in a series of heat exchangers:

Shift feed product exchanger

HP Steam Generator
MP Steam Generator

Inlet temperature to the second stage shift is controlled to 250°C. Outlet temperature from second shift is 331°C. The hot shifted syngas outlet from the second stage is cooled in a series of heat exchangers:

MP Steam Generator
LP Steam Generator
VLP Steam Generator
Condensate from CCU Preheater

The final cooling step of the syngas takes place in a cooling water cooler, where syngas is cooled with cooling water. Process condensate separated in syngas cooling is recycled back to the Sour Water Stripper of the Gasification Island.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
The syngas is then split in two equal streams: one is fed to the hydrogen production unit; the remaining clean syngas is preheated with VLP steam and sent to the gas turbine (Unit 3000).

Unit 2300: Acid Gas Removal (AGR)

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (27 bar g) and an extremely high CO_2/H_2S ratio (205/1).

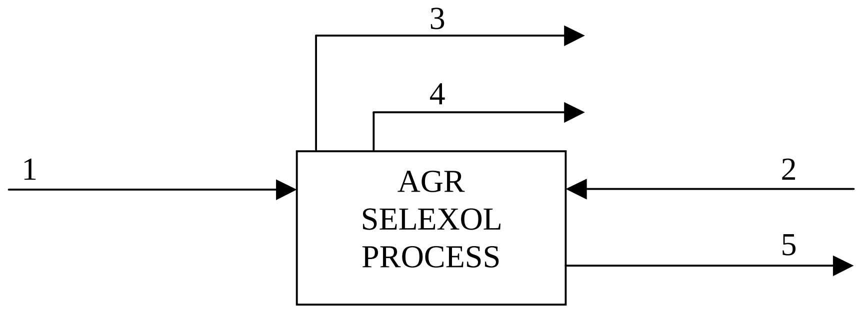
The interfaces of the process are the following, as shown in the scheme:

Entering Streams

1. Raw syngas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit.

<u>Exit Streams</u>

3. Treated Gas
4. CO_2 to compression
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treament	**Recycle Gas (tail gas) from SRU**	**Treated gas**	**CO_2 to compression**	**Acid gas to SRU**
Temperature (°C)	38	38	34	(1)	49
Pressure (bar)	27.8	27.0	27.0	(1)	1.8
Mass flow (kg/h)	714433	13011	164839	549273	13419
Molar flow (kgmole/h)	37113	332	24480	12728	336
Composition (vol %)					
H_2	56.51	4.10	85.35	1.74	0.28
CO	2.51	0.15	3.74	0.19	0.03
CO_2	36.91	76.63	5.24	97.69	72.41
N_2	3.10	17.78	4.93	0.06	0.01
CH_4	0.00	0.00	0.00	0.00	0.00
H_2S	0.18	0.72	0.00	0.01	20.25
COS	0.00	0.01	0.00	0.00	0.02
Ar	0.48	0.19	0.72	0.03	0.01
H2O	0.31	0.42	0.03	0.28	6.46

Note (1): CO_2 stream is the combination of three different streams at following pressure levels 26.0, 3.5 and 0.5 barg;

The Selexol solvent consumption, to make-up losses, is 120 m^3/year.

The proposed process matches the process specification with reference to H_2S+COS concentration of the treated gas exiting the Unit (H_2S+COS concentration is 3 ppm). This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power Consumption = 41% of the overall AGR Power requirement) before flowing to the CO_2 absorber.

The CO_2 removal rate is 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 22% dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1,7% vol and to an overall thermal power of 14,9 MWth, i.e. almost 5 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 100 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

<u>Unit 2400: SRU and TGT</u>

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 5.0 for the general information about the technology.

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of 56.4 t/day, and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 27 barg.

Unit 2500: CO_2 Compression and Drying

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 6.0 for the general information about the technology.

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 26.0 barg
- LP stream : 3.5 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is mainly composed of CO_2 and CO. The main properties of the stream are as follows:

- Product stream : 550 t/h.
- Product stream : 110 bar.
- Composition :

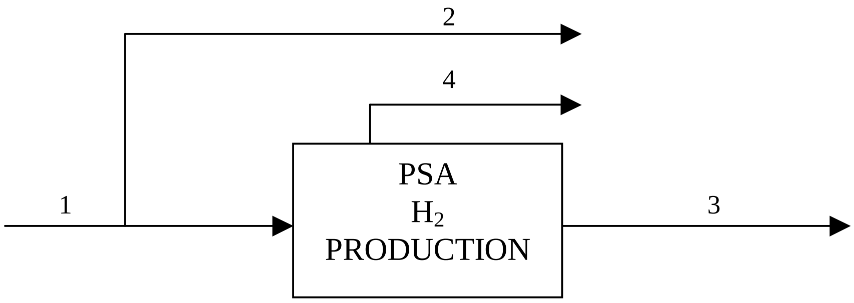
	%wt
CO_2	99.8
CO	0.1
Others	0.1
TOTAL	100.0

Unit 2600: H_2 Production

This Unit is a Package Unit supplied by specialised Vendor.
Reference is made to Section C, para. 8.0 for the general information about the technology.
A small portion of the syngas entering the unit bypasses the PSA and is sent to the post firing system of the HRSG together with the PSA off gas to make the burners flame stable.

The interfaces of the process are the following, as shown in the scheme:

1. Total clean syngas from AGR
2. Bypass to post firing
3. Hydrogen
4. Offgas to post firing



The main process data of the hydrogen production unit are summarised in following table:

	1 Syngas	2 By pass	3 H_2	4 Offgas
Hydrogen	85.35	85.35	99.50	43.73
Nitrogen	4.93	4.93	0.40	18.25
Argon	0.72	0.72	0.10	2.54
Carbon Monoxide	3.74	3.74		14.74
Carbon Dioxide	5.24	5.24		20.65
Methane	0.00	0.00		0.00
Water	0.02	0.02		0.08
Hydrogen Sulfide	0.00	0.00		0.00
	100.00	100.00	100.00	100.00
Flow (Nm^3/h)	274,296	5,149	200,858	68,289
Flow (kmol/h)	12,238	230	8,961	3,047
(kg/h)	82,571	1,550	19,303	61,717
p (barg)	26.0	26.0	25.2	0.7
T (°C)	34	34	39	26

Unit 3000: Power Island

Reference is made to Section C, para. 9.0 for the general information about the technology.

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

- HP steam (160 barg) : steam imported from Syngas Treatment and Conditioning Line.
- MHP steam (70 barg) : steam imported from Gasification section.

- MP steam (40 barg) : steam exported to Syngas Treatment and Conditioning Line to meet the water requirement of the shift reaction. A small quantity of steam is also generated in the Gasification Island and in the Sulphur Recovery Unit.
- LP steam (6.5 barg): steam exported to the following Process Units: AGR, ASU, Utility and Offsite Unit. LP steam is also generated in the Syngas Treatment and Conditioning Line.
- VLP steam (3.2 barg): steam imported from Syngas Treatment and Conditioning Line.
- BFW : HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate : All the condensate recovered from the condensation of the steam utilised in the Process Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

The MHP saturated steam at 70 bar from the gasification island, is superheated in a dedicated coil and sent to a dedicated ST section where is expanded. The exhaust steam is mixed with the exhaust steam from the ST IP section and flows to the ST LP main section. This steam turbine is coupled to the same generator of the main steam turbine. A dedicated clutch allows isolating the smaller steam turbine during the start-up of the plant.

Flow rate of the above interfaces of the Plant are shown in table attached to para 4.3, Utility Consumption.

4.3 Utility Consumption

The utility consumption of the process / utility and offsite units are shown in the attached Tables.

FOSTER WHEELER

CLIENT: IEA GHG
PROJECT: Hydrogen and Electricity co-production
LOCATION: the Netherlands

REVISION	Rev.0				
DATE	July 2007				
ISSUED BY	LV				
CHECKED BY	PC				
APPROVED BY	SA				

UTILITY CONSUMPTION SUMMARY - CASE 4 - Optimum H_2/EE Production Plant

UNIT	DESCRIPTION UNIT	HP Steam 160 barg [t/h]	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS											
1000	**Gasification Section**		-317.4				390.9				41.3	32.2
2100	**Air Separation Unit**				16.8						16.8	
2200	**Syngas Treatment and Conditioning line**	-40.6		267.3	-75.5	-103.4	40.6	150.5	75.5	120.3	16.9	417.8
2300	**Acid Gas Removal**				82.4						82.4	
2400	**Sulphur Recovery (SRU) - Tail gas treatment (TGT)**			-0.7	-1.0			4.3	1.0		3.6	
3000	**POWER ISLANDS UNITS**	40.6	317.4	-266.6	-32.1	103.4	-431.5	-154.8	-76.5	-120.3		
4100 to 5300	**UTILITY and OFFSITE UNITS**				9.4						9.4	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	170.4	450.0

Note: Minus prior to figure means figure is generated

4.4 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex.

SHELL

Case 4 - Low Pressure gasification with CO_2 capture, separate removal of H_2S and CO_2

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	273.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWt**	**1962.5**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWt	1638.2
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWt	1467.2
Syngas treatment efficiency	%	89.6
Hydrogen production (99.5% purity)	Nm^3/h	200,858
Hydrogen Thermal Power (E)	MWt	599.0
Equivalent H_2 based combined cycle net efficiency	%	56.0
Gas turbines total power output	MWe	286.0
Steam turbine power output	MWe	232.1
Equivalent Electric Power from H_2	MWe	335.4
ACTUAL GROSS ELECTRIC POWER	**MWe**	**518.1**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**853.5**
ASU power consumption	MWe	99.1
Process Units consumption	MWe	48.0
Utility Units consumption	MWe	2.5
Offsite Units consumption (including sea cooling water system)	MWe	7.5
Power Islands consumption	MWe	11.3
CO_2 compression and Drying	MWe	32.6
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**201.0**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**317.1**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**652.5**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**43.5**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**33.3**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**16.2**
Net H_2 output efficiency (E/A*100) (based on coal LHV)	**%**	**30.5**
H_2 thermal power Net Electric power generated ratio (E/B)		**1.89**

The following Table shows the overall CO_2 removal efficiency of the IGCC Complex.

	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	14701
Slag (Carbon =~0.4% wt) *	61
Net Carbon flowing to Process Units (A)	14640
Liquid Storage	
CO	24
CO_2	12434
Total to storage (B)	12458
Emission	
CO_2	2177
CO	6
Total Emission	2183
Overall CO_2 removal efficiency, % (B/A)	85.1

* The percentage of unreacted C stated by Shell is 0.2%. However, the carbon mass balance of the whole IGCC results in a 0.4% carbon less. This value is conservatively assumed.

4.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown in Section B - para 2.0, and co-produce electric power and hydrogen. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

4.5.1 Gaseous Emissions

Main Emissions

In normal operation at full load, the main continuous emissions are the combustion flue gas of single train of the Power Island, proceeding from the combustion of the Syngas in one gas turbine, and the emission from the coal Drying process.

The next table summarises expected flow rate and concentration of the combustion flue gas from the Power Island train.

Table 4.5 – Expected gaseous emissions from the Power Island train.

	Normal Operation	
Wet gas flow rate, kg/s	716.0	
Flow, Nm^3/h(1)	3,195,400	
Temperature, °C	129	
Composition	(%vol)	
Ar	0.97	
N_2	73.07	
O_2	8.80	
CO_2	2.32	
H_2O	14.84	
Emissions	mg/Nm^3 (1)	kg/h
NOx	73	233.6
SOx	1.6	5
CO	31	99
Particulate	5	16

(1) Dry gas, O_2 content 15%vol

In normal operation at full load, the following emission to the atmosphere is foreseen from the Coal Drying Process:

Flow rate	:	39	t/h
N_2	:	80	% vol.
$H_2O+O_2+CO_2$	:	20	% vol.
Particulate	:	<10	mg/Nm^3, wet basis.

Minor Emissions

The remainder gaseous emissions within the IGCC Complex are created by process vents and fugitive emissions.
Some of the vent points emit continuously; others during process upsets or emergency conditions only. All vent streams containing, potentially, undesirable gaseous components are sent to a flare system. Venting via the flare will be minimal during normal operation, but will be significant during emergencies, process upsets, start up and shutdown.

A small continuous emission is generated in the Waste Water Treatment plant; in fact a small burner is installed to destroy the biogas stream coming from the anaerobic section of the plant.

Fugitive emissions are those emissions caused by storage and handling of materials (solids transfer, leakage, etc.). Proper design and operation prevent them.

4.5.2 Liquid Effluent

Waste Water Treatment (Unit 4600)

Part of the effluent from the Waste Water Treatment (Unit 4600) is recovered and recycled back to the gasification island as process water, closing the Gasification water balance. The other part is sent to a dedicated treatment where the reverse osmosis process allows recovering almost 60% of the treated water. This recovered water is recycled back to the Demi Water System, Unit 4200, and used as raw water for the Demineralised water plant. The remaining 40% of water is discharged together with the sea cooling water return stream. The expected flow rate of this stream is as follows:

• Flow rate	:	46	m^3/h

Sea Water System (Unit 4100)

Sea water in open circuit is used for cooling.
The return stream Water is treated with meta-bisulphite in the Dechlorination System to reduce the Cl_2 concentration. Main characteristics of the water are listed in the following:

• Maximum flow rate	:	92,010	m^3/h
• Temperature	:	19	°C
• Cl_2	:	<0.05	ppm

4.5.3 Solid Effluent

The process does not produce any solid waste, except for typical industrial plant waste e.g. (sludge from Waste Water Treatment etc.). In any case, the waste water sludge (expected flow rate: 2 m^3/h) can be recovered, recycled back to the Gasification Island and burned into the Gasifier.

In addition, the Gasification Island is expected to produce the following solid by-products:

Slag from Slag Removal Unit

Flow rate	:	40.5	t/h
Water content	:	10	%wt

Slag product can be sold to be commercially used as major components in concrete mixtures to make road, pads, storage bins.

Flyash from Dry Solids Removal Unit

Flow rate : 1.3 t/h

Flyash can be dispatched to cement industries.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : CASE 5 – H_2 / ELECTRIC ENERGY FLEXIBLE PRODUCTION PLANT

ISSUED BY : L.VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L.Valota	P. Cotone	S. Arienti

SECTION G.5 HYDROGEN AND ELECTRICITY COPRODUTION BASIC INFORMATION FOR EACH ALTERNATIVE

I N D E X

SECTION G.5 BASIC INFORMATION FOR EACH ALTERNATIVE

5.0 Case 5

5.1 Introduction

The main features of the Case 5 configuration of the IGCC Complex are:

- Low pressure (39 bar g) Shell Gasification;
- Coal Nitrogen Dry Feed;
- Gasifier Heat Recovery Type;
- Double stage dirty shift;
- Separate removal of H_2S and CO_2;
- PSA unit for Hydrogen production with eventual Off-Gas Compression
- Gas Turbine (9FA)

The separate removal of acid gases, H_2S and CO_2, is based on the Selexol process.
The degree of integration between the Air Separation Unit (ASU) and the GT is 15%. Gas Turbine NO_x emission reduction is achieved diluting the syngas with compressed N_2 from ASU.
The Sulphur Recovery (SRU) is an O_2 assisted Claus Unit, with Tail gas catalytic treatment (SCOT type) and recycle of the treated tail gas to AGR.
Since this plant has been design to satisfy a wide range of hydrogen and net electricity production ratio, performance parameter and consumption will be shown at the maximum and at the minimum value of the ratio.

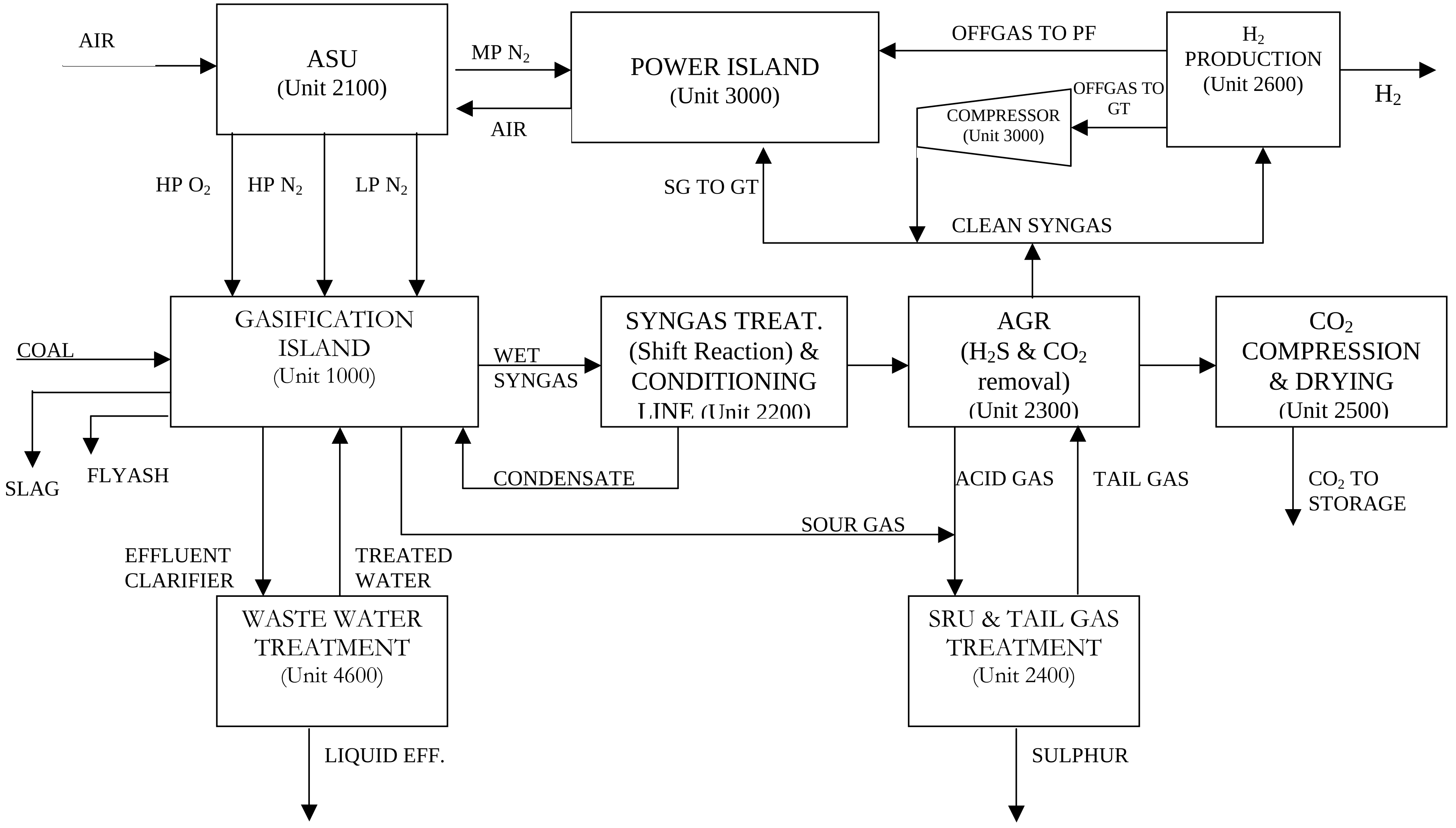
The arrangement of the process units is:

Unit		Trains
900	Coal milling and drying	4 x 33 %
1000	Coal pressurization/feeding	6 x 20 %
	Gasification heat recovery	2 x 50 %
	Slag removal	2 x 50 %
	Dry solids removal	2 x 50 %
	Wet scrubbing	2 x 50 %
	Sour slurry and sour water stripper	1 x 100 %
2100	ASU	2 x 50%
2200	Syngas Treatment and Conditioning Line	2 x 50%

2300	AGR	2 x 50%
2400	SRU	2 x 100%
	TGT	1 x 100%
2500	CO_2 Compression and Drying	2 x 50%
2600	H_2 production	1 x 100%
3000	Gas Turbine (PG9351FA)	1 x 100%
	HRSG	1 x 100%
	Steam Turbine	1 x 100%

Reference is made to the attached Block Flow Diagram of the IGCC Complex.

CASE 5 – IGCC COMPLEX BLOCK FLOW DIAGRAM



5.2 Process Description

Unit 1000: Gasification Island

Shell Gasification Island relevant information are collected in para. 1.1 of Section C.
The following table summarised the main process data of the Gasification Island for this alternative.

STREAM	FUEL FEED (COAL)	HP OXYGEN	HP NITROGEN	LP NITROGEN	SATURATED SYNGAS
Temperature (°C)	AMB.	80	80	70	160
Pressure (bar)		40	69	7.5	37
TOTAL FLOW					
Mass flow (kg/h)	273,100	214,550	87,000	33,680	568,200
Molar flow (kmol/h)			3,100	1,200	28,850
Composition (% vol)					
H_2					26.25
CO					49.60
CO_2					1.24
N_2		3.5	99.88	99.88	4.00
Ar		1.5	0.08	0.08	0.62
O_2		95.0	0.04	0.04	0.00
H_2S + COS					0.23
H_2O					18.05
Others					0.01

Unit 2100: Air Separation Unit (ASU)

This Unit is treated as a package supplied by specialised Vendors. For a general description of the Air Separation Unit refer to Section C, para. 3.0

The integration value between ASU and Gas Turbine is the percentage of the air extracted from the GT and sent to ASU over the total air required by ASU. It has been fixed to a value of 15%. Thus, when the gasification operates, the air required by the ASU to obtain the design oxygen production is entirely derived from self-standing compressor units.

The main process data and the main consumption of the ASU for both low hydrogen production - high electricity production (low R value) and high hydrogen production - low electricity production (high R value) are summarised in following tables.

Case low H_2 production and high electricity production (low R value)		
Process Data	Mass Flow (kg/h)	
Air from ambient	804,250	
Air from GT	141,900	
Oxygen to gasifier (95% vol)	214,550	
LP Nitrogen to Gasification Island (98% vol)	33,700	
HP Nitrogen to Gasification Island (98% vol)	87,000	
Nitrogen to Power Island (for syngas dilution)	304,350	
Consumption		
Main air compressor	64,500	kW
Oxygen compressor	11,000	kW
Nitrogen compressor	22,200	kW
Miscellanea	1,400	kW
Total	99,100	kW

Case high H_2 production and low electricity production (high R value)		
Process Data	Mass Flow (kg/h)	
Air from ambient	804,250	
Air from GT	141,900	
Oxygen to gasifier (95% vol)	214,550	
LP Nitrogen to Gasification Island (98% vol)	33,700	
HP Nitrogen to Gasification Island (98% vol)	87,000	
Nitrogen to Power Island (for syngas dilution)	248,550	
Consumption		
Main air compressor	64,500	kW
Oxygen compressor	11,000	kW
Nitrogen compressor	19,600	kW
Miscellanea	1,000	kW
Total	96,100	kW

Unit 2200: Syngas Treatment and Conditioning Line

Saturated raw syngas from wet scrubbing in Unit 1000, at approximately 36 barg and 160°C, enters the Sour Shift section of Unit 2200. The syngas is first heated in a gas/gas exchanger by the hot shift effluent and then enters the Shift Reactor, where CO is shifted to H_2 and CO_2 and COS is converted to H_2S. The exothermic shift reaction brings the syngas temperature up to 451°C. Due to the low water content of the syngas, the injection of MP steam to the syngas is required before entering the shift reactor. In order to meet the required degree of CO_2 removal, a double stage shift containing sulphur tolerant shift catalyst (dirty shift) is used. The hot shifted syngas outlet from the first stage is cooled in a series of heat exchangers:

Shift feed product exchanger
HP Steam Generator
MP Steam Generator

Inlet temperature to the second stage shift is controlled to 250°C. Outlet temperature from second shift is 331°C. The hot shifted syngas outlet from the second stage is cooled in a series of heat exchangers:

MP Steam Generator
LP Steam Generator
VLP Steam Generator
Condensate from CCU Preheater

The final cooling step of the syngas takes place in a cooling water cooler, where syngas is cooled with cooling water. Process condensate separated in syngas cooling is recycled back to the Sour Water Stripper of the Gasification Island.
Cold syngas flows to Unit 2300 and returns to Unit 2200, as clean syngas, after H_2S and CO_2 removal.
The syngas is then split in two streams. The first one is fed to the hydrogen production unit while the second stream is preheated with VLP steam, diluted with nitrogen from Air Separation Unit (Unit 2100) and sent to the gas turbine (Unit 3000). In the case of required low hydrogen production and high electricity production (low R value), the syngas sent to the hydrogen production unit consists of about 40% of the total syngas while in the case of high hydrogen production and low electricity production (high R value) it's around 61% of the total.

Unit 2300: Acid Gas Removal (AGR)

The removal of acid gases, H_2S and CO_2 is an important step of the IGCC operation. In fact this unit is not only capital intensive and a large consumer of energy, but also is a key factor for the control of the environmental performance of the IGCC. The right selection of the process and of the solvent used to capture the acid gases is important for the performance of the entire complex.

This Unit utilises Selexol as acid gas solvent.

Unit 2300 is characterised by a low syngas pressure (27 bar g) and an extremely high CO_2/H_2S ratio (205/1).

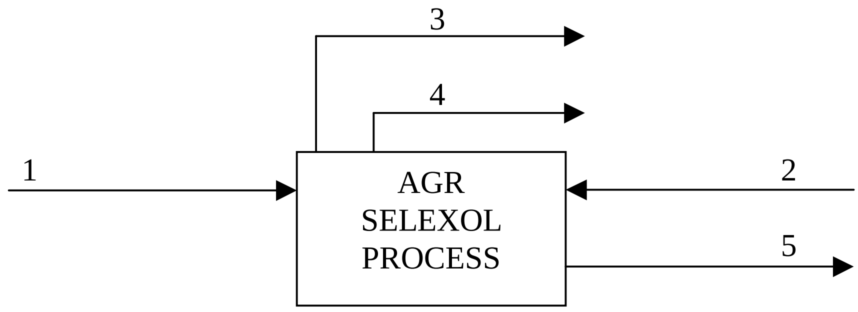
The interfaces of the process are the following, as shown in the scheme:

Entering Streams

1. Raw syngas from Syngas Treatment & Conditioning Line
2. Recycle Gas (Tail Gas) from Sulphur Recovery Unit.

Exit Streams

3. Treated Gas
4. CO_2 to compression
5. Acid Gas to Sulphur Recovery Unit



The main process data of the AGR unit are summarised in following table:

	1	2	3	4	5
	Raw SYNGAS from Syngas Treament	**Recycle Gas (tail gas) from SRU**	**Treated gas**	**CO_2 to compression**	**Acid gas to SRU**
Temperature (°C)	38	38	34	(1)	49
Pressure (bar)	27.8	27.0	27.0	(1)	1.8
Mass flow (kg/h)	714433	13011	164839	549273	13419
Molar flow (kgmole/h)	37113	332	24480	12728	336
Composition (vol %)					
H_2	56.51	4.10	85.35	1.74	0.28
CO	2.51	0.15	3.74	0.19	0.03
CO_2	36.91	76.63	5.24	97.69	72.41
N_2	3.10	17.78	4.93	0.06	0.01
CH_4	0.00	0.00	0.00	0.00	0.00
H_2S	0.18	0.72	0.00	0.01	20.25
COS	0.00	0.01	0.00	0.00	0.02
Ar	0.48	0.19	0.72	0.03	0.01
H2O	0.31	0.42	0.03	0.28	6.46

Note (1): CO_2 stream is the combination of three different streams at following pressure levels 26.0, 3.5 and 0.5 barg.

The Selexol solvent consumption, to make-up losses, is 120 m^3/year.

The proposed process matches the process specification with reference to H_2S+COS concentration of the treated gas exiting the Unit (H_2S+COS concentration is 3 ppm). This is due to the integration of CO_2 removal with the H_2S removal, which makes available a large circulation of the solvent that is cooled down by a refrigerant package (Power Consumption = 41% of the overall AGR Power requirement) before flowing to the CO_2 absorber.

The CO_2 removal rate is 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with large power consumption.

The acid gas H_2S concentration is 22% dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2 exiting the Unit, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1,7% vol and to an overall thermal power of 14.9 MWth, i.e. almost 5 MWe.
- A very low quantity of H_2S, corresponding to a concentration of about 100 ppmvd.

The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and H_2 at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

Unit 2400: SRU and TGT

This Unit is treated as a package supplied by specialised Vendors. For general information about the technology refer to Section C, para. 6.0

The Sulphur Recovery Section consists of two trains each having a normal sulphur production of 56.4 t/day, and normally operating at 50%.

The hydrogenated tail gas is recycled to Unit 2300, Acid Gas Removal, for the capture of H_2S by means of a compressor at a pressure of 27 barg.

Unit 2500: CO_2 Compression and Drying

This Unit is treated as a package supplied by specialised Vendors. For general information about the technology refer to Section C, para. 7.0

The incoming stream of Unit 2500 flows from Unit 2300, Acid Gas Removal, and is the combination of three different streams delivered at the following pressure levels:

- MP stream : 26.0 barg
- LP stream : 3.5 barg
- VLP stream : 0.5 barg

The product stream sent to final storage is mainly composed of CO_2 and CO. The main properties of the stream are as follows:

- Product stream : 550 t/h.
- Product stream : 110 bar.

- Composition :

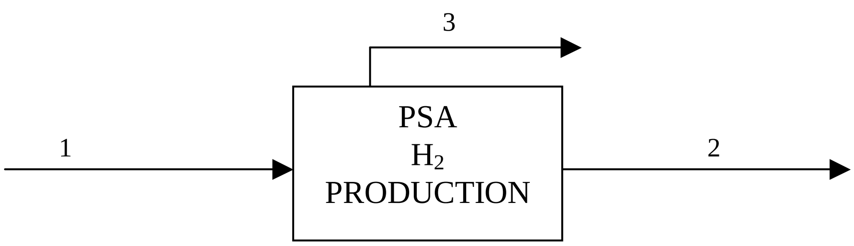
	%wt
CO_2	99.8
CO	0.1
Others	0.1
TOTAL	100.0

Unit 2600: H_2 Production

This Unit is treated as a package supplied by specialised Vendors. For general information about the technology refer to Section C, para. 8.0

The interfaces of the process are the following, as shown in the scheme:

1. Total clean syngas from AGR
2. Hydrogen
3. Offgas to post firing



The main process data of the hydrogen production unit for both low production of hydrogen and high production of electricity (low R value) as well as high production of hydrogen and low production of electricity (high R value) are summarised in following table:

Case low H_2 production and high electricity production (**low** R value)

	1 Syngas	**2 H_2**	**3 Offgas**
Hydrogen	85.35	99.50	43.73
Nitrogen	4.93	0.40	18.25
Argon	0.72	0.10	2.54
Carbon Monoxide	3.74		14.74
Carbon Dioxide	5.24		20.65
Methane	0.00		0.00
Water	0.02		0.08
Hydrogen Sulfide	0.00		0.00
Flow (Nm^3/h)	217,396	162,238	55,159
Flow (kmol/h)	9,699	7,238	2,461
(kg/h)	65,442	15,592	49,851
P (barg)	26.0	25.2	0.7
T (°C)	34	39	26

Case high H_2 production and low electricity production (**high** R value)

	1 Syngas	**2 H_2**	**3 Offgas**
Hydrogen	85.35	99.50	43.73
Nitrogen	4.93	0.40	18.25
Argon	0.72	0.10	2.54
Carbon Monoxide	3.74		14.74
Carbon Dioxide	5.24		20.65
Methane	0.00		0.00
Water	0.02		0.08
Hydrogen Sulfide	0.00		0.00
Flow (Nm^3/h)	329,848	246,157	83,690
Flow (kmol/h)	14,716	10,982	3,734
(kg/h)	99,293	23,657	75,637
P (barg)	26.0	25.2	0.7
T (°C)	34	39	26

In the case of high R value, the offgas is compressed in a compressor mixed with the clean syngas from AGR and sent to the Gas Turbine inlet (Unit 3000).

In the case of low R value, all the offgas is mixed with clean syngas from AGR and sent to the Post Firing (Unit 3000), thus there is no need of a clean syngas bypass to the HRSG post firing.

Gradually moving the production ratio between the two extremes, the offgas is split in two streams: the first is mixed with the bypass and sent to the Post Firing (Unit 3000) while the second is compressed in an external compressor, mixed with the clean syngas from AGR and sent to the Gas Turbine inlet (Unit 3000).

The offgas compression system is based on one integrally geared compressor unit with inter-cooling. Using inlet guide vanes it can control the quantity of delivered gas. Since the minimum deliverable gas is around 50% of the flowrate, a recirculation valve is included in order to be able to deliver gas even at low flow rates.

<u>Unit 3000: Power Island</u>

For general information about the Power Island technology refer to Section C, para. 9.0

For this configuration, the integration between the Process Units and the Power Island consists of the following interfaces:

<u>Imported</u>

- HP steam (160 barg) : steam imported from Syngas Treatment and Conditioning Line.
- MHP steam (70 barg) : steam imported from Gasification section.
- VLP steam (3.2 barg): steam imported from Syngas Treatment and Conditioning Line.
- Condensate from ST : All the Condensate from the Condenser is exported to the polishing unit (Unit 4200), pre-heated in the Syngas Treatment and Conditioning Line and recycled back to the HRSG.

<u>Exported</u>

- MP steam (40 barg): steam exported to Syngas Treatment and Conditioning Line to meet the water requirement of the shift reaction. A small quantity of steam is also generated in the Sulphur Recovery Unit and in the Tail Gas Treatment Unit.
- LP steam (6.5 barg): steam exported to the following Process Units: AGR, ASU, Utility and Offsite Unit. LP steam is also generated in the Syngas Treatment and Conditioning Line.

- BFW: HP, MP, LP, VLP Boiler Feed Water is exported to the Process Units to generate the above mentioned steam production.
- Process Condensate: All the condensate recovered from the condensation of the steam utilised in the Process Unit is recycled back to the HRSG after polishing in Unit 4200, Demi Water/Condensate Recovery.

The steam turbine in the Power Island consists of two sections: One High Pressure Steam turbine (HP ST) and one dedicated Medium High Pressure Steam turbine (MHP ST).

The HP saturated steam from the Syngas Treatment and Conditioning line (Unit 2200) is mixed with the HP steam generated in the coil, superheated and sent to HP ST where it's expanded. This turbine presents two extractions and one reheating.

The MHP saturated steam at 70 bar from the gasification island, is superheated in a coil, sent to the dedicated MHP ST and sent to the last stages of the HP steam turbine (LP Section) to be expanded down to condenser pressure.

Operative steam turbine pressures in the LP section are dependent from the R value that the plant is running. Essentially, for low R values, the post firing is maximum (250 MWth) and the steam turbine works at highest pressure and capacity. In the meanwhile, in case of high value of R, there is no post firing. Thus minimum steam production is perform, and the turbine operates in sliding pressure with consequently minimum turbine electricity production.

The thermal input of the post firing is delivered to the flue gas in two sections of the HRSG due to limits in the upper flue gas temperature.

Flow rate of the above interfaces are shown in table attached to para 5.3, Utility Consumption.

The off-gas compressor is included in Unit 3000

5.3 Utility Consumption

The utility consumption of the process / utility and offsite units of the Plant for both low production of hydrogen and high production of electricity (low R value) as well as high production of hydrogen and low production of electricity (high R value) are shown in the attached Tables.

CLIENT: IEA GHG
PROJECT: Hydrogen and Electricity co-production
LOCATION: the Netherlands

REVISION	Rev.0				
DATE	July 2007				
ISSUED BY	LV				
CHECKED BY	PC				
APPROVED BY	SA				

UTILITY CONSUMPTION SUMMARY - CASE G.5 - Shell gasification, with CO_2 capture, with H_2 production, with flexible H_2/EE ratio - LOW R VALUE

UNIT	DESCRIPTION UNIT	HP Steam 170 barg [t/h]	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS											
1000	**Gasification Section**		-317.4				390.9				41.3	32.2
2100	**Air Separation Unit**				16.8						16.8	
2200	**Syngas Treatment and Conditioning line**	-38.3		267.2	-75.5	-103.4	38.3	150.6	75.5	120.3	16.9	417.8
2300	**Acid Gas Removal**				82.4						82.4	
2400	**Sulphur Recovery (SRU) - Tail gas treatment (TGT)**			-0.7	-1.0			4.3	1.0		3.6	
3000	**POWER ISLANDS UNITS**	38.3	317.4	-266.5	-32.1	103.4	-429.2	-154.9	-76.5	-120.3		
4100 to 5300	**UTILITY and OFFSITE UNITS**				9.4						9.4	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	170.4	450.0

Note: Minus prior to figure means figure is generated

FOSTER WHEELER	CLIENT:	IEA GHG
	PROJECT:	Hydrogen and Electricity co-production
	LOCATION:	the Netherlands

REVISION	Rev.0				
DATE	July 2007				
ISSUED BY	LV				
CHECKED BY	PC				
APPROVED BY	SA				

UTILITY CONSUMPTION SUMMARY - CASE G.5 - Shell gasification, with CO_2 capture, with H_2 production, with flexible H_2/EE ratio - HIGH R VALUE

UNIT	DESCRIPTION UNIT	HP Steam 110 barg [t/h]	MHP Steam 70 barg [t/h]	MP Steam 40 barg [t/h]	LP Steam 6.5 barg [t/h]	VLP Steam 3.2 barg [t/h]	HP BFW [t/h]	MP BFW [t/h]	LP BFW [t/h]	VLP BFW [t/h]	condensate recovery [t/h]	Losses [t/h]
	PROCESS UNITS											
1000	**Gasification Section**		-317.4				390.9				41.3	32.2
2100	**Air Separation Unit**				16.8						16.8	
2200	**Syngas Treatment and Conditioning line**	-66.6		297.2	-75.5	-103.4	66.6	120.6	75.5	120.3	16.9	417.8
2300	**Acid Gas Removal**				82.4						82.4	
2400	**Sulphur Recovery (SRU) - Tail gas treatment (TGT)**			-0.7	-1.0			4.3	1.0		3.6	
3000	**POWER ISLANDS UNITS**	66.6	317.4	-296.5	-32.1	103.4	-457.5	-124.9	-76.5	-120.3		
4100 to 5300	**UTILITY and OFFSITE UNITS**				9.4						9.4	
	BALANCE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	170.4	450.0

Note: Minus prior to figure means figure is generated

5.4 IGCC Overall Performance

The following Table shows the overall performance of the IGCC Complex running at low H_2 production and high electricity production (low value of R) as well as high production of hydrogen and low production of electricity (high R value).

CASE 5 - R LOW

Shell gasification, with CO_2 capture, with H_2 production, with flexible H2/EE ratio

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	273.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWth**	**1962.5**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWth	1638.2
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWth	1467.2
Syngas treatment efficiency	%	89.6
Hydrogen production (99.5% purity)	Nm^3/h	162,240
Hydrogen Thermal Power (E)	MWth	484.0
Equivalent H_2 based combined cycle net efficiency	%	56.0
Gas turbines total power output	MWe	286.0
Steam turbine power output	MWe	279.0
Equivalent Electric Power from H_2	MWe	271.0
ACTUAL GROSS ELECTRIC POWER OUTPUT	**MWe**	**565.0**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**836.0**
ASU power consumption	MWe	99.1
Process Units consumption	MWe	48.0
Utility Units consumption	MWe	2.6
Offsite Units consumption (including sea cooling water system)	MWe	7.6
Power Islands consumption	MWe	12.0
CO_2 compression and Drying	MWe	32.6
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**201.9**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**363.1**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**634.1**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**42.6**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**32.3**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**18.5**
Net H_2 output efficiency (E/A*100) (based on coal LHV)	**%**	**24.7**
H_2 thermal power Net Electric power generated ratio (E/B)		**1.33**

CASE 5 - R HIGH

Shell gasification, with CO_2 capture, with H_2 production, with flexible H_2/EE ratio

OVERALL PERFORMANCES OF THE IGCC COMPLEX		
Coal Flowrate (fresh, air dried basis)	t/h	273.1
Coal LHV (air dried basis)	kJ/kg	25869.5
THERMAL ENERGY OF FEEDSTOCK (based on coal LHV) (A)	**MWth**	**1962.5**
Thermal Power of Raw Syngas exit Scrubber (dry, based on LHV)	MWth	1638.2
Gasification Efficiency (based on coal LHV)	%	83.5
Thermal Power of Clean Syngas (based on LHV)	MWth	1467.2
Syngas treatment efficiency	%	89.6
Hydrogen production (99.5% purity)	Nm^3/h	246,160
Hydrogen Thermal Power (E)	MWth	734.1
Equivalent H_2 based combined cycle net efficiency	%	56.0
Gas turbines total power output	MWe	286.0
Steam turbine power output	MWe	157.4
Equivalent Electric Power from H_2	MWe	411.1
ACTUAL GROSS ELECTRIC POWER OUTPUT	**MWe**	**443.4**
EQUIVALENT IGCC GROSS ELECTRIC POWER OUTPUT (D)	**MWe**	**854.5**
ASU power consumption	MWe	96.1
Process Units consumption	MWe	48.0
Utility Units consumption (including compressor)	MWe	15.1
Offsite Units consumption (including sea cooling water system)	MWe	7.6
Power Islands consumption	MWe	7.4
CO_2 compression and Drying	MWe	32.6
ELECTRIC POWER CONSUMPTION OF IGCC COMPLEX	**MWe**	**206.8**
NET ELECTRIC POWER OUTPUT (B)	**MWe**	**236.6**
EQUIVALENT NET ELECTRIC POWER OUTPUT OF IGCC (C)	**MWe**	**647.7**
Equivalent Gross electrical efficiency (D/A *100) (based on coal LHV)	**%**	**43.5**
Equivalent Net electrical efficiency (C/A*100) (based on coal LHV)	**%**	**33.0**
Net electrical efficiency (B/A*100) (based on coal LHV)	**%**	**12.1**
Net H_2 output efficiency (E/A*100) (based on coal LHV)	**%**	**37.4**
H_2 thermal power Net Electric power generated ratio (E/B)		**3.10**

The following tables show the overall CO_2 removal efficiency of the IGCC Complex for low R value and high R value:

LOW R VALUE	
	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	14701
Slag (Carbon =~0.4% wt) *	61
Net Carbon flowing to Process Units (A)	14640
Liquid Storage	
CO	24
CO_2	12434
Total to storage (B)	12458
Emission	
CO_2	2179
CO	4
Total Emission	2183
Overall CO_2 removal efficiency, % (B/A)	85.1

* The percentage of unreacted C stated by Shell is 0.2%. However, the carbon mass balance of the whole IGCC results in a 0.4% carbon less. This value is conservatively assumed.

HIGH R VALUE	
	Equivalent flow of CO_2, kmol/h
Coal (Carbon=82.5%wt)	14701
Slag (Carbon =~0.4% wt) *	61
Net Carbon flowing to Process Units (A)	14640
Liquid Storage	
CO	24
CO_2	12434
Total to storage (B)	12458
Emission	
CO_2	2180
CO	3
Total Emission	2183
Overall CO_2 removal efficiency, % (B/A)	85.1

* The percentage of unreacted C stated by Shell is 0.2%. However, the carbon mass balance of the whole IGCC results in a 0.4% carbon less. This value is conservatively assumed.

5.5 Environmental Impact

The IGCC Complex is designed to process coal, whose characteristics are shown in Section B - para 2.0, and co-produce electric power and hydrogen. The advanced technology allows to reach a high efficiency and to minimise environmental impact.

5.5.1 Gaseous Emissions

Main Emissions

In normal operation at full load, the main continuous emissions are the combustion flue gas of single train of the Power Island, proceeding from the combustion of the Syngas in one gas turbine, and the emission from the coal Drying process.

Next tables summarize expected flow rate and concentration of the combustion flue gas from the Power Island train for low R value and high R value.

LOW R VALUE		
	Normal Operation	
Wet gas flow rate, kg/s	716.2	
Flow, Nm^3/h(1)	3,196,000	
Temperature, °C	130	
Composition	(%vol)	
Ar	0.97	
N_2	73.08	
O_2	8.80	
CO_2	2.31	
H_2O	14.84	
Emissions	mg/Nm^3 (1)	kg/h
NOx	76.4	245
SOx	1.6	5
CO	32.5	104
Particulate	5	16

(1) Dry gas, O_2 content 15%vol

HIGH R VALUE		
	Normal Operation	
Wet gas flow rate, kg/s	697.6	
Flow, Nm^3/h(1)	2,025,400	
Temperature, °C	129	
Composition	(%vol)	
Ar	0.91	
N_2	73.74	
O_2	11.17	
CO_2	2.41	
H_2O	11.77	
Emissions	mg/Nm^3 (1)	kg/h
NOx	74	184.3
SOx	2.5	5
CO	31	78
Particulate	5	10.1

(1) Dry gas, O_2 content 15%vol

In normal operation at full load, the following emission to the atmosphere is foreseen from the Coal Drying Process:

Flow rate	39	t/h
N_2	80	% vol.
$H_2O+O_2+CO_2$	20	% vol.
Particulate	<10	mg/Nm^3, wet basis.

Minor Emissions

The remainder gaseous emissions within the IGCC Complex are created by process vents and fugitive emissions.
Some of the vent points emit continuously; others during process upsets or emergency conditions only. All vent streams containing, potentially, undesirable gaseous components are sent to a flare system. Venting via the flare will be minimal during normal operation, but will be significant during emergencies, process upsets, start up and shutdown.

A small continuous emission is generated in the Waste Water Treatment plant; in fact a small burner is installed to destroy the biogas stream coming from the anaerobic section of the plant.

Fugitive emissions are those emissions caused by storage and handling of materials (solids transfer, leakage, etc.). Proper design and operation prevent them.

5.5.2 Liquid Effluent

Waste Water Treatment (Unit 4600)

Part of the effluent from the Waste Water Treatment (Unit 4600) is recovered and recycled back to the gasification island as process water, closing the Gasification water balance. The other part is sent to a dedicated treatment where the reverse osmosis process allows recovering almost 60% of the treated water. This recovered water is recycled back to the Demi Water System, Unit 4200, and used as raw water for the Demineralised water plant. The remaining 40% of water is discharged together with the sea cooling water return stream. The expected flow rate of this stream is as follows:

- Flow rate: 46 m^3/h

Sea Water System (Unit 4100)

Sea water in open circuit is used for cooling.
The return stream Water is treated with meta-bisulphite in the Dechlorination System to reduce the Cl_2 concentration. Main characteristics of the water are listed in the following:

- Maximum flow rate: 93.600 m^3/h
- Temperature : 19 °C
- Cl_2 : <0.05 ppm

5.5.3 Solid Effluent

The process does not produce any solid waste, except for typical industrial plant waste e.g. (sludge from Waste Water Treatment etc.). In any case, the waste water sludge (expected flow rate: 2 m^3/h) can be recovered, recycled back to the Gasification Island and burned into the Gasifier.

In addition, the Gasification Island is expected to produce the following solid by-products:

Slag from Slag Removal Unit

Flow rate	:	40.5	t/h
Water content	:	10	%wt

Slag product can be sold to be commercially used as major components in concrete mixtures to make road, pads, storage bins.

Flyash from Dry Solids Removal Unit

Flow rate	:	1.3	t/h

Flyash can be dispatched to cement industries.

CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION ECONOMICS

ISSUED BY : L.VALOTA
CHECKED BY : P. COTONE
APPROVED BY : S. ARIENTI

Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L.Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L.Valota	P. Cotone	S. Arienti

SECTION H

Hydrogen and Electricity Co-Production Economics

I N D E X

SECTION H ECONOMICS

1.0 Introduction

This section summarises the economic data evaluated for each alternative of the study described in section G, including:

a. Investment cost;
b. Operation & Maintenance costs;
c. Electric power production cost.

2.0 Basis of Investment Cost Evaluation

2.1 Basis of the Estimate

The basis of the estimate for each alternative is the technical documentation collected in Sections C and G of this Study.

In particular the investment cost of the following Units or blocks of Units is detailed:

Unit 900	:	Coal Handling and Storage
Unit 1000	:	Gasification Section
Unit 2100	:	Air Separation Unit
Unit 2200	:	Syngas Treatment and Conditioning Line
Unit 2300	:	Acid Gas Removal
Unit 2400	:	Sulphur Recovery Unit and Tail Gas Treatment
Unit 2500	:	CO_2 Compression and Drying
Unit 2600	:	H_2 Production Unit
Unit 3000	:	Power Island
Units 4000 to 5200	:	Utilities and Offsites

The overall investment cost of each Unit or block of Units is split into the following items:

- Direct Materials, including equipment and bulk materials;
- Construction, including mechanical erection, instrument and electrical installation, civil works and, where applicable, buildings and site preparation;
- Other Costs, including temporary facilities, solvents, catalysts, chemicals, training, commissioning and start-up costs, spare parts etc.;
- EPC Services including Contractor's home office services and construction supervision.

2.2 Estimate Methodology and Cost Basis

Estimate methodology and cost basis are the same as described in Section E

3.0 Investment Cost of the Alternatives

As shown in section G, the following alternatives have been considered.

Case 1 consists of a electric energy production plant, without CO_2 capture and without hydrogen production, based on Shell gasification (Section G1).

Case 2 consists of a electric energy production plant, with CO_2 capture and without hydrogen production, based on Shell gasification (Section G2).

Case 3 consists of a electric energy production plant, with CO_2 capture and with maximum hydrogen production, based on Shell gasification (Section G3)

Case 4 consists of a electric energy production plant, with CO_2 capture and with hydrogen production at a specific ratio, based on Shell gasification (Section G4).

Case 5 consists of a flexible electric energy production plant, with CO_2 capture and with hydrogen production, based on Shell gasification (Section G5).

3.1 Case 1

The following Table H.3.1 shows the investment break down and the total figures for the case 1.

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Table H.3.1 - ESTIMATE SUMMARY

SHELL CASE 1

Client : IA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		900 €	1000 €	2100 €	2200 €	2300 €	2400 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
1	DIRECT MATERIALS		37,923,160	129,779,000	102,833,000	11,651,094	10,208,016	13,701,870	287,750,000	138,658,270	732,504,410	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		12,230,100	58,683,179	23,335,181	4,371,200	4,906,400	4,663,600	65,056,200	57,685,900	230,931,759	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		2,137,500	7,041,781	3,005,888	3,416,200	6,626,600	1,401,700	26,022,000	10,132,000	59,783,669	
4	EPC SERVICES		5,691,100	29,341,089	11,272,079	2,619,000	2,287,500	1,502,200	20,818,000	20,265,800	93,796,768	**900 Coal Handling & Storage**
												1000 Gasification Section
			______	______	______	______	______	______	______	______	______	**2100 Air Separation Unit**
												2200 Syngas Treat.&Condt. Line
A	**Installed costs** (contingency excluded)		**57,981,860**	**224,845,049**	**140,446,147**	**22,057,494**	**24,028,516**	**21,269,370**	**399,646,200**	**226,741,970**	**1,117,016,606**	**2300 Acid Gas Removal**
												2400 SRU & TGT
B	Contingency	%	7	7	5	7	7	7	7	5	6.3	**2500 CO2 Compression&Drying**
		Euro	4,058,730	15,739,153	7,022,307	1,544,025	1,681,996	1,488,856	27,975,234	11,337,099	70,847,400	**3000 Power Island**
												4000+ Utilities&Offsites
C	Fees (2% of A)		1,159,637	4,496,901	2,808,923	441,150	480,570	425,387	7,992,924	4,534,839	22,340,332	
D	Land Purchases; surveys (5% of A)		2,899,093	11,242,252	7,022,307	1,102,875	1,201,426	1,063,469	19,982,310	11,337,099	55,850,830	
			______	______	______	______	______	______	______	______	______	
	TOTAL INVESTMENT COST		**66,099,320**	**256,323,356**	**157,299,685**	**25,145,543**	**27,392,508**	**24,247,082**	**455,596,668**	**253,951,007**	**1,266,055,169**	

3.2 Case 2

The following Table H.3.2 shows the investment break down and the total figures for the case 2.

Table H.3.2 - ESTIMATE SUMMARY

SHELL CASE 2

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		UNIT 900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
1	DIRECT MATERIALS		40,041,100	137,377,000	110,062,100	29,828,000	56,040,900	24,727,200	24,871,400	285,710,000	170,783,500	879,441,200	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		12,913,100	62,118,800	24,975,600	10,893,300	26,118,100	8,416,500	4,703,200	64,595,000	71,051,000	285,784,700	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		2,256,900	7,454,000	3,217,200	15,242,600	24,007,000	2,529,300	908,500	25,838,000	12,479,000	93,932,500	
4	EPC SERVICES		6,008,900	31,058,900	12,064,500	6,524,800	12,181,900	2,710,000	1,272,400	20,671,000	24,961,100	117,453,500	
A	**Installed costs** (contingency excluded)		**61,220,000**	**238,008,800**	**150,319,500**	**62,488,700**	**118,347,900**	**38,383,000**	**31,755,500**	**396,814,000**	**279,274,600**	**1,376,611,900**	
B	Contingency	%	7	7	7	7	7	7	7	7	7	7.0	
		Euro	4,285,400	16,660,600	7,516,000	4,374,200	8,284,400	2,686,800	1,587,800	27,777,000	13,963,700	87,135,800	
C	Fees (2% of A)		1,224,400	4,760,200	3,006,400	1,249,800	2,367,000	767,700	635,100	7,936,300	5,585,500	27,532,200	
D	Land Purchases; surveys (5% of A)		3,061,000	11,900,400	7,516,000	3,124,400	5,917,400	1,919,200	1,587,800	19,840,700	13,963,700	68,830,600	
	TOTAL INVESTMENT COST		**69,790,800**	**271,330,000**	**168,357,800**	**71,237,100**	**134,916,600**	**43,756,700**	**35,566,200**	**452,368,000**	**312,787,500**	**1,560,110,600**	

900 **Coal Handling & Storage**
1000 **Gasification Section**
2100 **Air Separation Unit**
2200 **Syngas Treat.&Condt. Line**
2300 **Acid Gas Removal**
2400 **SRU & TGT**
2500 **CO2 Compression&Drying**
3000 **Power Island**
4000+ **Utilities&Offsites**

3.3 Case 3

The following Table H.3.3 shows the investment break down and the total figures for the case 3.

FOSTER WHEELER

Table H.3.3 - ESTIMATE SUMMARY

SHELL CASE 3

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		UNIT 900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	2600 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
1	DIRECT MATERIALS		40,041,100	137,377,000	103,787,000	29,828,000	56,040,900	24,727,200	24,871,400	11,964,000	90,813,000	130,838,200	650,287,800	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		12,913,100	62,118,800	23,551,700	10,893,300	26,118,100	8,416,500	4,703,200	5,982,000	20,531,500	54,432,500	229,660,700	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		2,256,900	7,454,000	3,033,800	15,242,600	24,007,000	2,529,300	908,500	717,800	8,213,000	9,561,000	73,924,000	
4	EPC SERVICES		6,008,900	31,058,900	11,376,700	6,524,800	12,181,900	2,710,000	1,272,400	5,264,200	6,570,000	19,122,800	102,090,500	**900 Coal Handling & Storage**
														1000 Gasification Section
			_______	_______	_______	_______	_______	_______	_______	_______	_______	_______	_______	**2100 Air Separation Unit**
														2200 Syngas Treat.&Condt. Line
A	**Installed costs** (contingency excluded)		**61,220,000**	**238,008,800**	**141,749,100**	**62,488,700**	**118,347,900**	**38,383,000**	**31,755,500**	**23,928,000**	**126,127,500**	**213,954,500**	**1,055,963,000**	**2300 Acid Gas Removal**
														2400 SRU & TGT
B	Contingency	%	7	7	7	7	7	7	7	7	7	7	7.000	**2500 CO2 Compression&Drying**
		Euro	4,285,400	16,660,600	7,087,500	4,374,200	8,284,400	2,686,800	1,587,800	1,675,000	8,828,900	10,697,700	66,168,200	**2600 Hydrogen production unit**
														3000 Power Island
C	Fees (2% of A)		1,224,400	4,760,200	2,835,000	1,249,800	2,367,000	767,700	635,100	478,600	2,522,600	4,279,100	21,119,300	**4000+ Utilities&Offsites**
D	Land Purchases; surveys (5% of A)		3,061,000	11,900,400	7,087,500	3,124,400	5,917,400	1,919,200	1,587,800	1,196,400	6,306,400	10,697,700	52,798,100	
			_______	_______	_______	_______	_______	_______	_______	_______	_______	_______	_______	
	TOTAL INVESTMENT COST		**69,790,800**	**271,330,000**	**158,759,000**	**71,237,100**	**134,916,600**	**43,756,700**	**35,566,200**	**27,277,900**	**143,785,400**	**239,629,000**	**1,196,048,600**	

3.4 Case 4

The following Table H.3.4 shows the investment break down and the total figures for the case 4.

FOSTER WHEELER

Table H.3.4 - ESTIMATE SUMMARY

SHELL CASE 4

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	2600 €	3000 €	UTIL&OFF €	TOTAL €	REMARKS
			UNIT											
1	DIRECT MATERIALS		40,041,100	137,377,000	102,096,500	29,828,000	56,040,900	24,727,200	24,871,400	8,009,500	169,134,000	146,343,300	738,469,000	1) ESTIMATE ACCURACY +/- 30%
2	CONSTRUCTION		12,913,100	62,118,800	23,168,100	10,893,300	26,118,100	8,416,500	4,703,200	4,004,800	38,238,800	60,883,100	251,457,700	2) TODAY COSTS (ESCALATION NOT INCLUDED)
3	OTHER COSTS		2,256,900	7,454,000	2,984,400	15,242,600	24,007,000	2,529,300	908,500	480,600	15,295,000	10,694,000	81,852,300	
4	EPC SERVICES		6,008,900	31,058,900	11,191,400	6,524,800	12,181,900	2,710,000	1,272,400	3,524,200	12,237,000	21,389,000	108,098,400	**900 Coal Handling & Storage**
														1000 Gasification Section
			———	———	———	———	———	———	———	———	———	———	———	**2100 Air Separation Unit**
														2200 Syngas Treat.&Condt. Line
A	**Installed costs** (contingency excluded)		**61,220,000**	**238,008,800**	**139,440,300**	**62,488,700**	**118,347,900**	**38,383,000**	**31,755,500**	**16,019,000**	**234,904,800**	**239,309,400**	**1,179,877,400**	**2300 Acid Gas Removal**
														2400 SRU & TGT
B	Contingency	%	7	7	7	7	7	7	7	7	7	7	7.0	**2500 CO2 Compression&Drying**
		Euro	4,285,400	16,660,600	6,972,000	4,374,200	8,284,400	2,686,800	1,587,800	1,121,300	16,443,300	11,965,500	74,381,300	**2600 Hydrogen production unit**
														3000 Power Island
C	Fees (2% of A)		1,224,400	4,760,200	2,788,800	1,249,800	2,367,000	767,700	635,100	320,400	4,698,100	4,786,200	23,597,500	**4000+ Utilities&Offsites**
D	Land Purchases; surveys (5% of A)		3,061,000	11,900,400	6,972,000	3,124,400	5,917,400	1,919,200	1,587,800	801,000	11,745,200	11,965,500	58,993,900	
			———	———	———	———	———	———	———	———	———	———	———	
	TOTAL INVESTMENT COST		**69,790,800**	**271,330,000**	**156,173,200**	**71,237,100**	**134,916,600**	**43,756,700**	**35,566,200**	**18,261,700**	**267,791,500**	**268,026,600**	**1,336,850,100**	

3.5 Case 5

The following Table H.3.5 shows the investment break down and the total figures for the case 5.

FOSTER WHEELER

Table H.3.5 - ESTIMATE SUMMARY

SHELL CASE 5

Client : IEA GREENHOUSE GASR & D PROGRAMME
Location : THE NETHERLANDS
Date : July 2007 REV. 1

FIGURE IN EURO

POS	DESCRIPTION		UNIT 900 €	1000 €	2100 €	2200 €	2300 €	2400 €	2500 €	2600 €	3000 €	UTIL&OFF €	TOTAL €
1	DIRECT MATERIALS		40,041,100	137,377,000	102,099,000	29,828,000	56,040,900	24,727,200	24,871,400	9,141,500	175,030,000	147,089,200	746,245,300
2	CONSTRUCTION		12,913,100	62,118,800	23,168,600	10,893,300	26,118,100	8,416,500	4,703,200	4,570,800	39,571,800	61,193,400	253,667,600
3	OTHER COSTS		2,256,900	7,454,000	2,984,400	15,242,600	24,007,000	2,529,300	908,500	548,500	15,829,000	10,748,000	82,508,300
4	EPC SERVICES		6,008,900	31,058,900	11,191,600	6,524,800	12,181,900	2,710,000	1,272,400	4,022,300	12,663,000	21,498,000	109,131,800
A	**Installed costs** (contingency excluded)		**61,220,000**	**238,008,800**	**139,443,700**	**62,488,700**	**118,347,900**	**38,383,000**	**31,755,500**	**18,283,000**	**243,093,800**	**240,528,600**	**1,191,553,000**
B	Contingency	%	7	7	7	7	7	7	7	7	7	7	7.0
		Euro	4,285,400	16,660,600	6,972,200	4,374,200	8,284,400	2,686,800	1,587,800	1,279,800	17,016,600	12,026,400	75,174,200
C	Fees (2% of A)		1,224,400	4,760,200	2,788,900	1,249,800	2,367,000	767,700	635,100	365,700	4,861,900	4,810,600	23,831,100
D	Land Purchases; surveys (5% of A)		3,061,000	11,900,400	6,972,200	3,124,400	5,917,400	1,919,200	1,587,800	914,200	12,154,700	12,026,400	59,577,600
	TOTAL INVESTMENT COST		**69,790,800**	**271,330,000**	**156,176,900**	**71,237,100**	**134,916,600**	**43,756,700**	**35,566,200**	**20,842,600**	**277,126,900**	**269,392,100**	**1,350,135,800**

REMARKS

1) ESTIMATE ACCURACY +/- 30%

2) TODAY COSTS (ESCALATION NOT INCLUDED)

900	**Coal Handling & Storage**
1000	**Gasification Section**
2100	**Air Separation Unit**
2200	**Syngas Treat.&Condt. Line**
2300	**Acid Gas Removal**
2400	**SRU & TGT**
2500	**CO2 Compression&Drying**
2600	**Hydrogen production unit**
3000	**Power Island**
4000+	**Utilities&Offsites**

4.0 Operation and Maintenance Cost of the Alternatives

Operating and Maintenance (O&M) costs include:

- Feedstock
- Chemicals
- Catalysts
- Solvents
- Raw Water make-up
- Direct Operating labour
- Maintenance
- Overhead Charges

O&M costs are generally allocated as variable and fixed costs.
Variable operating costs are directly proportional to the amount of kilowatt-hours and Hydrogen produced and are referred as incremental costs.
Fixed operating costs are essentially independent of the amount of products.
However, accurately distinguishing the variable and fixed operating costs is not always simple. Certain cost items may have both, variable and fixed, components; for instance the planned maintenance and inspection of the gas turbine, that are known to occur based on number of running hours.
In this study these costs have been considered fixed, assuming that the complex operates at design capacity and with the expected design service factor.

4.1 Variable Costs

The consumption of the various items and the corresponding costs are yearly, based on the expected equivalent availability of 7446 equivalent hours of operation in one year with syngas. Another 554 equivalent hours of operation of the power plant in one year with natural gas as back-up fuel is expected, provided the resulting greenhouse gas emissions are acceptable, but this operation has conservatively not been considered in the economic analysis.

The following Tables H.4.1/2/3/4/5 show the total yearly operating costs for the five alternatives.

FOSTER WHEELER

Client : IEA GHG
Date : July 2007 REV. 1

Table H.4.1 - Shell Case 1 Yearly Variable Costs

Yearly Operating hours =	7446	Shell - Case 1		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	Yearly t/y	**Oper. Costs (yearly basis)**
Feedstock				
Coal	31.0	250,600	1,865,968	57,844,996
Flux	15.0	7,767	57,833	867,496
Auxiliary feedstock				
Natural Gas (Flare)	113.0	75	558	63,105
Make-up water	0.1	224,000	1,667,904	166,790
Solvents				
MDEA	4500.0	8	62	280,119
Catalyst				74,633
Chemicals				1,218,974
Waste Disposal	7.0	37,200	276,991	1,938,938
TOTAL YEARLY OPERATING COSTS, Euro/year				**62,455,051**

FOSTER WHEELER	Client : IEA GHG Date : July 2007 REV. 1			
Table H.4.2 - Shell Case 2 Yearly Variable Costs				
Yearly Operating hours =	7446	**Shell - Case 2**		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	Yearly t/y	**Oper. Costs (yearly basis)**
Feedstock				
Coal	31.0	273,100	2,033,503	63,038,581
Flux	15.0	8,340	62,097	931,461
Auxiliary feedstock				
Natural Gas (Flare)	113.0	75	558	63,105
Make-up water	0.1	406,000	3,023,076	302,308
Solvents				
Selexol	6500	16.76	124.8	811,200
Catalyst				1,683,899
Chemicals				1,326,607
Waste Disposal	7.0	40,500	301,563	2,110,941
TOTAL YEARLY OPERATING COSTS, Euro/year				**70,268,101**

FOSTER WHEELER		Client : IEA GHG Date : July 2007 REV. 1		
Table H.4.3 - Shell Case 3 Yearly Variable Costs				
Yearly Operating hours =	7446	**Shell - Case 3**		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	 Yearly t/y	**Oper. Costs** **(yearly basis)**
Feedstock				
Coal	31.0	273,100	2,033,503	63,038,581
Flux	15.0	8,340	62,097	931,461
Auxiliary feedstock				
Natural Gas (Flare)	113.0	75	558	63,105
Make-up water	0.1	406,000	3,023,076	302,308
Solvents				
Selexol	6500	16.76	124.8	811,200
Catalyst				1,683,899
Chemicals				1,305,749
Waste Disposal	7.0	40,500	301,563	2,110,941
TOTAL YEARLY OPERATING COSTS, Euro/year				**70,247,243**

FOSTER WHEELER	Client : IEA GHG Date : July 2007 REV. 1		
Table H.4.4 - Shell Case 4 Yearly Variable Costs			

Yearly Operating hours =	7446	**Shell - Case 4**		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	 Yearly t/y	**Oper. Costs** **(yearly basis)**
Feedstock				
Coal	31.0	273,100	2,033,503	63,038,581
Flux	15.0	8,340	62,097	931,461
Auxiliary feedstock				
Natural Gas (Flare)	113.0	75	558	63,105
Make-up water	0.1	406,000	3,023,076	302,308
Solvents				
Selexol	6500	16.76	124.8	811,200
Catalyst				1,683,899
Chemicals				1,315,364
Waste Disposal	7.0	40,500	301,563	2,110,941
TOTAL YEARLY OPERATING COSTS, Euro/year				**70,256,858**

FOSTER WHEELER

Client : IEA GHG
Date : July 2007 REV. 1

Table H.4.5 - Shell Case 5 Yearly Variable Costs

Yearly Operating hours =	7446	Shell - Case 5		
Consumables	**Unit Cost** Euro/t	**Consumption** Hourly kg/h	Yearly t/y	**Oper. Costs (yearly basis)**
Feedstock				
Coal	31.0	273,100	2,033,503	63,038,581
Flux	15.0	8,340	62,097	931,461
Auxiliary feedstock				
Natural Gas (Flare)	113.0	75	558	63,105
Make-up water	0.1	406,000	3,023,076	302,308
Solvents				
Selexol	6500	16.76	124.8	811,200
Catalyst				1,683,899
Chemicals				1,320,246
Waste Disposal	7.0	40,500	301,563	2,110,941
TOTAL YEARLY OPERATING COSTS, Euro/year				**70,261,740**

4.2 Fixed Costs

Fixed costs have been evaluated following the same methodology of Section E.

The attached table H.4.6 shows the total maintenance costs for the five cases.

FOSTER WHEELER ITALIANA

Table H.4.6 - Maintenance Costs

Client : IEA GHG
Date : July 2007

Complex section	Maint %	Case 1 Capex Eurox10^3	Case 1 Maint. 10^3 Euro/y	Case 2 Capex Eurox10^3	Case 2 Maint. 10^3 Euro/y	Case 3 Capex Eurox10^3	Case 3 Maint. 10^3 Euro/y	Case 4 Capex Eurox10^3	Case 4 Maint. 10^3 Euro/y	Case 5 Capex Eurox10^3	Case 5 Maint. 10^3 Euro/y
ASU, AGR, SRU & TGT, CO_2 Comp., Coal St, H2 prod (Units: 900, 2100, 2300, 2400, 2500, 2600)	2.5	243,726	6,090	400,026	10,000	415,384	10,380	405,166	10,130	407,433	10,190
Gasification, Syngas Treat., (Units: 1000,2200)	4.0	246,903	9,880	300,497	12,020	300,497	12,020	300,497	12,020	300,497	12,020
Power Island (Unit: 3000)	5.0	399,646	19,982	396,814	19,841	126,128	6,306	234,905	11,745	243,094	12,155
Common facilities (Utilities, Offsite, etc.)	1.7	226,742	3,855	279,275	4,748	213,954	3,637	239,309	4,068	240,529	4,089
TOTAL		1,117,017	39,807	1,376,612	46,608	1,055,963	32,344	1,179,877	37,964	1,191,553	38,454
		Maint. % =	3.6	Maint. % =	3.4	Maint. % =	3.1	Maint. % =	3.2	Maint. % =	3.2

NOTES: (1) Including the Gas Turbine Long Term Service Agreement.

4.3 Summary

The following table H.4.7 summarizes the total Operating and Maintenance Costs on yearly basis for all the cases.

Table H.4.7 – Total O&M Costs

		Case 1 Euro/year	Case 2 Euro/year	Case 3 Euro/year	Case 4 Euro/year	Case 5 Euro/year
Fixed Costs	direct labor	6,400,000	6,400,000	6,400,000	6,400,000	6,400,000
	adm./gen overheads	1,920,000	1,920,000	1,920,000	1,920,000	1,920,000
	maintenance	39,807,000	46,608,000	32,344,000	37,964,000	38,454,000
	Subtotal	48,127,000	54,928,000	40,664,000	46,284,000	46,774,000
Variable Costs		62,455,000	70,268,000	70,247,000	70,257,000	70,262,000
TOTAL O&M COSTS		110,582,000	125,196,000	110,911,000	116,541,000	117,036,000

5.0 Evaluation of the Electric Power Cost of the alternatives

The following Tables summarize the economic analyses performed on each alternative in order to evaluate the electric power production cost, based on the following assumptions:

- 7446 equivalent operating hours of IGCC fed by syngas at 100% capacity;
- Total investment cost and O&M costs as evaluated in Section E;
- 10% discount rate on the investment cost over 25 operating years;
- No selling price is attributed to CO_2;
- Other financial parameters as per Project Design Basis, Section B, para. 2.7

The attached tables H.5.1/6 show the economical analysis for the alternatives G1, G2, G3, G4 and G5 (High and low). For case G3 the analysis has been based on the hydrogen production cost.

A sensitivity analysis with 5% discount rate on the investment cost is shown in table H.5.7/12.

The attached Table H.5.13 shows the economic analysis for alternatives and the sensitivity analysis.

TABLE H.5.1 - CASE G.1 - Cost Evaluation - Discount Rate = 10%

Production		
Coal Flowrate	250.6	t/h
Net Power Output	762.3	MW
Sold Sulphur	2.15	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost

(*) 1 USD= 1.00 Euro

Capital Expenditures	**MM Euro**	
Installed Costs		1117.0
Land purchase; surveys	5%	55.9
Fees	2%	22.3
Average Contingencies	6.3%	70.8
Total Investment Cost		1266.1

Operating Costs [MM Euro/year] **at 85% load factor**	
Fuel Cost	57.8
Maintenance	39.8
Waste Disposal (7€/t)	1.9
Chemicals + Consumable	2.7
Insurance and local taxes	22.3

Working Capital	**MM Euro**
30 days Chemical Storage	0.3
30 days Coal Storage	5.6
Total Working capital	5.9

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.052	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	294.7	MM Euro/year

NPV **0.00**
IRR **10.00%**

CASH FLOW ANALYSYS Millions Euro	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				155.1	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	293.0	
Sulphur				0.9	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	
Operating Costs																													
Fuel Cost				-30.6	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	
Maintenance				-26.5	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-1.4	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	
Waste Disposal				-1.0	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	
Insurance				-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	
Working Capital Cost				-5.9																									5.9
Fixed Capital Expenditures	-253.2	-569.7	-443.1																										
Total Cash flow (yearly)	-253.2	-569.7	-443.1	59.9	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	161.8	5.9
Total Cash flow (cumulated)	-253.2	-822.9	-1266.1	-1206.2	-1044.4	-882.6	-720.9	-559.1	-397.4	-235.6	-73.8	87.9	249.7	411.5	573.2	735.0	896.7	1058.5	1220.3	1382.0	1543.8	1705.5	1867.3	2029.1	2190.8	2352.6	2514.4	2676.1	2682.0
Discounted Cash Flow (Yearly)	-230.2	-470.8	-332.9	40.9	100.4	91.3	83.0	75.5	68.6	62.4	56.7	51.5	46.9	42.6	38.7	35.2	32.0	29.1	26.4	24.0	21.9	19.9	18.1	16.4	14.9	13.6	12.3	11.2	0.4
Discounted Cash Flow (Cumul.)	-230.2	-701.0	-1034.0	-993.1	-892.6	-801.3	-718.3	-642.8	-574.2	-511.9	-455.2	-403.6	-356.8	-314.2	-275.4	-240.2	-208.2	-179.1	-152.7	-128.6	-106.8	-86.9	-68.9	-52.4	-37.5	-23.9	-11.6	-0.4	0.0

TABLE H.5.2 - CASE G.2 - Cost Evaluation - Discount Rate = 10%

Production		
Coal Flowrate	273.1	t/h
Net Power Output	655.8	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost

(*) 1 USD= 1.00 Euro

Capital Expenditures		**MM Euro**
Installed Costs		1376.6
Land purchase; surveys	5%	68.8
Fees	2%	27.5
Average Contingencies	6.3%	87.1
Total Investment Cost		1560.1

Operating Costs [MM Euro/year] at 85% load factor	
Fuel Cost	63.0
Maintenance	46.6
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	27.5

Working Capital	**MM Euro**
30 days Chemical Storage	0.5
30 days Coal Storage	6.1
Total Working capital	6.6

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.072	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	351.9	MM Euro/year

NPV	**0.00**
IRR	**10.00%**

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				185.3	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	350.1	
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Operating Costs																													
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance				-31.1	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance				-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	
Working Capital Cost				-6.6																									6.6
Fixed Capital Expenditures	-312.0	-702.0	-546.0																										
Total Cash flow (yearly)	-312.0	-702.0	-546.0	75.6	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	199.1	6.6
Total Cash flow (cumulated)	-312.0	-1014.1	-1560.1	-1484.5	-1285.4	-1086.3	-887.1	-688.0	-488.8	-289.7	-90.5	108.6	307.7	506.9	706.0	905.2	1104.3	1303.5	1502.6	1701.7	1900.9	2100.0	2299.2	2498.3	2697.5	2896.6	3095.7	3294.9	3301.5
Discounted Cash Flow (Yearly)	-283.7	-580.2	-410.2	51.6	123.7	112.4	102.2	92.9	84.5	76.8	69.8	63.5	57.7	52.4	47.7	43.3	39.4	35.8	32.6	29.6	26.9	24.5	22.2	20.2	18.4	16.7	15.2	13.8	0.4
Discounted Cash Flow (Cumul.)	-283.7	-863.9	-1274.1	-1222.5	-1098.8	-986.4	-884.2	-791.3	-706.9	-630.1	-560.3	-496.9	-439.2	-386.7	-339.1	-295.7	-256.3	-220.5	-187.9	-158.3	-131.4	-107.0	-84.7	-64.5	-46.1	-29.4	-14.2	-0.4	0.0

Table H.5.3 - CASE G.3 - Cost Evaluation - Discount Rate = 10%

Production		
Coal Flowrate	273.1	t/h
Net Power Output	0.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	372,400	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1056.0
Land purchase; surveys	5%	52.8
Fees	2%	21.1
Average Contingencies	6.3%	66.2
Total Investment Cost		1196.0

Operating Costs [MM Euro/year] at 85% load factor	
Fuel Cost	63.0
Maintenance	32.3
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	21.1

Working Capital	**MM Euro**
30 days Chemical Storage	0.5
30 days Coal Storage	6.1
Total Working capital	6.6

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.072	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	284.9	MM Euro/year
Hydrogen price	0.102	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electric Energy				0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8
Hydrogen				150	283	283	283	283	283	283	283	283	283	283	283
Operating Costs															
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0
Maintenance				-21.6	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1
Insurance				-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1
Working Capital Cost				-6.6											
Fixed Capital Expenditures	-239.2	-538.2	-418.6												
Total Cash flow (yearly)	-239.2	-538.2	-418.6	56.0	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9
Total Cash flow (cumulated)	-239.2	-777.4	-1196.0	-1140.0	-987.1	-834.3	-681.4	-528.5	-375.7	-222.8	-69.9	82.9	235.8	388.7	541.5
Discounted Cash Flow (Yearly)	-217.5	-444.8	-314.5	38.3	94.9	86.3	78.4	71.3	64.8	58.9	53.6	48.7	44.3	40.3	36.6
Discounted Cash Flow (Cumul.)	-217.5	-662.3	-976.8	-938.5	-843.6	-757.3	-678.9	-607.5	-542.7	-483.8	-430.2	-381.5	-337.2	-297.0	-260.4

CASH FLOW ANALYSYS Millions Euro	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**
	13	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electric Energy	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	
Sulphur	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen	283	283	283	283	283	283	283	283	283	283	283	283	283	
Operating Costs														
Fuel Cost	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	
Working Capital Cost														6.6
Fixed Capital Expenditures														
Total Cash flow (yearly)	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	152.9	6.6
Total Cash flow (cumulated)	694.4	847.3	1000.1	1153.0	1305.9	1458.7	1611.6	1764.4	1917.3	2070.2	2223.0	2375.9	2528.8	2535.4
Discounted Cash Flow (Yearly)	33.3	30.2	27.5	25.0	22.7	20.7	18.8	17.1	15.5	14.1	12.8	11.7	10.6	0.4
Discounted Cash Flow (Cumul.)	-227.1	-196.8	-169.4	-144.4	-121.6	-101.0	-82.2	-65.1	-49.6	-35.5	-22.7	-11.0	-0.4	0.0

Production		
Coal Flowrate	273.1	t/h
Net Power Output	317.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	200,860	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1179.9
Land purchase; surveys	5%	59.0
Fees	2%	23.6
Average Contingencies	6.3%	74.4
Total Investment Cost		1336.85

Operating Costs [MM Euro/year]		
at 85% load factor		
Fuel Cost		63.0
Maintenance		38.0
Waste Disposal	(7€/t)	2.1
Chemicals + Consumable		5.1
Insurance and local taxes		23.6

Working Capital		**MM Euro**
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6
Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.071	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	310.9	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				88.4	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	167.0	
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen				75	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	
Operating Costs																													
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance				-25.3	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance				-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	
Working Capital Cost				-6.6																									6.6
Fixed Capital Expenditures	-267.4	-601.6	-467.9																										
Total Cash flow (yearly)	-267.4	-601.6	-467.9	63.6	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	170.8	6.6
Total Cash flow (cumulated)	-267.4	-869.0	-1336.9	-1273.3	-1102.5	-931.7	-761.0	-590.2	-419.4	-248.7	-77.9	92.9	263.6	434.4	605.1	775.9	946.7	1117.4	1288.2	1459.0	1629.7	1800.5	1971.3	2142.0	2312.8	2483.6	2654.3	2825.1	2831.7
Discounted Cash Flow (Yearly)	-243.1	-497.2	-351.5	43.4	106.0	96.4	87.6	79.7	72.4	65.8	59.9	54.4	49.5	45.0	40.9	37.2	33.8	30.7	27.9	25.4	23.1	21.0	19.1	17.3	15.8	14.3	13.0	11.8	0.4
Discounted Cash Flow (Cumul.)	-243.1	-740.2	-1091.8	-1048.3	-942.3	-845.9	-758.3	-678.6	-606.2	-540.4	-480.5	-426.1	-376.6	-331.7	-290.8	-253.6	-219.9	-189.1	-161.2	-135.8	-112.8	-91.8	-72.7	-55.4	-39.6	-25.3	-12.3	-0.4	0.0

FOSTER WHEELER

Table H.5.5 - CASE G.5 - LOW R - Cost Evaluation - Discount Rate = 10%

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Production		
Coal Flowrate	273.1	t/h
Net Power Output	363.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	162,240	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1191.6
Land purchase; surveys	5%	59.6
Fees	2%	23.8
Average Contingencies	6.3%	75.2
Total Investment Cost		1350.1

Operating Costs [MM Euro/year]		
at 85% load factor		
Fuel Cost		63.0
Maintenance		38.5
Waste Disposal	(7€/t)	2.1
Chemicals + Consumable		5.1
Insurance and local taxes		23.8

Working Capital		**MM Euro**
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.073	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	313.3	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electric Energy				104.2	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8
Hydrogen				61	115	115	115	115	115	115	115	115	115	115
Operating Costs														
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0
Maintenance				-25.6	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1
Insurance				-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8
Working Capital Cost				-6.6										
Fixed Capital Expenditures	-270.0	-607.6	-472.5											
Total Cash flow (yearly)	-270.0	-607.6	-472.5	64.3	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5
Total Cash flow (cumulated)	-270.0	-877.6	-1350.1	-1285.8	-1113.4	-940.9	-768.5	-596.0	-423.6	-251.1	-78.7	93.8	266.2	438.7
Discounted Cash Flow (Yearly)	-245.5	-502.1	-355.0	43.9	107.1	97.3	88.5	80.5	73.1	66.5	60.4	54.9	50.0	45.4
Discounted Cash Flow (Cumul.)	-245.5	-747.6	-1102.6	-1058.7	-951.6	-854.3	-765.8	-685.3	-612.2	-545.7	-485.3	-430.3	-380.4	-335.0

CASH FLOW ANALYSYS Millions Euro	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	12	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electric Energy	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	196.7	
Sulphur	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen	115	115	115	115	115	115	115	115	115	115	115	115	115	115	
Operating Costs															
Fuel Cost	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	
Working Capital Cost															6.6
Fixed Capital Expenditures															
Total Cash flow (yearly)	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	6.6
Total Cash flow (cumulated)	611.2	783.6	956.1	1128.5	1301.0	1473.4	1645.9	1818.3	1990.8	2163.2	2335.7	2508.1	2680.6	2853.0	2859.6
Discounted Cash Flow (Yearly)	41.3	37.5	34.1	31.0	28.2	25.6	23.3	21.2	19.3	17.5	15.9	14.5	13.2	12.0	0.4
Discounted Cash Flow (Cumul.)	-293.7	-256.1	-222.0	-191.0	-162.8	-137.2	-113.9	-92.7	-73.4	-55.9	-40.0	-25.5	-12.4	-0.4	0.0

Table H.5.6 - CASE G.5 - HIGH R - Cost Evaluation - Discount Rate = 10%

Production		
Coal Flowrate	273.1	t/h
Net Power Output	236.6	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	246,160	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		MM Euro
Installed Costs		1191.6
Land purchase; surveys	5%	59.6
Fees	2%	23.8
Average Contingencies	6.3%	75.2
Total Investment Cost		1350.1

Operating Costs [MM Euro/year] at 85% load factor	
Fuel Cost	63.0
Maintenance	38.5
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	23.8

Working Capital		MM Euro
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6
Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.078	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	10.00	%
Revenues / year	313.3	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**10.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electric Energy				72.7	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8
Hydrogen				92	174	174	174	174	174	174	174	174	174	174	174
Operating Costs															
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0
Maintenance				-25.6	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1
Insurance				-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8
Working Capital Cost				-6.6											
Fixed Capital Expenditures	-270.0	-607.6	-472.5												
Total Cash flow (yearly)	-270.0	-607.6	-472.5	64.3	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5
Total Cash flow (cumulated)	-270.0	-877.6	-1350.1	-1285.8	-1113.4	-940.9	-768.5	-596.0	-423.6	-251.1	-78.7	93.8	266.2	438.7	611.2
Discounted Cash Flow (Yearly)	-245.5	-502.1	-355.0	43.9	107.1	97.3	88.5	80.5	73.1	66.5	60.4	54.9	50.0	45.4	41.3
Discounted Cash Flow (Cumul.)	-245.5	-747.6	-1102.6	-1058.7	-951.6	-854.3	-765.8	-685.3	-612.2	-545.7	-485.3	-430.3	-380.4	-335.0	-293.7

CASH FLOW ANALYSYS Millions Euro	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	13	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electric Energy	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	137.4	
Sulphur	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen	174	174	174	174	174	174	174	174	174	174	174	174	174	
Operating Costs														
Fuel Cost	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	
Working Capital Cost														6.6
Fixed Capital Expenditures														
Total Cash flow (yearly)	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	172.5	6.6
Total Cash flow (cumulated)	783.6	956.1	1128.5	1301.0	1473.4	1645.9	1818.3	1990.8	2163.2	2335.7	2508.1	2680.6	2853.0	2859.6
Discounted Cash Flow (Yearly)	37.5	34.1	31.0	28.2	25.6	23.3	21.2	19.3	17.5	15.9	14.5	13.2	12.0	0.4
Discounted Cash Flow (Cumul.)	-256.1	-222.0	-191.0	-162.8	-137.2	-113.9	-92.7	-73.4	-55.9	-40.0	-25.5	-12.4	-0.4	0.0

Production		
Coal Flowrate	250.6	t/h
Net Power Output	762.3	MW
Sold Sulphur	2.15	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost

(*) 1 USD= 1.00 Euro

Capital Expenditures		**MM Euro**
Installed Costs		1117.0
Land purchase; surveys	5%	55.9
Fees	2%	22.3
Average Contingencies	6.3%	70.8
Total Investment Cost		1266.1

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	57.8
Maintenance	39.8
Waste Disposal (7€/t)	1.9
Chemicals + Consumable	2.7
Insurance and local taxes	22.3

Working Capital		**MM Euro**
30 days Chemical Storage		0.3
30 days Coal Storage		5.6
Total Working capital		5.9

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.040	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	5.00	%
Revenues / year	231.4	MM Euro/year

NPV	**0.00**
IRR	**5.00%**

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electric Energy				121.6	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7
Sulphur				0.9	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
Operating Costs														
Fuel Cost				-30.6	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8
Maintenance				-26.5	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-1.4	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7
Waste Disposal				-1.0	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9
Insurance				-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3
Working Capital Cost				-5.9										
Fixed Capital Expenditures	-253.2	-569.7	-443.1											
Total Cash flow (yearly)	-253.2	-569.7	-443.1	26.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4
Total Cash flow (cumulated)	-253.2	-822.9	-1266.1	-1239.7	-1141.2	-1042.8	-944.3	-845.9	-747.5	-649.0	-550.6	-452.1	-353.7	-255.2
Discounted Cash Flow (Yearly)	-241.2	-516.8	-382.8	21.7	77.1	73.5	70.0	66.6	63.5	60.4	57.6	54.8	52.2	49.7
Discounted Cash Flow (Cumul.)	-241.2	-757.9	-1140.7	-1119.0	-1041.9	-968.4	-898.4	-831.8	-768.3	-707.9	-650.3	-595.5	-543.3	-493.6

CASH FLOW ANALYSYS Millions Euro	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	12	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electric Energy	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	229.7	
Sulphur	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7	
Operating Costs															
Fuel Cost	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	-57.8	
Maintenance	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	-39.8	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	-2.7	
Waste Disposal	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	-1.9	
Insurance	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	-22.3	
Working Capital Cost															5.9
Fixed Capital Expenditures															
Total Cash flow (yearly)	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	98.4	5.9
Total Cash flow (cumulated)	-156.8	-58.3	40.1	138.6	237.0	335.5	433.9	532.3	630.8	729.2	827.7	926.1	1024.6	1123.0	1128.9
Discounted Cash Flow (Yearly)	47.4	45.1	43.0	40.9	39.0	37.1	35.3	33.7	32.1	30.5	29.1	27.7	26.4	25.1	1.4
Discounted Cash Flow (Cumul.)	-446.2	-401.1	-358.2	-317.3	-278.3	-241.2	-205.9	-172.2	-140.2	-109.7	-80.6	-52.9	-26.5	-1.4	0.0

FOSTER WHEELER

TABLE H.5.8 - CASE G.2 - Cost Evaluation - Discount Rate = 5%

Production

Coal Flowrate	273.1	t/h
Net Power Output	655.8	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost

(*) 1 USD= 1.00 Euro

Capital Expenditures	**MM Euro**	
Installed Costs		1376.6
Land purchase; surveys	5%	68.8
Fees	2%	27.5
Average Contingencies	6.3%	87.1
Total Investment Cost		1560.1

Operating Costs [MM Euro/year] at 85% load factor	
Fuel Cost	63.0
Maintenance	46.6
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	27.5

Working Capital	**MM Euro**
30 days Chemical Storage	0.5
30 days Coal Storage	6.1
Total Working capital	6.6

Labour Cost	**MM Euro/year**
# operators	128
Salary	0.05
Direct Labour Cost	6.4
Administration 30% L.C.	1.9
Total Labour Cost	8.3

Electricity Production Cost	0.056	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	5.00	%
Revenues / year	273.9	MM Euro/year
NPV	**0.00**	
IRR	**5.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				144.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	272.1	
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Operating Costs																													
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance				-31.1	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	-46.6	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance				-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	-27.5	
Working Capital Cost				-6.6																									6.6
Fixed Capital Expenditures	-312.0	-702.0	-546.0																										
Total Cash flow (yearly)	-312.0	-702.0	-546.0	34.3	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	121.2	6.6
Total Cash flow (cumulated)	-312.0	-1014.1	-1560.1	-1525.8	-1404.6	-1283.4	-1162.2	-1041.0	-919.8	-798.6	-677.5	-556.3	-435.1	-313.9	-192.7	-71.5	49.7	170.9	292.1	413.3	534.5	655.7	776.9	898.1	1019.3	1140.5	1261.6	1382.8	1389.4
Discounted Cash Flow (Yearly)	-297.2	-636.8	-471.7	28.2	95.0	90.4	86.1	82.0	78.1	74.4	70.9	67.5	64.3	61.2	58.3	55.5	52.9	50.4	48.0	45.7	43.5	41.4	39.5	37.6	35.8	34.1	32.5	30.9	1.6
Discounted Cash Flow (Cumul.)	-297.2	-933.9	-1405.6	-1377.4	-1282.5	-1192.0	-1105.9	-1023.9	-945.7	-871.3	-800.5	-733.0	-668.7	-607.5	-549.2	-493.7	-440.8	-390.5	-342.5	-296.8	-253.3	-211.9	-172.4	-134.9	-99.1	-65.0	-32.5	-1.6	0.0

Production		
Coal Flowrate	273.1	t/h
Net Power Output	0.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	372,400	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1056.0
Land purchase; surveys	5%	52.8
Fees	2%	21.1
Average Contingencies	6.27%	66.2
Total Investment Cost		1196.05

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	63.0
Maintenance	32.3
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	21.1

Working Capital		**MM Euro**
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6

Labour Cost		**MM Euro/year**
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.072	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	5.00	%
Revenues / year	225.0	MM Euro/year
Hydrogen price	0.080	Euro/Nm3
NPV	**0.00**	
IRR	**5.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electric Energy				0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8
Hydrogen				118	223	223	223	223	223	223	223	223	223	223	223
Operating Costs															
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0
Maintenance				-21.6	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1
Insurance				-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1
Working Capital Cost				-6.6											
Fixed Capital Expenditures	-239.2	-538.2	-418.6												
Total Cash flow (yearly)	-239.2	-538.2	-418.6	24.4	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0
Total Cash flow (cumulated)	-239.2	-777.4	-1196.0	-1171.7	-1078.7	-985.6	-892.6	-799.6	-706.6	-613.6	-520.5	-427.5	-334.5	-241.5	-148.5
Discounted Cash Flow (Yearly)	-227.8	-488.2	-361.6	20.0	72.9	69.4	66.1	63.0	60.0	57.1	54.4	51.8	49.3	47.0	44.7
Discounted Cash Flow (Cumul.)	-227.8	-716.0	-1077.6	-1057.6	-984.7	-915.3	-849.2	-786.2	-726.3	-669.1	-614.8	-563.0	-513.6	-466.7	-421.9

CASH FLOW ANALYSYS Millions Euro	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**
	13	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electric Energy	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	
Sulphur	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen	223	223	223	223	223	223	223	223	223	223	223	223	223	
Operating Costs														
Fuel Cost	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	-32.3	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	-21.1	
Working Capital Cost														6.6
Fixed Capital Expenditures														
Total Cash flow (yearly)	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	93.0	6.6
Total Cash flow (cumulated)	-55.5	37.6	130.6	223.6	316.6	409.6	502.7	595.7	688.7	781.7	874.7	967.8	1060.8	1067.4
Discounted Cash Flow (Yearly)	42.6	40.6	38.7	36.8	35.1	33.4	31.8	30.3	28.8	27.5	26.2	24.9	23.7	1.6
Discounted Cash Flow (Cumul.)	-379.3	-338.7	-300.1	-263.2	-228.2	-194.8	-163.0	-132.7	-103.9	-76.4	-50.2	-25.3	-1.6	0.0

FOSTER WHEELER

Table H.5.10 - CASE G.4 - Cost Evaluation - Discount Rate = 5%

Production		
Coal Flowrate	273.1	t/h
Net Power Output	317.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	200,860	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1179.9
Land purchase; surveys	5%	59.0
Fees	2%	23.6
Average Contingencies	6.3%	74.4
Total Investment Cost		1336.85

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	63.0
Maintenance	38.0
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	23.6

Working Capital	**MM Euro**	
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6
Labour Cost	**MM Euro/year**	
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.042	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	5.00	%
Revenues / year	244.1	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**5.00%**	

CASH FLOW ANALYSYS Millions Euro	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031
	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				53.0	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	100.2	
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen				75	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	142	
Operating Costs																													
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance				-25.3	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	-38.0	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance				-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	-23.6	
Working Capital Cost				-6.6																									6.6
Fixed Capital Expenditures	-267.4	-601.6	-467.9																										
Total Cash flow (yearly)	-267.4	-601.6	-467.9	28.2	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	103.9	6.6
Total Cash flow (cumulated)	-267.4	-869.0	-1336.9	-1308.7	-1204.7	-1100.8	-996.9	-893.0	-789.1	-685.2	-581.2	-477.3	-373.4	-269.5	-165.6	-61.7	42.3	146.2	250.1	354.0	457.9	561.8	665.8	769.7	873.6	977.5	1081.4	1185.3	1191.9
Discounted Cash Flow (Yearly)	-254.6	-545.7	-404.2	23.2	81.4	77.5	73.9	70.3	67.0	63.8	60.8	57.9	55.1	52.5	50.0	47.6	45.3	43.2	41.1	39.2	37.3	35.5	33.8	32.2	30.7	29.2	27.8	26.5	1.6
Discounted Cash Flow (Cumul.)	-254.6	-800.3	-1204.5	-1181.3	-1099.9	-1022.3	-948.5	-878.1	-811.1	-747.3	-686.6	-628.7	-573.6	-521.1	-471.1	-423.5	-378.2	-335.0	-293.9	-254.7	-217.4	-181.9	-148.1	-115.9	-85.2	-55.9	-28.1	-1.6	0.0

Table H.5.11 - CASE G.5 - LOW R - Cost Evaluation - Discount Rate = 5%

Production		
Coal Flowrate	273.1	t/h
Net Power Output	363.1	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	162,240	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1,191.6
Land purchase; surveys	5%	59.6
Fees	2%	23.8
Average Contingencies	6.3%	75.2
Total Investment Cost		1350.1

Operating Costs [MM Euro/year]		
at 85% load factor		
Fuel Cost		63.0
Maintenance		38.5
Waste Disposal	(7€/t)	2.1
Chemicals + Consumable		5.1
Insurance and local taxes		23.8

Working Capital	**MM Euro**	
30 days Chemical Storage		0.5
30 days Coal Storage		6.1
Total Working capital		6.6
Labour Cost	**MM Euro/year**	
# operators		128
Salary		0.05
Direct Labour Cost		6.4
Administration	30% L.C.	1.9
Total Labour Cost		8.3

Electricity Production Cost	0.048	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	5.00	%
Revenues / year	245.8	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**5.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electric Energy				68.4	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	129.2	
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen				61	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	115	
Operating Costs																													
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance				-25.6	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance				-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	
Working Capital Cost				-6.6																									6.6
Fixed Capital Expenditures	-270.0	-607.6	-472.5																										
Total Cash flow (yearly)	-270.0	-607.6	-472.5	28.6	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	6.6
Total Cash flow (cumulated)	-270.0	-877.6	-1350.1	-1321.6	-1216.6	-1111.7	-1006.7	-901.8	-796.9	-691.9	-587.0	-482.0	-377.1	-272.1	-167.2	-62.2	42.7	147.7	252.6	357.5	462.5	567.4	672.4	777.3	882.3	987.2	1092.2	1197.1	1203.7
Discounted Cash Flow (Yearly)	-257.2	-551.1	-408.2	23.5	82.2	78.3	74.6	71.0	67.6	64.4	61.4	58.4	55.7	53.0	50.5	48.1	45.8	43.6	41.5	39.6	37.7	35.9	34.2	32.5	31.0	29.5	28.1	26.8	1.6
Discounted Cash Flow (Cumul.)	-257.2	-808.2	-1216.4	-1193.0	-1110.7	-1032.4	-957.8	-886.8	-819.2	-754.7	-693.4	-634.9	-579.3	-526.3	-475.8	-427.7	-381.9	-338.3	-296.8	-257.2	-219.6	-183.7	-149.5	-117.0	-86.0	-56.5	-28.4	-1.6	0.0

Table H.5.12 - CASE G.5 - HIGH R - Cost Evaluation - Discount Rate = 5%

Production		
Coal Flowrate	273.1	t/h
Net Power Output	236.6	MW
Sold Sulphur	2.35	t/h
Fuel Price	31.0	Euro/t
Insurance and local taxes	2%	Installed cost
Hydrogen production	246,160	Nm3/h
(*) 1 USD= 1.00 Euro		

Capital Expenditures		**MM Euro**
Installed Costs		1,191.6
Land purchase; surveys	5%	59.6
Fees	2%	23.8
Average Contingencies	6.3%	75.2
Total Investment Cost		1350.1

Operating Costs [MM Euro/year]	
at 85% load factor	
Fuel Cost	63.0
Maintenance	38.5
Waste Disposal (7€/t)	2.1
Chemicals + Consumable	5.1
Insurance and local taxes	23.8

Working Capital	**MM Euro**
30 days Chemical Storage	0.5
30 days Coal Storage	6.1
Total Working capital	6.6

Labour Cost	**MM Euro/year**
# operators	128
Salary	0.05
Direct Labour Cost	6.4
Administration 30% L.C.	1.9
Total Labour Cost	8.3

Electricity Production Cost	0.040	Euro/kWh
Sulphur Price	103.3	Euro/t
Inflation	0.00	%
Taxes	0.00	%
Discount rate	5.00	%
Revenues / year	245.8	MM Euro/year
Hydrogen price	0.095	Euro/Nm3
NPV	**0.00**	
IRR	**5.00%**	

CASH FLOW ANALYSYS Millions Euro	**2003**	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**
	000	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electric Energy				37.0	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9
Sulphur				1.0	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8
Hydrogen				92	174	174	174	174	174	174	174	174	174	174	174
Operating Costs															
Fuel Cost				-33.4	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0
Maintenance				-25.6	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5
Labour				-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3
Chemicals & Consumables				-2.7	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1
Waste Disposal				-1.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1
Insurance				-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8
Working Capital Cost				-6.6											
Fixed Capital Expenditures	-270.0	-607.6	-472.5												
Total Cash flow (yearly)	-270.0	-607.6	-472.5	28.6	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9
Total Cash flow (cumulated)	-270.0	-877.6	-1350.1	-1321.6	-1216.6	-1111.7	-1006.7	-901.8	-796.9	-691.9	-587.0	-482.0	-377.1	-272.1	-167.2
Discounted Cash Flow (Yearly)	-257.2	-551.1	-408.2	23.5	82.2	78.3	74.6	71.0	67.6	64.4	61.4	58.4	55.7	53.0	50.5
Discounted Cash Flow (Cumul.)	-257.2	-808.2	-1216.4	-1193.0	-1110.7	-1032.4	-957.8	-886.8	-819.2	-754.7	-693.4	-634.9	-579.3	-526.3	-475.8

CASH FLOW ANALYSYS Millions Euro	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
	13	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electric Energy	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	69.9	
Sulphur	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	1.8	
Hydrogen	174	174	174	174	174	174	174	174	174	174	174	174	174	
Operating Costs														
Fuel Cost	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	-63.0	
Maintenance	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	-38.5	
Labour	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	-8.3	
Chemicals & Consumables	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	-5.1	
Waste Disposal	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	-2.1	
Insurance	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	-23.8	
Working Capital Cost														6.6
Fixed Capital Expenditures														
Total Cash flow (yearly)	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	104.9	6.6
Total Cash flow (cumulated)	-62.2	42.7	147.7	252.6	357.5	462.5	567.4	672.4	777.3	882.3	987.2	1092.2	1197.1	1203.7
Discounted Cash Flow (Yearly)	48.1	45.8	43.6	41.5	39.6	37.7	35.9	34.2	32.5	31.0	29.5	28.1	26.8	1.6
Discounted Cash Flow (Cumul.)	-427.7	-381.9	-338.3	-296.8	-257.2	-219.6	-183.7	-149.5	-117.0	-86.0	-56.5	-28.4	-1.6	0.0

Table H.5.13

ALTERNATIVE		**G1**	**G2**	**G3**	**G4**	**G5 Low**	**G5 High**
Discount rate		10%	10%	10%	10%	10%	10%
Coal Flowrate	[t/h]	250.6	273.1	273.1	273.1	273.1	273.1
Net Power Out.	[MWe]	762.3	655.8	0.1	317.1	363.1	236.6
Hydrogen Production	[Nm^3/h]	-	-	372,400	200,860	162,240	246,160
Total Inv. Cost	[10^6 Euro]	1266.1	1560.1	1196.0	1336.9	1350.1	1350.1
Hydrogen Cost	[$Euro/Nm^3$]	-	-	0.102	0.095	0.095	0.095
Revenues / year	[10^6 Euro/y]	294.7	351.9	284.9	310.9	313.3	313.3
Electricity Prod Cost	[Euro/kWh]	0.052	0.072	0.072	0.071	0.073	0.078

ALTERNATIVE		**G1**	**G2**	**G3**	**G4**	**G5 Low**	**G5 High**
Discount rate		5%	5%	5%	5%	5%	5%
Coal Flowrate	[t/h]	250.6	273.1	273.1	273.1	273.1	273.1
Net Power Out.	[MWe]	762.3	655.8	0.1	317.1	363.1	236.6
Hydrogen Production	[Nm^3/h]	-	-	372,400	200,860	162,240	246,160
Total Inv. Cost	[10^6 Euro]	1266.1	1560.1	1196.0	1336.9	1350.1	1350.1
Hydrogen Cost	[$Euro/Nm^3$]	-	-	0.080	0.095	0.095	0.095
Revenues / year	[10^6 Euro/y]	231.4	273.9	225.0	244.1	245.8	245.8
Electricity Prod Cost	[Euro/kWh]	0.040	0.056	0.072	0.042	0.048	0.040

6.0 Sensitivity to hydrogen pressure and purity

With reference to case 4 (electric energy production plant, with CO_2 capture and with hydrogen production at a specific ratio, based on Shell gasification (Section G4)) a sensitivity study has been performed in order to highlight the possible impact on PSA investment cost of a different hydrogen purity and pressure.

Hydrogen pressure

The hydrogen pressure in the reference case is 25 barg.
The pressure has been selected in order to match the lower gasification pressure among the three considered in section D and to avoid any compressor that would be necessary in case of higher hydrogen pressure was required.
The clean and shifted syngas available from Shell Gasification Island is around 26 barg.
In order to have Hydrogen at pressure higher than 25 barg, two alternatives could be selected:
A- To compress the syngas at PSA inlet (approx 80 t/h) at an adequate pressure (taking into account the PSA Unit pressure losses);
B- To compress the hydrogen at PSA outlet (approx 20 t/h) at the required pressure.

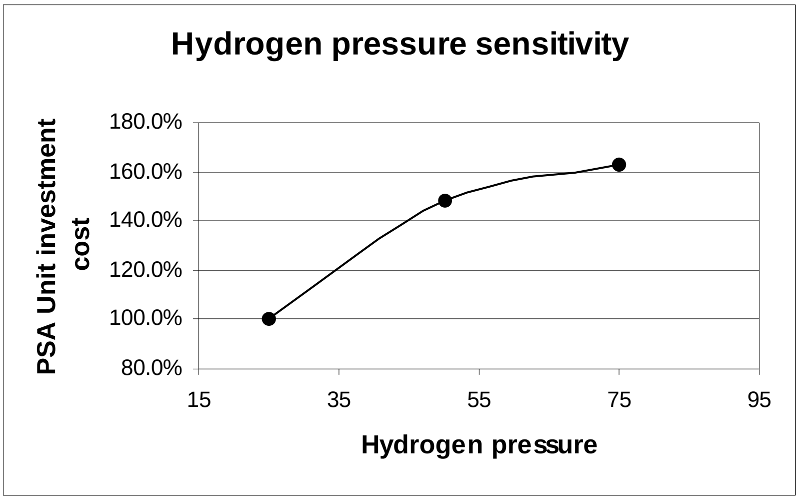
In alternative A the total flowrate to be compressed includes the syngas impurities that are successively discharged as PSA offgas.
Instead in alternative B the flowrate to be compressed is lower (around 25% as mass flow).
For this reason, alternative B is preferred.

The following Table H.6.1 and Figure H.6.1 show the hydrogen pressure percentage impact on investment cost considering 100% for reference case (Case 3 – 25 barg H_2 outlet pressure). In the table is also shown the power absorbed by the hydrogen compressor.

Table H.6.1

Pressure	Investment cost	Compressor Power absorption
25	100.0%	-
50	148.3%	6,275 kW
75	163.3%	10,250 kW

Figure H.6.1



As shown in the table, the sensitivity has been performed for three pressure values. The impact on cost is mainly due to the introduction of a compressor. For this reason, the higher impact is in the first step (25 barg to 50 barg). Successively, the investment cost increasing trend is lower as only the compressor incremental cost is considered.

In GEE case it would be different as the clean and shifted syngas is available at approx 55 barg. In the case described in section D1, the portion of syngas fed to the PSA unit is expanded down to 26 barg generating approx 5,600 kWe. In case it is necessary to export hydrogen at 50 barg it is sufficient to avoid the expansion abandoning the syngas expander. The impact is better than Shell case as the loss in power is lower (non production of 5,600 kWe vs. consumption of 6,275 kWe) and the investment cost is lower (abandoning the expander vs. introducing a compressor).

Hydrogen pressure impact on hydrogen cost

An estimation of the impact of hydrogen pressure on the cost of hydrogen has been performed. This consists of the capital related costs of increasing hydrogen pressure plus an estimation of the difference variable cost including extra O&M and extra cost of electricity consumption divided by the annual hydrogen output.

Capital costs are weighted on a 6 years as estimated payback time. In formulas, the difference in hydrogen cost is:

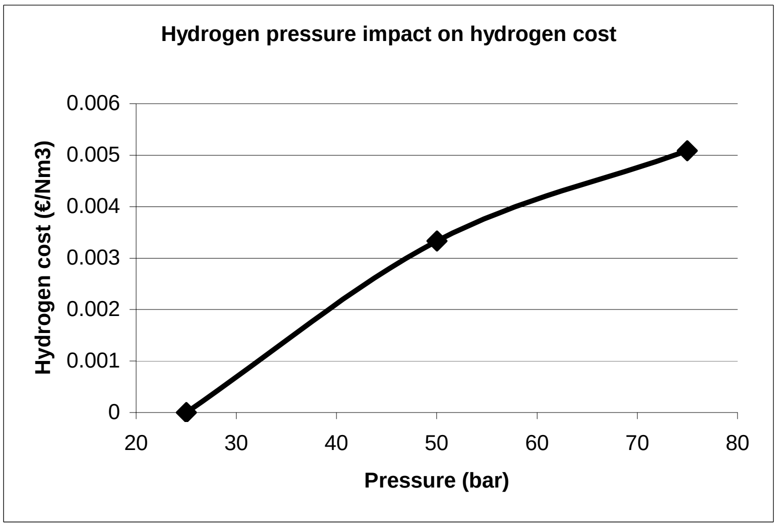
$$\Delta Hydrogen\_Cost = \frac{\frac{Extra\_Capex}{6} + Extra\_O\&M + Cost\_extra\_power\_cons}{Total\_hydrogen\_yearly\_production}$$

The following Table H.6.2 and Figure H.6.2 show the hydrogen cost impact on hydrogen cost.

Table H.6.2

Pressure	Extra Cost of hydrogen €/Nm^3
25	0
50	0.0033
75	0.0051

Figure H.6.2



Hydrogen purity

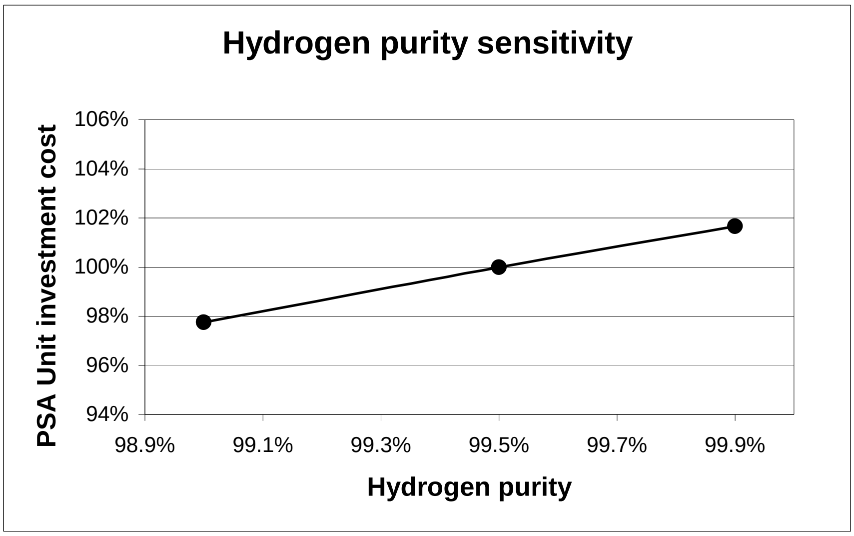
Hydrogen purity in the reference case is 99.5%.
The purity has been selected in order to match the average purity required by the different users considered in section J.
The sensitivity study has been performed based on rough cost evaluation provided by UOP for the PSA unit.

The following Table H.6.3 and Figure H.6.3 show the hydrogen purity percentage impact on investment cost considering 100% for reference case (Case 3 – 99.5% H_2 outlet purity).

Table H.6.3

H2 Purity	Investment cost
99.0%	97.8%
99.5%	100.0%
99.9%	101.7%

Figure H.6.3



As shown in the table, the sensitivity has been performed for three purity values centered in the reference one.

The trend of the investment cost is approximately linear in the range considered. The increase from 99.0% to 99.9% purity is less than 5% as the big issue in the PSA unit PSA is to achieve high purity of hydrogen (>99%); the difference in a so close range of value is not significantly high.

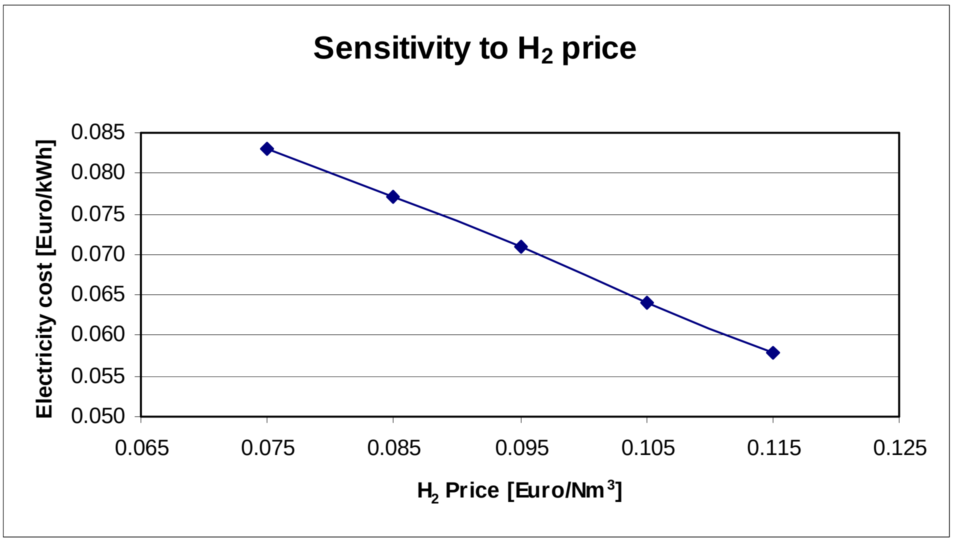
7.0 Sensitivity of electricity cost to hydrogen price

Tables H.7.1 and figure H.7.1 show the sensitivity analysis of electricity cost to hydrogen price for case G4.

Table H.7.1

Coal Flowrate	[t/h]	273.1	273.1	273.1	273.1	273.1
Net Power Out.	[MWe]	317.1	317.1	317.1	317.1	317.1
Hydrogen production	[MWe equiv]	335.4	335.4	335.4	335.4	335.4
Total Inv. Cost	[MM Euro]	1336.9	1336.9	1336.9	1336.9	1336.9
H_2 price	**[Euro/Nm^3]**	**0.075**	**0.085**	**0.095**	**0.105**	**0.115**
Revenues / year	[MM Euro/y]	310.9	310.9	310.9	310.9	310.9
Electricity Prod Cost	**[Euro/kWh]**	**0.083**	**0.077**	**0.071**	**0.064**	**0.058**

Figure H.7.1



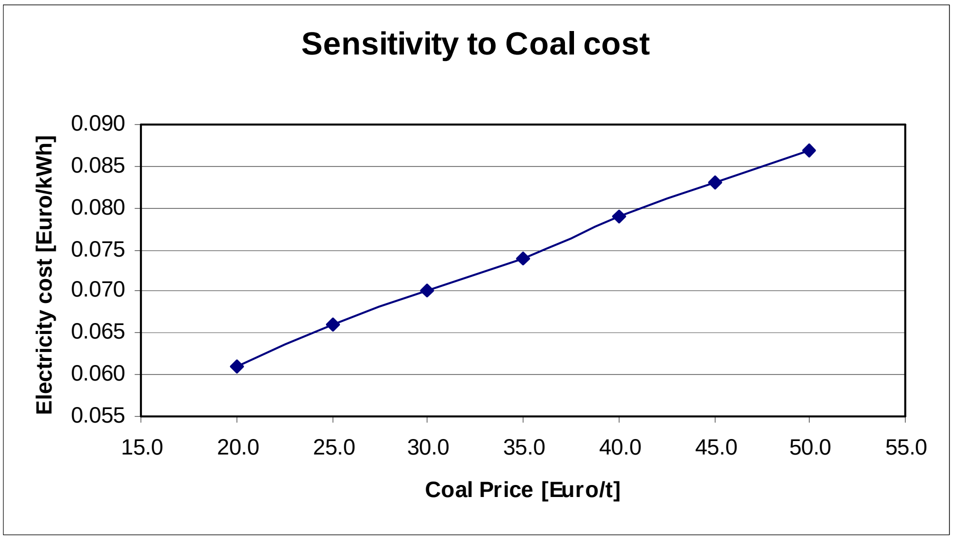
8.0 Sensitivity of electricity cost to coal price

Tables H.8.1 and figure H.8.1 show the sensitivity analysis of electricity cost to fuel price.

Table H.8.1

Coal Flowrate	[t/h]	273.1	273.1	273.1	273.1	273.1	273.1	273.1
Net Power Out.	[MWe]	317.1	317.1	317.1	317.1	317.1	317.1	317.1
Hydrogen production	[MWe equiv]	335.4	335.4	335.4	335.4	335.4	335.4	335.4
Total Inv. Cost	[MM Euro]	1336.9	1336.9	1336.9	1336.9	1336.9	1336.9	1336.9
Coal price	**[Euro/t]**	**20.0**	**25.0**	**30.0**	**35.0**	**40.0**	**45.0**	**50.0**
Revenues / year	[MM Euro/y]	288.3	298.6	308.9	319.1	329.4	339.6	349.9
Electricity Prod Cost	**[Euro/kWh]**	**0.061**	**0.066**	**0.070**	**0.074**	**0.079**	**0.083**	**0.087**

Figure H.8.1



CLIENT : IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME : HYDROGEN AND ELECTRICITY CO-PRODUCTION COMPARISON OF ALTERNATIVES

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Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L.Valota	P.Cotone	S.Arienti
July 2007	Rev. 1	L.Valota	P.Cotone	S.Arienti

SECTION I

HYDROGEN AND ELECTRICITY COPRODUCTION - COMPARISON OF ALTERNATIVES

INDEX

1.0 INTRODUCTION

In order to co-produce hydrogen and electricity, five different plants have been analyzed and their performances evaluated. These plants have been described in Section G and their costs estimated in Section E. A review of these data is present in paragraph 2.0 of the current section.

Moreover, under a specific hypothesis, the demand of energy and thermal energy deliverable by hydrogen has been computed for two different regions. Those analyses are present in Section J, Attachment A, "Analysis of Hydrogen and Electricity Demand".

That document forecasts the quantity of hydrogen that would be required to fulfill the demand of energy if the conventional fossil fuel systems were replaced by hydrogen systems based on the state of the art technology.

In the current section, five scenarios have been presented. Each scenario is a combination of the five possible plants. Behind each one of them, a specific criterion is present in order to fulfill the demand of energy.

These five criterions are presented in paragraph 3.0.

For each scenario, several overall outputs are provided, such as average annual outputs and load factors, overall coal consumptions, CO_2 outputs (emissions to the atmosphere and CO_2 captured), capital costs and operating costs and others.

A software program, presented in 4.0, has been compiled in order to systematically achieve the required output information on the basis of different energy consumption values and to allow changes in relevant input.

The results present in this study are relevant to the two regions considered (The Netherlands and USA) but, as explained before, the methodology outlined can be applied also to the consumption in different regions, only changing the software input.

The program has been run with two different kinds of data: considering the monthly data energy consumption and the intra-day data energy consumption. The monthly analysis concerned The Netherlands and USA data, while the intra-day analysis only The Netherlands.

2.0 PLANT ALTERNATIVES REVIEW

In Section G, five different types of electric energy and hydrogen gasification coproduction plant have been analyzed:

Case 1 – Plant type 1 consists of an electricity-only energy production plant, without hydrogen production and without CO_2 capture (Section G1).

Case 2 – Plant type 2 consists of a plant with the maximum electric energy production without hydrogen production, with CO_2 capture (Section G2).

Case 3 – Plant type 3 consists of a coproduction plant with the maximum hydrogen production and electric energy production only for internal electrical consumption, with CO_2 capture (Section G3).

Case 4 – Plant type 4 consists of a coproduction plant, with electricity and hydrogen production at a specific ratio and with CO_2 capture (Section G4).

Case 5 – Plant type 5 consists of a flexible coproduction plant with electricity and hydrogen production with CO_2 capture (Section G5).

Relevant data from Section G for each case are reported in Table I.2.0. All the considered plants have an availability factor (potential working hours a year over hours in a year) of 85%.

FOSTER WHEELER

OVERALL ECONOMICS
PERFORMANCE and COST SUMMARY - TABLE I.2.0

Date: July 2007
Rev: Rev. 1
Made by: FWI

			Case #1 plant	Case #2 plant	Case #3 plant	Case #4 plant	Case #5 plant-R low	Case #5 plant-R high
			w/o CO_2 capture, w/o H_2 production	CO_2 capture; No H_2 production	CO_2 capture; maximum H_2 production	CO_2 capture; H_2 production; optimum fixed H_2/EE ratio;	CO_2 capture; H_2 production; flexible H_2/EE ratio; R low	CO_2 capture; H_2 production; flexible H_2/EE ratio; R high
Gasification	Coal consumption	t/h	**250.6**	**273.1**	**273.1**	**273.1**	**273.1**	**273.1**
PSA	Hydrogen production (99.5% purity)	Nm^3/h	**n/a**	**n/a**	**372,400.0**	**200,858.0**	**162,240.0**	**246,160.0**
	Hydrogen Thermal Power (E)	MWt	**n/a**	**n/a**	**1,110.7**	**599.0**	**484.0**	**734.1**
Consumption	Electric power consumption of IGCC complex	MWe	**129.6**	**219.2**	**208.5**	**201**	**201.9**	**206.8**
Power Island	Gas turbines total power output	MWe	**553.6**	**572**	**87.6**	**286**	**286**	**286**
	Steam turbine power output	MWe	**338.3**	**303**	**121**	**232.1**	**279**	**157.4**
	Actual gross electric power output	MWe	**891.9**	**875**	**208.6**	**518.1**	**565**	**443.4**
	Net electric power output (B)	MWe	**762.3**	**655.8**	**0.10**	**317.1**	**363.1**	**236.6**
CO2 capture	Net Carbon flowing to process unit	kmol/h	**n/a**	**14640**	**14640**	**14640**	**14640**	**14640**
	CO_2 to Storage	kmol/h	**n/a**	**12458**	**12458**	**12458**	**12458**	**12458**
	CO_2 Emissions	kmol/h	**n/a**	**2183**	**2183**	**2183**	**2183**	**2183**
Sold Sulphur	Sulphur	t/h	**2.15**	**2.35**	**2.35**	**2.35**	**2.35**	**2.35**
Emissions	NO_X	kg/h	**453.6**	**371.2**	**83.6**	**233.6**	**245**	**184.3**
	SO_X	kg/h	**28.3**	**5**	**5**	**5**	**5**	**5**
	CO	kg/h	**176**	**155.5**	**36**	**99**	**104**	**78**
	Particulate	kg/h	**28**	**25.1**	**6.3**	**16**	**16**	**10.1**
Cost	Capital cost	EUR	**1,041,278,700**	**1,560,120,000**	**1,196,050,000**	**1,336,860,000**	**1,350,140,000**	**1,350,140,000**
	O&M fixed cost	EUR/y	**39,560,000**	**54,930,000**	**40,670,000**	**46,290,000**	**46,780,000**	**46,780,000**
	O&M variable cost	EUR/y	**62,455,000**	**70,270,000**	**70,250,000**	**70,260,000**	**70,270,000**	**70,270,000**
Avaibility	Availability Factor		**0.85**	**0.85**	**0.85**	**0.85**	**0.85**	**0.85**

3.0 SCENARIOS ALTERNATIVES

In the current study five different scenarios are considered. A scenario is a combinations of the plants described in Section 2.0 which satisfies the required energy demand.

The five scenarios are:

1- Electricity-only and H_2-only production plants
2- Non flexible co-production plants, without hydrogen storage
3- Non flexible co-production plants, with hydrogen storage
4- Flexible Coproduction plants, without hydrogen storage
5- Flexible Coproduction plants, with hydrogen storage

For each scenario, there is a single method of organizing the operation of the plants. These are listed below.

In scenario 1, electricity-only plants (plant 2) would be used to satisfy the peak electricity demand. When demand is low, some plants would be shut down. Similarly for hydrogen, hydrogen-only plants (plant 3) would meet the demand peak for hydrogen and would be shut down when the demand is lower.

In scenario 2, non flexible co-production plants (plant 4) are used to satisfy the minimum hydrogen or electricity demands, whichever is the smaller. Peaks in electricity and hydrogen demand will be satisfied by electricity-only plants (plant 2) and hydrogen-only plants (plant 3) respectively.

In scenario 3, non flexible co-production plants (plant 4) are used to satisfy the peak electricity demand. The variation in hydrogen demand is satisfied by storing hydrogen at times of low demand, for use at times of high demand. If the overall annual hydrogen demand is not the same as the overall annual production, some of the peak electricity demand or hydrogen demand will be satisfied by electricity-only plants (plant 2) or hydrogen-only plants (plant 3).

In scenario 4, flexible co-production plants (plant 5) are installed, thus the amount of hydrogen and electricity produced vary to enable the hydrogen and electricity demand to be satisfied. If there are any periods when the hydrogen or electricity production is beyond the demand, either hydrogen-only plants (plant 2) or electricity-only plants (plant 3) are installed.

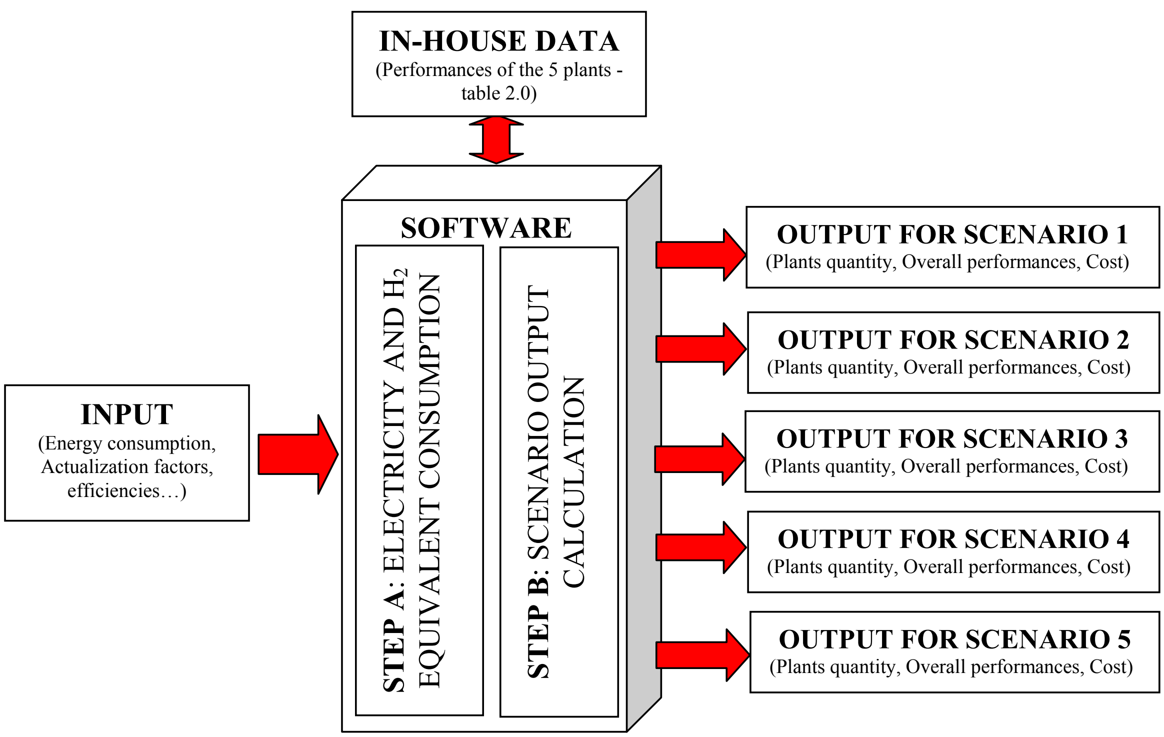
In scenario 5 flexible co-production plants (plant 5) are used to satisfy the energy demand and hydrogen storage is used to avoid the need for hydrogen-only and electricity-only plants.

Since the quantity of hydrogen to be stored in scenario 3 and 5 is high and the cheapest solution for this magnitude is underground storage, for the purpose of this study underground geological hydrogen storage has been considered. Refer to Section J, Attachment C for detail. Because the cost of underground storage widely varies and could strongly affect the final results, a sensitivity study has been also performed in paragraph 6.0.

4.0 SOFTWARE DESIGN AND DESCRIPTION

A software program has been developed in order to systematically achieve the output of each scenario and to allow eventual further studies (i.e. in different regions or different periods). It automatically combines the different plants and creates scenarios under the above criteria, computing the output values.

Two sections compose the software program: an input section and an output section.



The inputs data are:

- The monthly electrical energy consumption
- The percentage of the total energy produced from Nuclear and Renewable energy.
- The monthly natural gas consumption
- The percentage of the total consumed gas that is used for power generation and industrial chemical use.
- The monthly gasoline consumption used for transportation
- The monthly diesel consumption used for transportation
- The state-of-the-art fuel cell efficiency
- The state-of-the-art gasoline motors efficiency
- The state-of-the-art diesel motors efficiency
- Natural gas actualization factor.
- Gasoline actualization factor.
- Diesel actualization factor.
- Capital cost and operation cost of hydrogen storage (Euro/kg)

The performances of each plant shown in Section G and table 2.0 of this section are in-house data already integrated in the software and thus they do not have to be inserted by the user.

The output section consists of the following set of information, for each scenario:

- Number of each type of plant present in the scenario
- Monthly average load factors. This is the percentage of the time the plants are running when they are available. In other words it is the percentage of plants running at 100% when they are available.
- Max quantity of hydrogen present in storage (if present) and max quantity of hydrogen present in storage per each plant that includes hydrogen storage. These figures correspond to the required hydrogen storage volume. Eventual leakage from the storage is not considered
- Overall coal consumption
- Carbon dioxide capture and emission
- Plants capital cost (excluding storage)
- Underground storage capital cost (if applicable, including capital cost of extra compression of H_2 into the storage and extra PSA unit for purification of hydrogen removed from storage)
- Total scenario capital cost
- Total annual O&M cost

- Electricity production cost under the following main assumptions
  - 7446 (85% availability) equivalent operating hours in normal conditions at 100% capacity;
  - 10% discount rate on the investment cost over 25 operating years;
  - No selling price attributed to CO_2;
  - Hydrogen selling price 8.799 Euro/GJ.
- NO_x, SO_x, CO and particulate emissions based on monthly average

The capital and operating costs of the plants have been assumed to be equal for both the considered regions. In other words the costs are the same in the Netherlands as and in USA. The impact of this approximation has been considered not significant for the purpose of this study.

The program operates in two steps: step A (electricity and hydrogen equivalent consumption) and step B (scenario output calculation). In step A the program calculates the hydrogen equivalent consumption using the methodology described in Section J, Attachment A. This first step is a common methodology to all the scenarios. Step B uses the results from step A and computes the results for each scenario. Since the scenarios are widely different, different procedures have been used for each one. In particular:

Scenario 1 (Electricity-only and H_2-only production plants)

The program takes the demand of electricity and divides it by the energy production of one single plant type 2, finding the quantity of type 2 plants. It uses the same approach for hydrogen demand with plant type 3. Final performances of the calculated combination of plants 2 and 3 are provided.

Scenario 2 (Non flexible co-production plants, without hydrogen storage)

The program takes the demand of electricity and divides it by the energy production of one single plant type 4; then it takes the demand of hydrogen and divides it by the energy production of one single plant type 4 and takes the smallest between the two. Finally it adds plants type 2 or 3 in order to fulfill the request with the same methodology followed for scenario 1.

Scenario 3 (Non flexible co-production plants, with hydrogen storage)

1- The program takes the results from step A
2- It calculates the number of plants type 4 in order to do not have any excess of electricity (not storable). But the system may have hydrogen in excess or shortage. The hydrogen in excess is sent or taken from the storage for future usage.

3- The remaining request of hydrogen and electricity is satisfied installing plants type 2 and 3 and using the hydrogen previously sent to storage.

Scenario 4 (Flexible Coproduction plants, without hydrogen storage)

Plants type 5 are installed to follow the demand for each time period. The quantity has been selected in order to not have any excess of electricity (not storable). The plant performances in different periods are extrapolated from the two extreme R (hydrogen:electricity ratio) performances. The eventual period of peak electricity or hydrogen demand are satisfied installing plants type 2 and 3.

Scenario 5 (Flexible Coproduction plants, with hydrogen storage)

1- The program takes the results from step A
2- It guesses a number of plants type 5
3- Hypothetically the plants are set to run for the entire month at the specific ratio R of the demand in that month. R is the ratio of the hydrogen production and electricity production over a given period. The plants produce a quantity of hydrogen and electricity following the performances of the type 5 flexible plants.
4- Since the number of plants has been guessed, there will be a shortage or an excess of hydrogen or electricity. If extra electric energy is produced, it will tune the flexible plants to produce more hydrogen instead of electricity. In this way the system will never have excess electricity (not storable). But the system may have hydrogen in excess or shortage. The excess is sent to storage.
5- At this point a new R value is computed, including the fact that the plant is not producing excess electric energy, and hydrogen is sent or taken from storage. The plants will run at this new value of R. Since the number of plants has been guessed, the extra production of hydrogen and its shortage will be different. Thus we take another guess of the number of plants to exactly match the two numbers. Iteratively we get to the exact quantity. In other words the system is producing enough hydrogen to supply not the monthly demand, but the demand of the month plus considering eventual shortages or excess later in the year. The H_2 stored is used later on in the year and nothing stays in the store at the end of the cycle.

5.0 RESULTS

The program has been run with two different kinds of data: one considering the monthly energy consumption data while the second with the intra-day energy consumption data. The monthly analysis concerned The Netherlands and USA data, while the intra-day analysis only The Netherlands.

5.1 Monthly analysis

The Netherlands energy consumption data for 2004-2005 has been used (Section J, Attachments A). The input data are shown in table I.5.1 while the output is shown in table I.5.2.

Date: July 2007
Rev: Rev. 1
Made by: FWI

MONTHLY ENERGY CONSUMPTION - TABLE I.5.1

NL	EE consumption	EE Without Nucl/Ren	EE Without Nucl/Ren	NG consumption	NG Without Power Gen & Ind.	Actualized NG	Gasoline	Actualized Gasoline	Gasoline FC	Diesel	Actualized Diesel	Diesel FC	H2	H2	H2/EE
	TJ	*TJ*	*GWh*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*TJ*	*GWh*	
jan-04	28,237	26,232	7,287	212,472	89,238	53,543	13,833	13,833	4,940	20,095	20,095	11,483	69,966	19,435	2.667
feb-04	27,700	25,733	7,148	199,850	83,937	50,362	14,353	14,353	5,126	20,644	20,644	11,797	67,285	18,690	2.615
mar-04	28,920	26,867	7,463	177,005	74,342	44,605	15,994	15,994	5,712	24,042	24,042	13,738	64,055	17,793	2.384
apr-04	28,377	26,362	7,323	122,671	51,522	30,913	16,877	16,877	6,027	22,398	22,398	12,799	49,739	13,816	1.887
may-04	27,404	25,458	7,072	114,782	48,208	28,925	14,511	14,511	5,182	21,171	21,171	12,098	46,205	12,835	1.815
june-04	28,474	26,453	7,348	97,690	41,030	24,618	15,773	15,773	5,633	22,919	22,919	13,097	43,347	12,041	1.639
july-04	27,810	25,835	7,176	87,171	36,612	21,967	14,353	14,353	5,126	21,350	21,350	12,200	39,294	10,915	1.521
ago-04	28,060	26,068	7,241	85,725	36,004	21,603	14,038	14,038	5,013	20,095	20,095	11,483	38,099	10,583	1.462
sept-04	29,070	27,006	7,502	102,817	43,183	25,910	15,457	15,457	5,520	23,093	23,093	13,196	44,626	12,396	1.652
oct-04	29,188	27,116	7,532	131,973	55,429	33,257	14,748	14,748	5,267	23,145	23,145	13,225	51,750	14,375	1.908
nov-04	31,099	28,891	8,025	163,956	68,861	41,317	15,931	15,931	5,689	23,266	23,266	13,295	60,301	16,750	2.087
dec-04	31,796	29,538	8,205	200,244	84,102	50,461	15,899	15,899	5,678	23,324	23,324	13,328	69,468	19,297	2.352
jan-05	30,676	28,498	7,916	197,220	82,832	49,699	14,117	14,117	5,042	20,274	20,274	11,585	66,326	18,424	2.327
feb-05	31,290	29,068	8,074	202,085	84,876	50,925	14,890	14,890	5,318	20,581	20,581	11,760	68,003	18,890	2.339
mar-05	29,767	27,653	7,681	174,375	73,238	43,943	15,552	15,552	5,554	23,862	23,862	13,636	63,132	17,537	2.283
apr-05	28,148	26,149	7,264	121,652	51,094	30,656	15,410	15,410	5,504	22,919	22,919	13,097	49,256	13,682	1.884
may-05	28,086	26,092	7,248	111,429	46,800	28,080	15,047	15,047	5,374	22,786	22,786	13,020	46,475	12,910	1.781
june-05	28,444	26,424	7,340	96,046	40,339	24,204	15,284	15,284	5,459	23,961	23,961	13,692	43,354	12,043	1.641
july-05	27,528	25,573	7,104	90,031	37,813	22,688	13,801	13,801	4,929	20,992	20,992	11,995	39,612	11,003	1.549
ago-05	27,360	25,417	7,060	90,064	37,827	22,696	14,511	14,511	5,182	20,812	20,812	11,893	39,771	11,048	1.565
sept-05	28,732	26,692	7,414	97,492	40,947	24,568	15,221	15,221	5,436	23,787	23,787	13,593	43,597	12,110	1.633
oct-05	29,106	27,040	7,511	113,993	47,877	28,726	14,416	14,416	5,149	22,786	22,786	13,020	46,895	13,027	1.734
nov-05	31,333	29,108	8,086	160,471	67,398	40,439	15,631	15,631	5,582	24,134	24,134	13,791	59,812	16,615	2.055
dec-05	30,353	28,198	7,833	192,191	80,720	48,432	15,615	15,615	5,577	23,503	23,503	13,431	67,439	18,733	2.392

Nuclear and Renewable Energy % of Total Electric Power Production	7.1%
Power Generation and Industrial Natural Gas % of Total consumed Gas	58.0%
Natural Gas actualization factor	0.60
Gasoline actualization factor	1.00
Diesel actualization factor	1.00
Gasoline Motor Efficiency	25.0%
Diesel Motor Efficiency	40.0%
Fuel Cell Efficiency	70.0%

LEGEND: Input data from user (yellow)

	Underground	Compressed gas
	euro/kg	euro/kg
capital cost	1.5	1500
O&M costs	0.05	0.78

FOSTER WHEELER

OVERALL ECONOMICS AND ADVANTAGES OF COPRODUCTION - THE NETHERLANDS - TABLE I.5.2

Date: July 2007
Rev: Rev. 1
Made by: FWI

	SCENARIO 1	SCENARIO 2	SCENARIO 3	SCENARIO 4	SCENARIO 5
	EE PLANT AND H_2 PLANT ONLY	NON FLEX COPROD PLANT W/O H_2 STORAGE	NON FLEX COPROD PLANT WITH H_2 STORAGE	FLEXIBLE COPROD PLANT W/O STORAGE	FLEXIBLE COPROD PLANT WITH STORAGE - monthly
Quantity Plants #1	0	0	0	0	0
Quantity Plants #2	21	7	4	7	0
Quantity Plants #3	29	13	5	9	0
Quantity Plants #4	0	29	36	0	0
Quantity Plants #5	0	0	0	33	41
Total quantity of plant	50	49	45	49	41
Monthly average installed plants #1 load factor					
Monthly average installed plants #2 load factor	89.1%	66.5%	35.1%	45.9%	
Monthly average installed plants #3 load factor	75.0%	47.1%	32.5%	45.6%	
Monthly average installed plants #4 load factor		100.0%	100.0%		
Monthly average installed plants #5 load factor				100.0%	99.1%
Max quantity hydrogen in storage (million Nm3)	n/a	n/a	2,389	n/a	6,822
Max quantity hydrogen in storage per plant with storage (million Nm3)	n/a	n/a	66	n/a	166
Overall coal consumption (t/h)	9392	9234	9060	9358	9432
CO_2 capture (kg/h)	18,855,935	18,537,750	18,189,524	18,787,583	18,935,567
CO_2 emission (kg/h)	3,304,102	3,248,347	3,187,328	3,292,125	3,318,056
Plants Capital Cost (excluding storage) (milions EUR)	67,448	65,238	60,348	66,240	55,356
Underground Storage Capital Cost (including extra PSA unit) (milions EUR)	n/a	n/a	390	n/a	962
Total Capital Cost (underground)(milions EUR)	67,448	65,238	60,738	66,240	56,318
Total O&M Cost million EUR/y (underground) (base on monthly average)	5,176	5,050	4,843	5,127	4,786
Electricity Prod Cost [Euro/kWh]	0.103	0.098	0.090	0.100	0.080
NOx EMISSION (kg/h) (including availability, month average	7,447	2,481	1,274	7,591	7,741
SOx EMISSION (kg/h) (including availability, month average)	172	169	166	171	173
CO EMISSION (kg/h) (including availability, month average)	3,138	3,243	3,265	3,216	3,283
PART EMISSION (kg/h) (including availability, month average)	516	563	528	482	483

The same procedure has been followed for 2004-2005 USA energy consumption data (Section J, Attachments A). The input data are shown in table I.5.3 while the output is shown in table I.5.4.

FOSTER WHEELER

MONTHLY ENERGY CONSUMPTION - TABLE I.5.3

Date: July 2007
Rev: Rev. 1
Made by: FWI

USA	EE consumption	EE Without Nucl/Ren	EE Without Nucl/Ren	NG consumption	NG Without Power Gen & Ind.	Actualized NG	Gasoline	Actualized Gasoline	Gasoline FC	Diesel	Actualized Diesel	Diesel FC	H2	H2	H2/EE
	TJ	TJ	GWh	TJ	TJ	TJ	TJ	TJ	TJ	TJ	TJ	TJ	TJ	GWh	
jan-04	1,247,566	1,089,125	302,535	2,604,930	1,015,923	609,554	1,373,022	1,373,022	490,365	484,422	484,422	276,813	1,376,731	382,425	1.264
feb-04	1,131,408	987,719	274,366	2,420,824	944,121	566,473	1,303,993	1,303,993	465,712	484,422	484,422	276,813	1,308,997	363,610	1.325
mar-04	1,111,723	970,534	269,593	2,124,855	828,693	497,216	1,423,343	1,423,343	508,337	522,620	522,620	298,640	1,304,193	362,276	1.344
apr-04	1,046,016	913,172	253,659	1,672,841	652,408	391,445	1,392,977	1,392,977	497,492	531,302	531,302	303,601	1,192,538	331,260	1.306
may-04	1,178,568	1,028,890	285,803	1,463,684	570,837	342,502	1,447,851	1,447,851	517,090	520,884	520,884	297,648	1,157,240	321,456	1.125
june-04	1,242,306	1,084,533	301,259	1,307,188	509,803	305,882	1,422,880	1,422,880	508,171	550,401	550,401	314,515	1,128,568	313,491	1.041
july-04	1,358,395	1,185,879	329,411	1,500,118	585,046	351,028	1,475,932	1,475,932	527,119	526,093	526,093	300,624	1,178,771	327,436	0.994
ago-04	1,326,380	1,157,930	321,647	1,515,791	591,158	354,695	1,471,068	1,471,068	525,381	531,302	531,302	303,601	1,183,677	328,799	1.022
sept-04	1,208,239	1,054,793	292,998	1,397,736	545,117	327,070	1,376,121	1,376,121	491,472	555,610	555,610	317,491	1,136,033	315,565	1.077
oct-04	1,124,820	981,968	272,769	1,477,558	576,248	345,749	1,434,826	1,434,826	512,438	559,082	559,082	319,476	1,177,662	327,128	1.199
nov-04	1,087,564	949,443	263,734	1,714,534	668,668	401,201	1,382,176	1,382,176	493,634	524,357	524,357	299,632	1,194,467	331,797	1.258
dec-04	1,231,013	1,074,674	298,521	2,240,004	873,602	524,161	1,451,973	1,451,973	518,562	515,675	515,675	294,672	1,337,394	371,498	1.244
jan-05	1,235,624	1,078,700	299,639	2,526,700	985,413	591,248	1,389,991	1,389,991	496,425	501,785	501,785	286,734	1,374,407	381,780	1.274
feb-05	1,072,584	936,366	260,102	2,218,690	865,289	519,173	1,262,435	1,262,435	450,869	494,840	494,840	282,766	1,252,809	348,002	1.338
mar-05	1,140,408	995,576	276,549	2,175,786	848,557	509,134	1,418,539	1,418,539	506,621	539,983	539,983	308,562	1,324,317	367,866	1.330
apr-05	1,038,838	906,906	251,918	1,716,022	669,249	401,549	1,393,232	1,393,232	497,583	543,456	543,456	310,546	1,209,678	336,022	1.334
may-05	1,129,583	986,126	273,924	1,592,050	620,900	372,540	1,463,451	1,463,451	522,661	534,774	534,774	305,585	1,200,786	333,552	1.218
june-05	1,301,299	1,136,034	315,565	1,483,648	578,623	347,174	1,430,634	1,430,634	510,941	567,764	567,764	324,436	1,182,551	328,486	1.041
july-05	1,437,307	1,254,769	348,547	1,555,305	606,569	363,941	1,503,748	1,503,748	537,053	529,565	529,565	302,609	1,203,603	334,334	0.959
ago-05	1,447,121	1,263,337	350,927	1,587,067	618,956	371,374	1,504,272	1,504,272	537,240	541,719	541,719	309,554	1,218,168	338,380	0.964
sept-05	1,255,723	1,096,246	304,513	1,399,674	545,873	327,524	1,360,806	1,360,806	486,002	555,610	555,610	317,491	1,131,017	314,171	1.032
oct-05	1,134,122	990,089	275,025	1,450,397	565,655	339,393	1,425,256	1,425,256	509,020	538,247	538,247	307,570	1,155,982	321,106	1.168
nov-05	1,097,636	958,236	266,177	1,746,539	681,150	408,690	1,391,324	1,391,324	496,901	543,456	543,456	310,546	1,216,138	337,816	1.269
dec-05	1,246,514	1,088,207	302,280	2,274,362	887,001	532,201	1,466,163	1,466,163	523,630	522,620	522,620	298,640	1,354,471	376,242	1.245

Nuclear and Renewable Energy % of Total Electric Power Production	12.7%
Power Generation and Industrial Natural Gas % of Total consumed Gas	61.0%
Natural Gas actualization factor	0.60
Gasoline actualization factor	1.00
Diesel actualization factor	1.00
Gasoline Motor Efficiency	25.0%
Diesel Motor Efficiency	40.0%
Fuel Cell Efficiency	70.0%

LEGEND: Input data from user

	Underground euro/kg	Compressed gas euro/kg
capital cost	1.5	1500
O&M costs	0.05	0.78

FOSTER WHEELER

OVERALL ECONOMICS AND ADVANTAGES OF COPRODUCTION - USA- TABLE I.5.4

Date: July 2007
Rev: Rev. 1
Made by: FWI

	SCENARIO 1	SCENARIO 2	SCENARIO 3	SCENARIO 4	SCENARIO 5
	EE PLANT AND H_2 PLANT ONLY	NON FLEX COPROD PLANT W/O H_2 STORAGE	NON FLEX COPROD PLANT WITH H_2 STORAGE	FLEXIBLE COPROD PLANT W/O STORAGE	FLEXIBLE COPROD PLANT WITH STORAGE monthly
Quantity Plants #1	0	0	0	0	0
Quantity Plants #2	875	461	425	170	0
Quantity Plants #3	563	101	0	97	0
Quantity Plants #4	0	856	932	0	0
Quantity Plants #5	0	0	0	1132	1253
Total quantity of plant	1438	1418	1357	1399	1253
Monthly average installed plants #1 load factor					
Monthly average installed plants #2 load factor	83.0%	67.7%	65.3%	40.0%	
Monthly average installed plants #3 load factor	89.2%	40.3%		46.7%	
Monthly average installed plants #4 load factor		100.0%	100.0%		
Monthly average installed plants #5 load factor				100.0%	99.99%
Max quantity hydrogen in storage (million Nm3)	n/a	n/a	37,830	n/a	85,016
Max quantity hydrogen in storage per plant with storage (million Nm3)	n/a	n/a	41	n/a	68
Overall coal consumption (t/h)	285091	280559	280730	289074	290832
CO_2 capture (kg/h)	572,350,688	563,251,509	563,593,831	580,345,618	583,874,557
CO_2 emission (kg/h)	100,292,306	98,697,868	98,757,853	101,693,248	102,311,620
Plants Capital Cost (excluding storage) (milions EUR)	2,038,481	1,984,369	1,909,005	1,909,596	1,691,725
Underground Storage Capital Cost (including extra PSA unit) (milions EUR)	n/a	n/a	5,717	n/a	13,718
Total Capital Cost (underground)(milions EUR)	2,038,481	1,984,369	1,914,721	1,909,596	1,705,444
Total O&M Cost million EUR/y (underground) (base on monthly average)	157,251	153,974	151,641	153,743	147,089
Electricity Prod Cost [Euro/kWh]	0.091	0.088	0.085	0.085	0.075
NOx EMISSION (kg/h) (including availability, month average	264,707	118,335	106,043	265,393	266,275
SOx EMISSION (kg/h) (including availability, month average)	5,220	5,137	5,140	5,292	5,325
CO EMISSION (kg/h) (including availability, month average)	111,307	114,506	115,085	112,576	113,052
PART EMISSION (kg/h) (including availability, month average)	18,175	18,837	18,592	17,573	17,561

5.2 Intra-day analysis

As outlined in Section J, only 2004-2005 electric energy consumption data for The Netherlands has been collected and analyzed. For scenarios 1, 2 and 4, since they do not include any storage, there will be no difference between the monthly analysis and the intra-day analysis. In fact the impact of the hourly analysis is that the hydrogen stored underground at a certain time could be used when necessary, even in the next hour. Thus the plant can work on different performance to exactly fit the consumption. Since from paragraph 5.1 it has been found that the scenario 5 is better (less cost of energy) than scenario 3, the intra-day analysis in carried on only for it.

For this analysis, the program is run in the same way as for the monthly analysis, including the following two considerations regarding the input data:

- The electric energy consumption is based on hourly average;
- The fuel consumptions (natural gas, gasoline and diesel fuel) are based on monthly averages. As a consequence the input of fuels will be the same as in the monthly analysis (table I.5.1).

The output for scenario 5 in The Netherlands both for year 2004 and for 2005 is shown in table I.5.5.

Intra-day analysis for the USA has not been performed since it was out of the scope and since the software lets the user reproduce any energy consumption scenarios by only changing the input values, so the information can be produced by others if required.

OVERALL ECONOMICS AND ADVANTAGES OF COPRODUCTION - THE NETHERLANDS -TABLE I.5.5	
Date:	July 2007
Rev:	Rev. 1
Made by:	FWI

	SCENARIO 6-2004	SCENARIO 6-2005
	FLEXIBLE COPROD PLANT WITH STORAGE - day	FLEXIBLE COPROD PLANT WITH STORAGE - day
Quantity Plants #1	0	0
Quantity Plants #2	0	0
Quantity Plants #3	0	0
Quantity Plants #4	0	0
Quantity Plants #5	40	40
Total quantity of plant	40	40
Monthly average installed plants #1 load factor		
Monthly average installed plants #2 load factor		
Monthly average installed plants #3 load factor		
Monthly average installed plants #4 load factor		
Monthly average installed plants #5 load factor	98.4%	98.3%
Max quantity hydrogen in storage (million Nm3)	6,449	6,945
Max quantity hydrogen in storage per plant with storage (million Nm3)	161	174
Overall coal consumption (t/h)	9133	9132
CO_2 capture (kg/h)	18,335,422	18,332,920
CO_2 emission (kg/h)	3,212,893	3,212,455
Plants Capital Cost (excluding storage) (milions EUR)	54,006	54,006
Underground Storage Capital Cost (including extra PSA unit) (milions EUR)	987	1,057
Total Capital Cost (underground)(milions EUR)	54,992	55,063
Total O&M Cost million EUR/y (underground) (base on monthly average)	4,605	4,605
Electricity Prod Cost [Euro/kWh]	0.075	0.075
NOx EMISSION (kg/h) (including availability month average	7,197	7,293
SOx EMISSION (kg/h) (including availability month average)	167	167
CO EMISSION (kg/h) (including availability month average)	3,051	3,092
PART EMISSION (kg/h) (including availability month average)	438	448

6.0 CONCLUSION

For each scenario several overall outputs are provided in order to be able to evaluate the performance and benefits of each scenario.

With reference to The Netherlands the following main conclusions can be drawn:

- If hydrogen storage is not considered, scenario 2 is slightly better than scenario 4 due to a lower capital cost. The slight advantage of the non-flexible scenario 2 compared to scenario 4 in The Netherlands region is due to the fact that the co-production plant with fixed (type 4) H_2/EE ratio is designed to produce hydrogen and electricity with a ratio that is close to that of The Netherlands' consumption. In fact, that advantage does not occur in the USA where the ratio is different.
- By considering hydrogen storage, instead, the flexible co-production scenario is better than the non-flexible scenario: electricity costs of 0.080 Euro/kWh vs. 0.90 Euro/kWh respectively. Flexible plants can vary the ratio of electricity and hydrogen produced in order to simultaneously match the requirement of the market and fill the storage cavern when the hydrogen requirement is lower than the production. When the hydrogen requirement is higher than the production it is possible to use the hydrogen available in storage.
- Same considerations can be made for the non flexible plants: when the hydrogen production is higher than the requirement, the storage cavern is filled and vice versa. Nevertheless, in the non flexible case it is not possible to vary the electricity to H_2 ratio, resulting in the necessity to introduce only H_2 plants and only electricity plants to provide for the peak demands of Hydrogen and Electricity.
- The final result is that Scenario 5 with respect to Scenario 3 has a more optimized system with a much lower number of plants (41 vs. 45), higher hydrogen storage volume (with a negligible effect on capital costs) and lower total O&M costs.
- Clearly the worst combination of plants is Scenario 1, both from a point of view of number of plants and electricity production cost.
- For Scenario 5, which is the lowest cost scenario, a simulation based on intra-day consumptions has been made. The result is a significantly lower cost of energy with respect to the cost of energy based on the monthly averages: 0.075 euro/kWh vs. 0.080 euro/kWh respectively. In the intra-day case it is possible to use the storage not only seasonally, but also daily (and hourly if necessary). This results in a lower number of plants (40 vs. 41) leading to a lower cost of energy. The storage size is almost the same

as the storage dimension depends on the seasonal need of hydrogen; the daily peak of hydrogen request can be managed with a reduced number of plants.

With reference to the USA, similar considerations can be made. The resulting best case is still the flexible co-production case with H_2 storage.
Due to very high demand both of hydrogen and electricity, the number of plants is much higher than in the Netherlands case (1253 plants vs. 41 plants in scenario 5), leading to a better use of the plants with a higher load factor for Scenario 5 plants (almost 100% vs. 99.1%).
This advantage in the plants utilization factor and the lower volume of storage required per plant, explains the lower cost of energy in the USA case with respect to the Netherlands case.
As the co-production plant with fixed H_2/EE ratio is designed to produce hydrogen and electricity with a ratio close to The Netherlands consumption (approx 1.9 vs. approx 1.2 for USA) in case of no hydrogen storage, Scenario 4 (flexible plants) is better than Scenario 2 (fixed ratio plants).

7.0 SENSITIVITY STUDY OF UNDERGROUND STORAGE COST

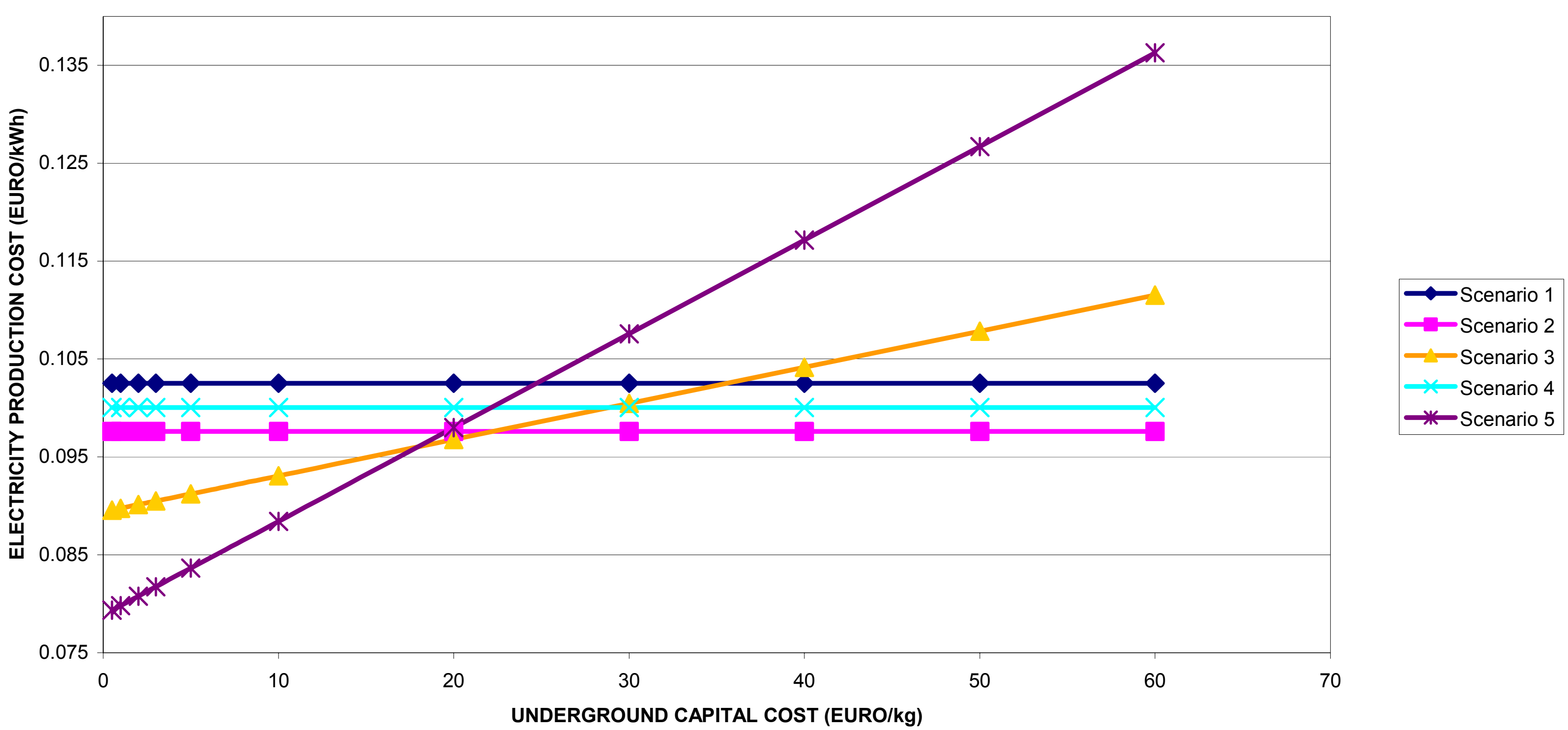
Because the cost of underground storage, varies widely, depending on the geological configuration of the area, it could strongly affect the outputs. Thus a sensitivity study has been also performed for The Netherlands.

Graph I.7.1 shows the dependence of the Electricity production cost on the hydrogen storage cost, based on monthly consumptions.

For hydrogen storage costs lower than approximately 20 Euro/kg, Scenario 5 remains the winning choice. For increasing storage costs the impact on overall investment costs become higher and both alternatives with hydrogen storage appear uncompetitive. In any case the cost considered for underground storage, is likely to be significantly lower than 20 Euro/kg.

On the contrary, above ground storage is not justified, having a cost at least one order of magnitude higher than 20 Euro/kg.

Underground storage cost sensitivity



CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAM
PROJECT NAME	:	HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME	:	ANALYSIS OF HYDROGEN AND ELECTRICITY DEMAND IN USA AND THE NETHERLANDS

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Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L. Valota	P. Cotone	S. Arienti
July 2007	Rev 1	L. Valota	P. Cotone	S. Arienti

ANALYSIS OF HYDROGEN AND ELECTRICITY DEMAND IN USA AND THE NETHERLANDS

INDEX

1.0 INTRODUCTION

Hydrogen is currently used on a large scale in ammonia plants and modern petroleum refineries. In the future it may also be used as energy carrier for vehicles, distributed heat and fuel cells power generation. Moreover hydrogen can be stored in above ground stores (drums and pipelines) or, with much lower costs, in geological underground stores.

The long term goal is to produce hydrogen from renewable energy sources but in a near term the cheapest way to produce hydrogen with low CO_2 emissions is expected to be by use of fossil fuels with CO_2 capture and storage.

Hydrogen can be produced in stand-alone plants but it may be advantageous to co-produce hydrogen and electricity, following the demand.

The aim of this study is to analyze the demand of energy in different regions and forecast the quantity of hydrogen that would be required to fulfill the demand if the conventional fossil fuel systems were replaced with hydrogen systems based on the state of the art technology.

Two sections compose the study. The first one is a collection and description of the energy consumption data such as electricity, natural gas, gasoline and diesel oil, of two different regions: The Netherlands and United States. These regions have been chosen because they represent, at a regional scale, two possible different world consumption scenarios; indeed The Netherlands presents a peak winter demand for electricity mostly due to electrical heaters while in the United States the electricity peak is during summertime for the massive use of electrical air conditioner.

The second section performs an estimate of the required quantity of hydrogen and electricity needed in such areas with the standard fossil fuel systems replaced as much as possible by hydrogen systems.

Thus the final output is the ratio of hypothetical hydrogen and electricity demand in The Netherlands and USA for 2004-2005 under a certain hypothesis of fossil fuel system conversion.

In this way, with the intention of exactly matching the demands, a combined H_2 and electricity production plant should meet the estimated demands to fully take advantage of flexible co-production.

2.0 THE NETHERLANDS ENERGY CONSUMPTION

2.1 Electricity Demand

At the present time The Netherlands is the eighth-greatest electricity producer in the EU and 23rd in the world, accounting for about 3.1% of total annual EU generation and about 0.8% of the world's annual total electricity generation. Although renewable energy is starting to make inroads into The Netherlands energy mix, more than 90% of its generation is via conventional thermal power plants. Table 2.1 is a breakdown of the total electrical installed capacity energy by source.

Table 2.1: Percentage of electricity installed capacity by energy source in The Netherlands (2006) (1)

Source	Percentage
Thermal	92.9%
Hydro	0.2%
Nuclear	2.1%
Renewable	4.8%

Overall The Netherlands generates about 25% more electricity annually than it did a decade ago, while consumption of electricity in The Netherlands has shown an even greater annual increase. This increase is mainly due to the more intensive use of electrical appliances in households.

Electricity monthly consumption in The Netherlands for 2004-2005 is shown in table 2.2 and plot in figure 2.1.

Intra-day electric energy consumption data has also been collected but due to their quantity, they are not showed in the current report but they are used in Section I.

Table 2.2: Monthly electricity consumption in The Netherlands for 2004-2005 [GWh-TJ] (1)

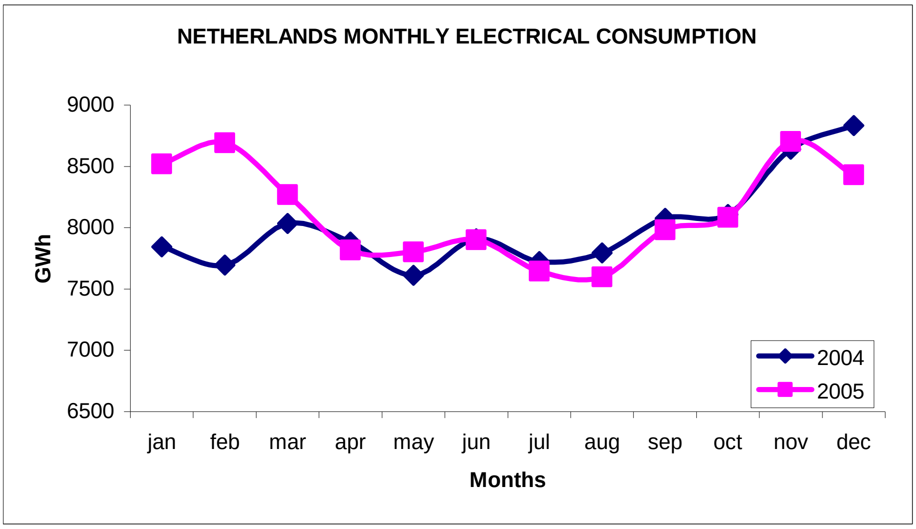
2004

Month	GWh	TJ
jan	7844	28237
feb	7694	27700
mar	8033	28920
apr	7882	28377
may	7612	27404
jun	7910	28474
jul	7725	27810
aug	7794	28060
sep	8075	29070
oct	8108	29188
nov	8639	31099
dec	8832	31796

2005

Month	GWh	TJ
jan	8521	30676
feb	8692	31290
mar	8269	29767
apr	7819	28148
may	7802	28086
jun	7901	28444
jul	7647	27528
aug	7600	27360
sep	7981	28732
oct	8085	29106
nov	8704	31333
dec	8431	30353

Figure 2.1: Monthly electricity consumption in The Netherlands for 2004-2005 [GWh]



2.2 Natural Gas Consumption

Proved reserves, as reported by The Netherlands WEC Member Committee, have been gradually declining during the last ten years, but still represent one of the largest gas resources in Western Europe. The enormous field of Groningen in the north-west of The Netherlands accounts for almost two-thirds of the country's proved reserves.
Gas production has tended to fluctuate in recent years, depending on weather conditions in Europe, thus demonstrating the flexibility that enables The Netherlands to play the role of a swing producer. Nearly 60% of 1999 output came from onshore fields, with Groningen contributing about 40%.

Table 2.3: Natural Gas reserves and production in The Netherlands (2)

Proved recoverable reserves (billion cubic meters)	1 714
Production (net billion cubic meters)	70.3
Recoverable / Production ratio (years)	24.4

Nearly half of Netherlands gas output is exported, principally to Germany but also to France, Belgium, Italy, Luxembourg and Switzerland.
The principal domestic market consists of electricity and heat generation for both industrial and residential sectors. The amount of natural gas used depends largely on the severity of the winters. Historical data summary of monthly natural gas consumption in year 2004/2005 and breakdown for energy utilization in The Netherlands in year 2006 is shown below.

Table 2.4: Monthly Natural Gas consumption in The Netherlands for 2004-2005 [million cubic meter - TJ] (1)

2004	NATURAL GAS CONSUMPTION	
Month	mcm	TJ
Jan	6,141	212,472
Feb	5,776	199,850
Mar	5,116	177,005
Apr	3,545	122,671
May	3,317	114,782
Jun	2,823	97,690
Jul	2,519	87,171
Aug	2,478	85,725
Sep	2,972	102,817
Oct	3,814	131,973
Nov	4,739	163,956
Dec	5,787	200,244

2005	NATURAL GAS CONSUMPTION	
Month	mcm	TJ
Jan	5,700	197,220
Feb	5,841	202,085
Mar	5,040	174,375
Apr	3,516	121,652
May	3,221	111,429
Jun	2,776	96,046
Jul	2,602	90,031
Aug	2,603	90,064
Sep	2,818	97,492
Oct	3,295	113,993
Nov	4,638	160,471
Dec	5,555	192,191

Figure 2.2: Natural gas consumption in Netherlands for 2004-2005 [million cubic meter]

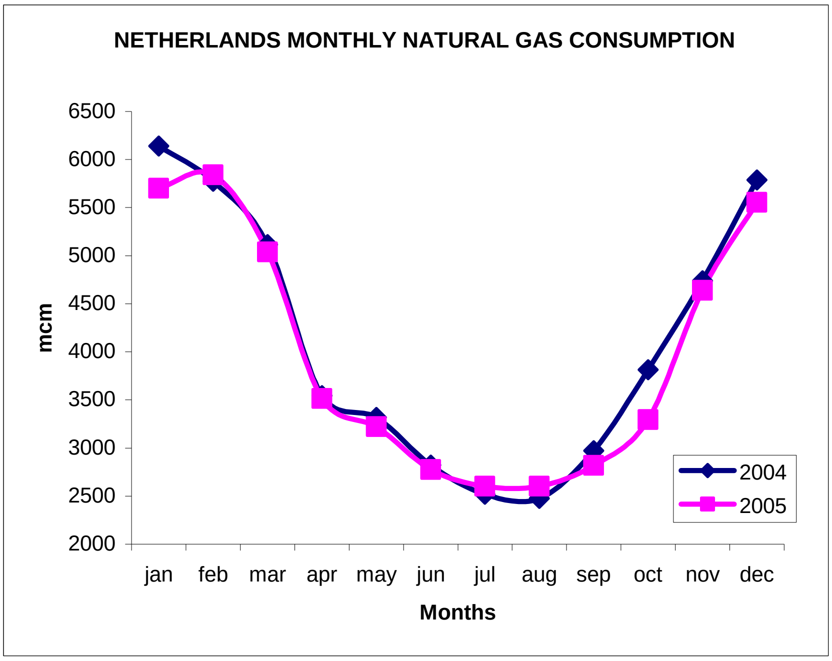


Table 2.5: Natural Gas consumption breakdown in the The Netherlands (2006) (1)

Use	Percentage
Resid. + Comm.	34%
Industrial	22%
Power	36%
Others	8%

2.3 Gasoline and Diesel Oil Demand

The Netherlands is a small country with a high density of population, especially in the Randstat zone. As a gateway of Europe, the port of Rotterdam and the related truck traffic are of major importance to the country. Consecutive transport master plans are oriented to the control of car traffic, either through alternative modes or taxation. Despite this, trucks are of major importance to the country.

Figures 2.3 and 2.4 plot data of gasoline and diesel oil consumption from IEA Oil Market Report. A close look at the graphs shows that the consumption is particularly high during spring and autumn, low during summer and winter.

Figure 2.3: Motor gasoline demand in Netherlands, 2003-06 [kbarrels/day] (3)

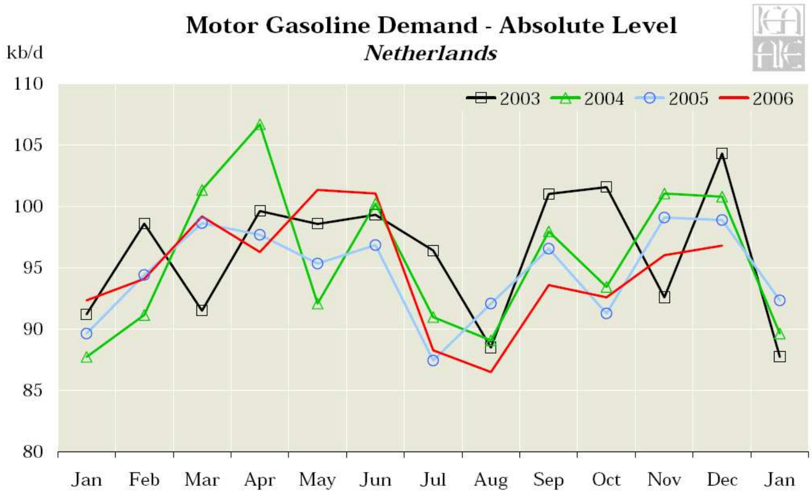
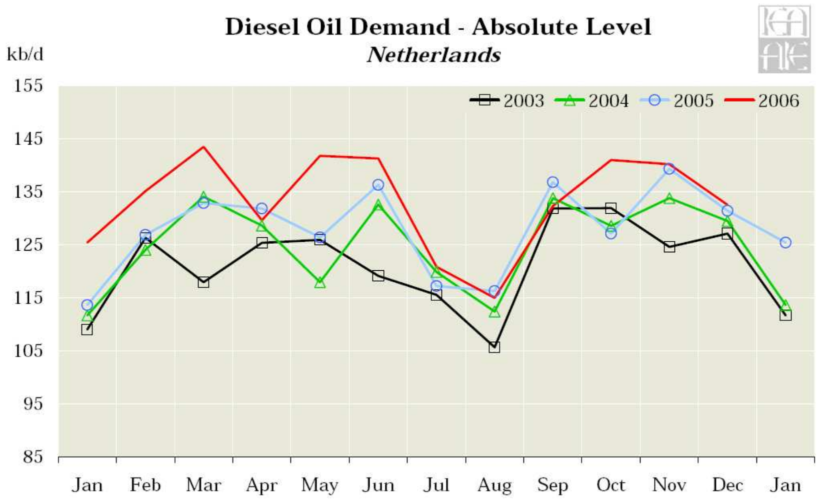


Figure 2.4: Diesel oil demand in Netherlands, 2003-06 [kbarrels/day] (3)



3.0 UNITED STATES

3.1 Electricity Demand

The United States is both the world's greatest producer and consumer of electricity, accounting for about one-fourth of both the world's annual electricity generation and consumption. By far, the majority of electricity generation in the United States is from fossil fuels, with coal by itself accounting for more than half of all generation.

Table 3.1: Electricity installed capacity percentage by source in USA (2006) (1)

Source	Percentage
Thermal	79.1%
Hydro	8.2%
Nuclear	10.6%
Renewable	2.1%

Most of the electricity consumed in the northeastern part of the United States is generated from hydroelectric sources in Canada's Québec and Ontario provinces, while the United States exports electricity to some Canadian markets. There is also electricity trade between the United States and Mexico, but inadequate cross-border power transmission infrastructure is currently a limiting factor.

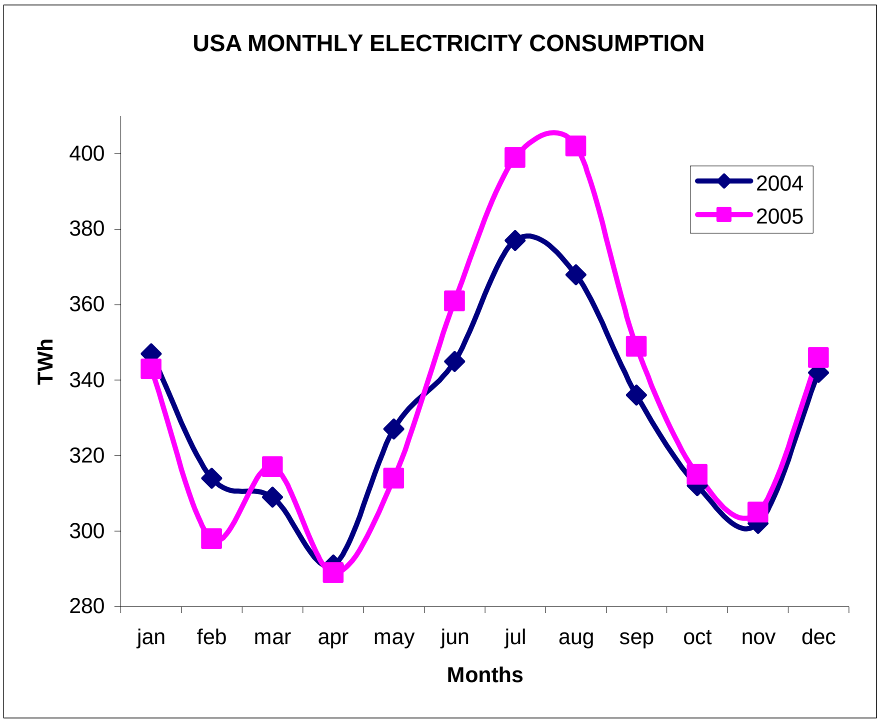
Demand for electricity in the United States has greatly increased, with electricity consumption now more than 20% higher than it was a decade ago.
Electricity demand increases during the summer period basically due to air conditioning systems. An historical summary of monthly electricity consumption in the United States for 2004-2005 is shown in Table 3.2 and plot in figure 3.1. Intra-day electric energy consumption analysis has not been performed.

Tables 3.2 : Monthly electricity consumption in USA for 2004-2005 [TWh-TJ] (1)

2004	ELECTRICITY CONSUMPTION	
Month	TWh	TJ
Jan	347	1,247,566
Feb	314	1,131,408
Mar	309	1,111,723
Apr	291	1,046,016
May	327	1,178,568
Jun	345	1,242,306
Jul	377	1,358,395
Aug	368	1,326,380
Sep	336	1,208,239
Oct	312	1,124,820
Nov	302	1,087,564
Dec	342	1,231,013

2005	ELECTRICITY CONSUMPTION	
Month	TWh	TJ
Jan	343	1,235,624
Feb	298	1,072,584
Mar	317	1,140,408
Apr	289	1,038,838
May	314	1,129,583
Jun	361	1,301,299
Jul	399	1,437,307
Aug	402	1,447,121
Sep	349	1,255,723
Oct	315	1,134,122
Nov	305	1,097,636
Dec	346	1,246,514

Figure 3.1: Monthly electricity consumption in USA, year 2004-2005 [TWh]



3.2 Natural Gas Consumption

The United States has proved gas reserves estimated at about 4740 billion cubic meters (January 2005), which represents about 3% of the current world total. The United States is currently the world's second-greatest producer of natural gas, after Russia, and accounts for about one-fifth of the world's annual natural gas production. It is also the world's greatest consumer of natural gas, accounting for nearly one-fourth of the world's total annual natural gas consumption. About one-fifth of all natural gas consumed is now imported, and more than 80% of U.S. natural gas imports are from the western provinces of Canada.

Table 3.3: Natural Gas reserves and production in USA (2)

Proved recoverable reserves (billion cubic meters)	4740
Production (net billion cubic meters)	527.3
Recoverable / Production ratio (years)	9.0

Demand for natural gas in the United States has been slowly increasing over the past decade and is now about 8% greater than it was a decade ago. More than one-third of the natural gas consumed in the United States is for industrial uses, almost another one third is for residential and commercial use, while another one-fourth is used for power production. An historical summary of monthly natural gas consumption in the United States is shown below.

Table 3.4: Monthly natural gas consumption in USA for 2004-2005 [million cubic meter -TJ] (1)

2004	NATURAL GAS CONSUMPTION	
Month	mcm	TJ
Jan	75,287	2,604,930
Feb	69,966	2,420,824
Mar	61,412	2,124,855
Apr	48,348	1,672,841
May	42,303	1,463,684
Jun	37,780	1,307,188
Jul	43,356	1,500,118
Aug	43,809	1,515,791
Sep	40,397	1,397,736
Oct	42,704	1,477,558
Nov	49,553	1,714,534
Dec	64,740	2,240,004

2005	NATURAL GAS CONSUMPTION	
Month	mcm	TJ
Jan	73,026	2,526,700
Feb	64,124	2,218,690
Mar	62,884	2,175,786
Apr	49,596	1,716,022
May	46,013	1,592,050
Jun	42,880	1,483,648
Jul	44,951	1,555,305
Aug	45,869	1,587,067
Sep	40,453	1,399,674
Oct	41,919	1,450,397
Nov	50,478	1,746,539
Dec	65,733	2,274,362

Figure 3.2: Monthly natural gas consumption in USA for 2004-2005 [million cubic meter]

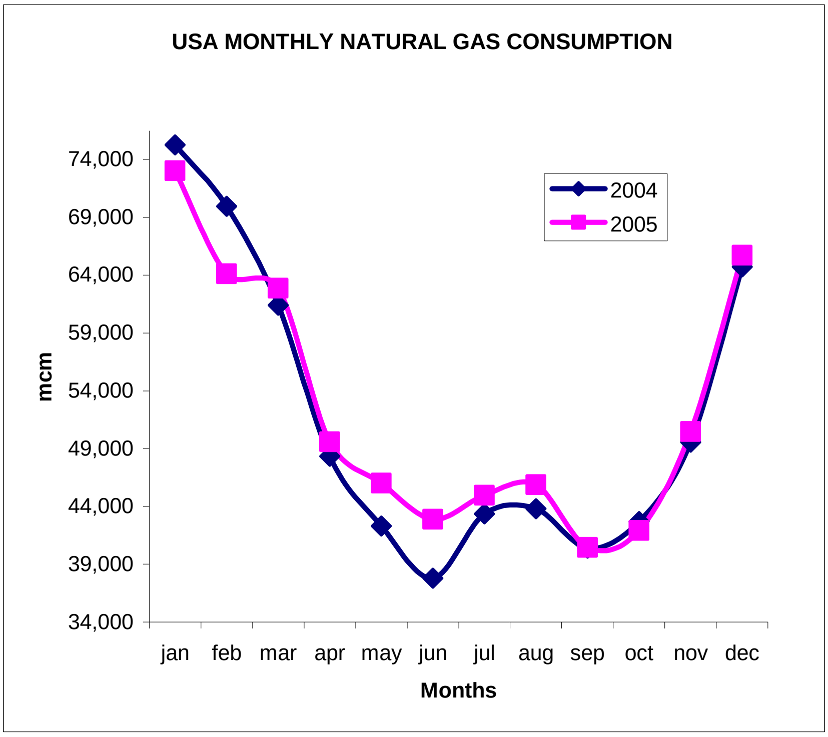


Table 3.5: Natural Gas consumption breakdown in USA (2006) (1)

Use	Percentage
Resid. + Comm.	36%
Industrial	35%
Power	26%
Others	3%

3.3 <u>Gasoline and Diesel Oil Demand</u>

The USA is both geographically and demographically a large country. The truck traffic is of major importance to the country, with a strong impact on fuel demand.

Figure 3.3 and 3.4 plot data of gasoline and diesel oil consumption from IEA Oil Market Report.

Figure 3.3: Motor gasoline demand in USA, 2003-06 [kbarrels/day] (3)

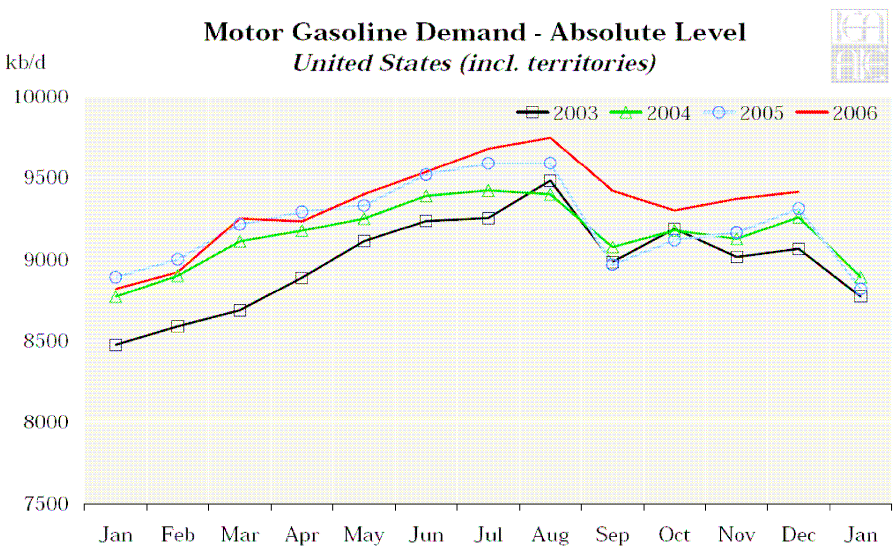
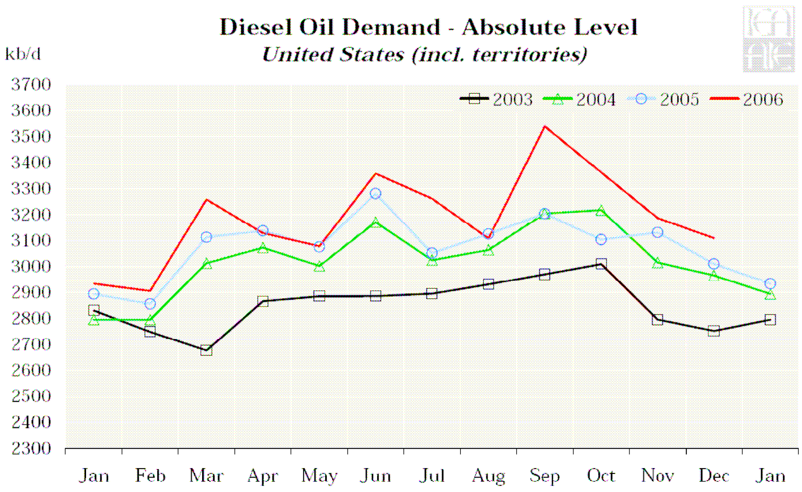


Figure 3.4: Diesel oil demand in USA, 2003-06 [kbarrels/day] (3)



4.0 H_2 AND ELECTRICITY CONSUMPTION ESTIMATION

The aim of this section is to analyze the demand of energy in different regions and forecast the quantity of energy that would be required to fulfill the demand if the conventional fossil fuel systems were as much as possible replaced with hydrogen systems based on the state of the art technology.

Given the consumption data provided in the previous paragraphs, equivalent quantities of electricity and hydrogen consumption have been calculated under certain criteria.

Electrical energy consumption

To convert the actual electric energy consumption to the hypothetical consumption (modified consumption) the following criterion has been applied:

- The production of electricity coming from renewable energy sources and nuclear is not converted in EE modified consumption. This because in a hypothetical hydrogen energy scenario, nuclear and renewable may still have their power production.

In formulas, the modified electrical consumption energy $EE^{\mathrm{mod}\_consumption}$ is:

$$EE^{\mathrm{mod}\_consumption} = EE \times (1 - RN\%)$$

where EE is the actual electrical energy consumption and $RN\%$ is the fraction of power generated by renewable, hydro and nuclear sources, equal to 0.071 for The Netherlands and 0.209 for USA (tables 2.1 and 3.1)

Hydrogen consumption

To convert the energy from fossil fuel to hydrogen energy consumption, the following hypotheses have been made:

1- The natural gas consumed by industry and power generation plants is not converted in to H_2 consumption. That is, gas will continue to be consumed by power plants.

2- 60% of the remaining part of the natural gas consumption is converted to hydrogen. The remaining 40% is kept as gas consumption.
3- The diesel and gasoline consumption is converted into hydrogen consumption considering the state-of-the-art fuel cell efficiency.

Three actualization factors (α_{NG}, α_G and α_D) are introduced in order to quantify the conversion factor of fossil fuels to hydrogen.

In formulas, the equivalent energy consumption of hydrogen $H_2^{equivalent}$ is:

$$H_2^{equivalent} = NG \times (1 - PI\%) \times \alpha_{NG} + G \times \frac{\eta_{GA}}{\eta_{FC}} \times \alpha_G + D \times \frac{\eta_{DIES}}{\eta_{FC}} \times \alpha_D$$

where:

NG is the actual natural gas consumption
$PI\%$ is the power and industry consumption percentage with respect to the overall natural gas consumption, equal to 0.58 for The Netherlands and 0.61 for USA (tables 2.5 and 3.5)
G is the actual motor gasoline consumption
η_{GA} is the efficiency of a standard car gasoline engine, equal to 0.25
η_{FC} is the efficiency of a standard state-of-the-art fuel cell, equal to 0.70
D is the actual diesel oil consumption
η_{DIES} is the efficiency of a standard car diesel oil engine, equal to 0.40
α_{NG} is the actualisation factor for natural gas that gives information on the quantity of natural gas consumption that can be converted into hydrogen consumption, for the purpose of this study set to 0.6. In other words, only 60% of the natural gas not used for industrial and power usage is converted into hydrogen consumption. This because the hypothetical system, for technological and realistic forecasts, cannot consist entirely of hydrogen based systems but may keep a fraction of natural gas consumption.
α_G is the actualisation factor for gasoline that gives information on the quantity of gasoline consumption that can be converted into hydrogen consumption, for the purpose of this study set to 1
α_D is the actualisation factor for diesel oil that gives information on the quantity of diesel consumption that can be converted into hydrogen consumption, for the purpose of this study set to 1

Even if both α_G and α_D are set to 1, it has been preferred to separately show the coefficients since the methodology outlined in this study could be applied also to different consumption scenarios of any different region.

Evaluation of R

The value R is the ratio between the equivalent consumption of hydrogen and the modified electrical consumption. It is a value significant to summarize the trend of the hydrogen and electricity consumptions in a co-production vision.

In formulas, the value R is given by equation:

$$R = \frac{H_2^{equivalent}}{EE^{\mathrm{mod\_}consumption}}$$

Under these assumptions, $H_2^{equivalent}$, $EE^{\mathrm{mod\_}consumption}$ and R The Netherlands and USA for years 2004-2005 are shown in the next tables.

Table 4.1: Hydrogen equivalent consumption and modified electrical consumption for The Netherlands 2004-2005 (TJ)

2004

Month	$H_2^{equivalent}$	$EE^{mod\_consumption}$
jan	69,966	26,232
feb	67,285	25,733
mar	64,055	26,867
apr	49,739	26,362
may	46,205	25,458
jun	43,347	26,453
jul	39,294	25,835
aug	38,099	26,068
sep	44,626	27,006
oct	51,750	27,116
nov	60,301	28,891
dec	69,468	29,538

2005

Month	$H_2^{equivalent}$	$EE^{mod\_consumpti}$
jan	66,326	28,498
feb	68,003	29,068
mar	63,132	27,653
apr	49,256	26,149
may	46,475	26,092
jun	43,354	26,424
jul	39,612	25,573
aug	39,771	25,417
sep	43,597	26,692
oct	46,895	27,040
nov	59,812	29,108
dec	67,439	28,198

Table 4.2: Hydrogen equivalent consumption and modified electrical consumption for USA 2004-2005 (TJ)

2004

Month	$H_2^{equivalent}$	$EE^{mod\_consumption}$
jan	1,376,731	1,089,125
feb	1,308,997	987,719
mar	1,304,193	970,534
apr	1,192,538	913,172
may	1,157,240	1,028,890
jun	1,128,568	1,084,533
jul	1,178,771	1,185,879
aug	1,183,677	1,157,930
sep	1,136,033	1,054,793
oct	1,177,662	981,968
nov	1,194,467	949,443
dec	1,337,394	1,074,674

2005

Month	$H_2^{equivalent}$	$EE^{mod\_consumpti}$
jan	1,374,407	1,078,700
feb	1,252,809	936,366
mar	1,324,317	995,576
apr	1,209,678	906,906
may	1,200,786	986,126
jun	1,182,551	1,136,034
jul	1,203,603	1,254,769
aug	1,218,168	1,263,337
sep	1,131,017	1,096,246
oct	1,155,982	990,089
nov	1,216,138	958,236
dec	1,354,471	1,088,207

Table 4.3: R values for The Netherlands and USA in 2004-2005

Netherlands

Month	2004	2005
jan	2.67	2.33
feb	2.61	2.34
mar	2.38	2.28
apr	1.89	1.88
may	1.81	1.78
jun	1.64	1.64
jul	1.52	1.55
aug	1.46	1.56
sep	1.65	1.63
oct	1.91	1.73
nov	2.09	2.05
dec	2.35	2.39

USA

Month	2004	2005
jan	1.26	1.27
feb	1.33	1.34
mar	1.34	1.33
apr	1.31	1.33
may	1.12	1.22
jun	1.04	1.04
jul	0.99	0.96
aug	1.02	0.96
sep	1.08	1.03
oct	1.20	1.17
nov	1.26	1.27
dec	1.24	1.24

Figure 4.1: R value for The Netherlands

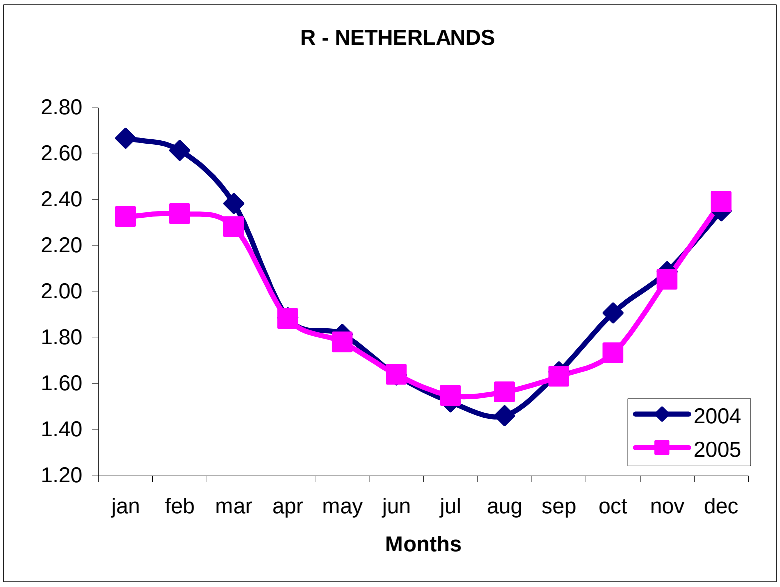


Figure 4.2: R value for USA

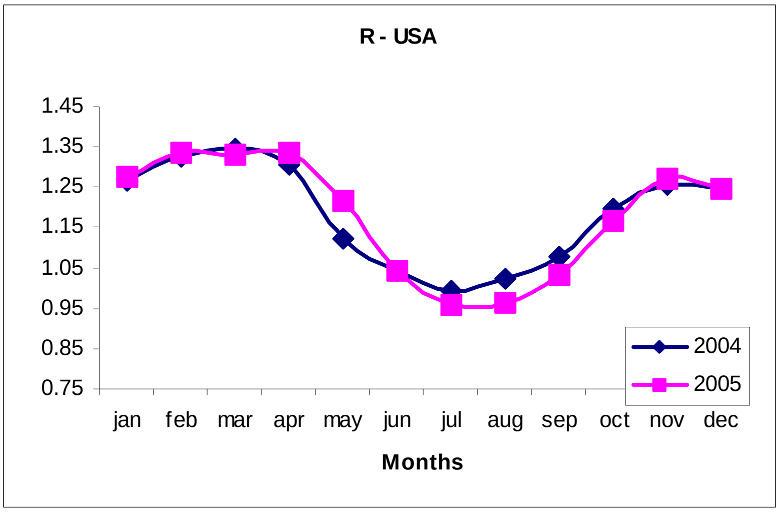


Table 4.4: R values summary table for The Netherlands and USA in the 2004-2005 period

Netherlands

	Average R	min	max
2004	2.00	1.46	2.67
2005	1.93	1.55	2.39

USA

	Average R	min	max
2004	1.18	0.99	1.34
2005	1.18	0.96	1.34

As shown in table 4.1 and 4.2, the R value trend is similar to the natural gas consumption trend. This is because, in comparison with the hydrogen demand, the electrical consumption is more constant.

R presents high values during winter for both the Netherlands and USA and low values in summer. This is because the maximum consumption of natural gas is during winter due to excessive use of heating systems. Moreover the absolute value of the maximum is different for the Netherlands and USA because, in comparison, USA uses less natural gas for heating.

Thus a flexible co-production plant able to perform production of H_2 and Electricity (i.e. perform at a given R) as shown in table 4.1 and 4.2, can, month after month, fulfill the energy demand taking maximum benefit from co-production.

5.0 CONCLUSIONS

Data on the total electric energy consumption and the fossil fuel consumption in The Netherlands and USA have been collected. Furthermore an estimation of the amount of fossil fuel that can be replaced by hydrogen has been made assuming the state-of-the-art utilization technology under a certain hypothesis (see paragraph 4.0).

The final outputs are the absolute demand values and the ratio of hydrogen and electricity demand in The Netherlands and USA for 2004-2005 under the conversion hypothesis.

6.0 REFERENCES

1. Energy Information Administration web site (http://www.eia.doe.gov).

2. World Energy Council web site (http://www.worldenergy.org/wec-geis/).

3. IEA Oil Market Report 11/10/2006

APPENDIX A: ENERGY DEMAND DATA SOURCE, MAIN TERMINOLOGY AND FUEL CONVERTION PARAMETERS

Many data are available in literature on the worldwide energy demand, such as electricity, gasoline and natural gas, but among all the following can be considered reliable sources and have been used for this report:

EIA: Energy Information Administration.

WEC: World Energy Council

Following paragraph represents a summary of the most common terminology used in the world energy reports:

Energy consumption:	The use of energy as a source of heat or power or as a raw material input to a manufacturing process.
Dry Natural Gas Production:	Marketed production less extraction loss
Natural gas:	A mixture of hydrocarbon compounds and small quantities of various nonhydrocarbons existing in the gaseous phase or solution with oil in natural underground reservoirs at reservoir conditions.
Motor gasoline:	A complex mixture of relatively volatile hydrocarbons with or without small quantities of additives, blended to form a fuel suitable for use in spark-ignition engines. Motor Gasoline includes conventional gasoline; all types of oxygenated gasoline, including gasohol; and reformulated gasoline, but excludes aviation gasoline.
Diesel fuel:	A fuel composed of distillates obtained in petroleum refining operation or blends of such distillates with residual oil used in motor vehicles. The boiling point and specific gravity are higher for diesel fuels than for gasoline.
Residential consumption:	Gas used in private dwellings, including apartments, for heating, cooking, water heating,

and other household uses.

For the estimation of the thermal power associated to fuels, the following parameters have been used:

Gasoline:
Low heating value = 32 MJ/liter
Gasoline density (average) = 0.73 metric tonnes/m3

Petro-diesel
Low heating value = 36.4 MJ/liter
Petro-diesel density (average) = 0.84 metric tonnes/m3

Natural Gas:
Low heating value = 34.6 MJ/m3

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAMME
PROJECT NAME	:	HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME	:	AGR TECHNICAL COMPARISON

ISSUED BY	:	M. GALLIO
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APPROVED BY	:	S. ARIENTI

Date	**Revised Pages**	**Issued by**	**Checked by**	**Approved by**
April 2007 July 2007	Draft Rev. 1	M. Gallio L. Valota	P. Cotone P. Cotone	S. Arienti S. Arienti

AGR TECHNICAL COMPARISON

INDEX

1.0 INTRODUCTION

IEA Greenhouse Gas R&D Programme (IEA GHG) retained Foster Wheeler to investigate alternative power and hydrogen generation plant designs, based on high rank coal gasification, aimed at assessing the potential advantage of flexible co-production of hydrogen and electricity with capture of CO_2.

The primary purpose of this study is, therefore, the evaluation of the technologies and the process alternatives that can be used in these complex power and hydrogen generation schemes to optimise efficiency and capital cost and reduce, at the same time, emissions to the atmosphere.

This report details the technologies available for capture of the acid gas (AGRU: Acid Gas Recovery Unit). The study as a whole has considered GEE, Shell and Siemens based coal gasification technologies.
The basic scheme investigated is therefore an IGCC with CO_2 capture and production of separate H_2S and CO_2 streams.

Sulphur is recovered from the acid gas by separate oxygen Claus Sulphur Removal Unit (SRU) so as to minimise sulphur emissions from the facility.

The purpose of this report is to compare the AGR schemes based on different physical solvents (Selexol and Rectisol) taking into account their relevant impacts on the downstream units (SRU and CO_2 compression unit). Suppliers of these solvents provide also the design of the AGRU, acting as licensor.
The comparison is applied to syngas coming from Shell and GEE technology as the solvent data were provided by Licensors only with reference to the above-mentioned cases; it is understood that the results could be also applied to Siemens gasification technology.

2.0 DESIGN BASIS

The following sections detail the design basis for the AGRU which has been used in licensor enquiries.

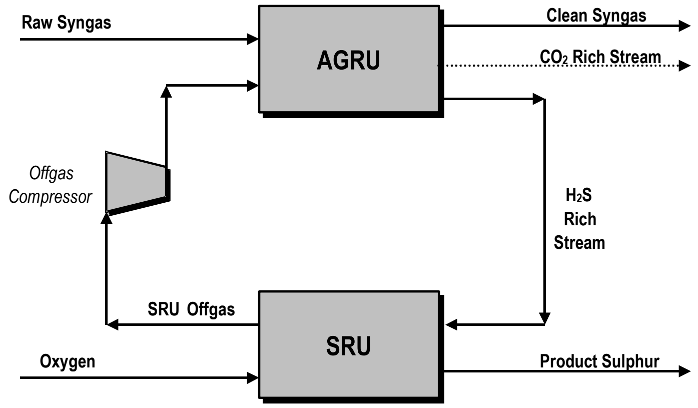
2.1 Case definition

The following cases have been investigated:

Case	Gasification	Pressure	Shift	CO_2 Capture
0A	GEE	High	Sour - Single stage	Not combined with H_2S
0B	Shell	Low	Sour - Double stage	Not combined with H_2S

2.2 Feedstock definition

The AGRU has been specified to treat also the offgas from the SRU to minimize emissions from the complex:



As a result, there are two feedstocks to the AGRU as detailed below:

2.2.1 Raw Syngas

		GEE Case 0A	Shell Case 0B
H_2	mol.%	55.04	56.41
N_2	mol.%	0.68	3.09
CO	mol.%	2.84	2.51
Ar	mol.%	0.79	0.48
CH_4	mol.%	0.02	0.00
CO_2	mol.%	40.22	37.02
H_2S	mol.%	0.22	0.18
H_2O	mol.%	0.19	0.31
COS	vppm	1	1
HCN	vppm	5	5
NH_3	vppm	10	45
Mol Wt		20.22	19.31
Flowrate	kmol/h	37276	37276
Pressure	barg	56.2	26.0
Temp	°C	38	38

2.2.2 Recycle Gas From SRU

		GEE – Case 0A		**Shell – Case 0B**	
		Selexol	**Rectisol**	**Selexol (1)**	**Rectisol (1)**
H_2	kmol/h	17.4	16.3	13.8	25.8
N_2	kmol/h	75.6	70.7	59.4	111.6
CO	kmol/h	0.2	0.1	0.2	0.2
Ar	kmol/h	0.8	0.8	0.6	1.2
CO_2	kmol/h	569.4	313.2	454.8	328.4
H_2S+COS	kmol/h	3.1	3.1	2.4	11.6
H_2O	kmol/h	1.6	0.9	1.4	2.8
Flowrate	kmol/h	668	405	533	482
Pressure	barg	As required	As required	As required	As required
Temp	°C	38	38	38	38

(1) Two parallel train are required

2.3 Product & Performance specifications

Product specifications are provided for the “clean” syngas and the recovered CO_2 and H_2S streams. In addition to these, there is also a recovery specification against CO_2 to ensure the overall target of 85% CO_2 capture for IGCC is achieved.

2.3.1 Clean Syngas

		GEE Case	Shell Case
H_2S+COS concentration	ppmv	< 40	< 40
CO_2 Washing-unit removal efficiency	%	91	91
Solvent content in syngas	ppmv	< 1	< 1

Definition of CO_2 washing unit removal efficiency is as follows:

$$\frac{CO_2 \text{ flow rate to B.L.}}{CO_2 \text{ flow rate in raw syngas to AGR}} \times 100$$

2.3.2 Acid Gas (H_2S Rich)

For this stream the Hydrogen Sulphide concentration is maximized such that the composition and operating conditions are suitable for downstream treatment in an Oxygen Claus Sulphur Recovery Unit. For purposes of design, this has been interpreted as a minimum target H_2S content of 15-20 mol%. For Rectisol cases the Hydrogen Sulphide concentration obtained is higher (approx 35%), having a positive impact on the downstream Sulphur Recovery Unit.

2.3.3 Acid Gas (CO_2 Rich)

A specification of max 100ppm H_2S in CO_2 has been adopted. Its worth noting that this specification if fairly arbitrary, and has been adopted to ensure a “sensible” separation between the two acid gases.

No hydrogen slippage specification was imposed, and the results have shown this to be a significant loss to the complex in terms of equivalent power production.

2.4 Utility conditions

The AGRU is a user of steam, electrical power, and cooling water.

For electrical power and steam, no limitations were put on designs in terms of quantities; LP steam was specified at 6.5 barg and VLP at 3.2 barg.

2.5 Turndown and availability

Turndown required is specified at 50%. The availability of an AGRU is expected to be higher than the remainder of the IGCC facility, and so no special considerations are required in the design.

2.6 Site and plot data

No limitations were specified.

2.7 Environmental standards

There are no direct emissions to the environment from an AGRU, so no environmental limits were specified. Sufficient tankage is specified for the total inventory of solvent.

2.8 Climatic data

The following data have been used in the specification of the units:

2.8.1 Air

Relative Humidity:	Average	60%
	Maximum	95%
	Minimum	40%
Temperature:	Minimum	-10°C
	Maximum	30°C
	Average	9°C

2.8.2 Cooling Water

Supply temperature:	Maximum	17°C
	Minimum	13°C
	Max increase	12°C

Design return temperature for cooling water cooler: 29 °C

Operating pressure at Users:	3.0 barg
Max allowable ΔP for Users:	1.0 bar
Design pressure:	5.0 barg
Design temperature:	60°C
Fouling Factor:	0.0002 h °C m^2/kcal

3.0 PROCESS/SOLVENT SELECTION

For removal of acid components from gas streams several methods are possible:

- Cryogenic separation
- Membrane separation
- Solvent processes:
    - Physical absorption
    - Chemical absorption

The first two processes, cryogenic and membrane separation, have not found yet commercial operation. Solvent processes have dominated the market.

The choice between physical and chemical solvent has been the subject of several studies and evaluation of many projects in the chemical industry. As a general rule chemical solvents, such as Amine, Potassium Carbonate etc., are suited when the acid gas partial pressure is low whereas physical solvents have generally a superior performance when the acid gas partial pressure is high.

Chemical solvents require more thermal energy for regeneration because the acid gas capture takes place through a foundation of a chemical bond between the acid gas and the solvent molecule. During regeneration, this chemical bond is broken with the use of thermal energy.

On the contrary, physical solvents require less thermal energy for regeneration because the acid gas is physically de-solved in the solvent and can be recovered during regeneration by a reduction of the pressure, possibly with the final thermal step only to regenerate more deeply the solvent.

It is interesting to exploit solvent selectivity properties in order to capture separately H_2S and CO_2. Chemical solvents selectivity is obtained by controlling the solvent acid gas contact time; with amine solvent a short time of contact permits to absorb preferentially H_2S instead of CO_2. With a physical solvent the selectivity is a physical characteristic of the solvent which entails a greater solubility of one acid gas versus the other.

4.0 PROCESS INFORMATION

In a previous FWI study (IEA GHG – Gasification Power generation study – 2003) it has been highlighted that, for the two cases (listed below) used as reference for the present study, physical solvent is the best choice to separately remove CO_2 and H_2S from syngas, as suggested directly by the solvents vendor:

- GEE (former Texaco) HP gasification with separate H_2S and CO_2 capture;
- Shell LP gasification with separate H_2S and CO_2 capture;

For this reason, for the present study, the analysis is focused on:

- Selexol
- Rectisol.

For both cases licensor designs have been used: UOP designed the Selexol cases and Linde the Rectisol cases.

In all the cases considered, some H_2 will be present in the stream of CO_2 sent to compression. The feasibility to separate and recover H_2 during the CO_2 compression was investigated. Due to the similar equilibrium constants of CO_2 and other components at super-critical CO_2 conditions, this separation is unfeasible, thus constituting a disadvantage of the process.

4.1 UOP (AmineGuard / Selexol)

Note that UOP now offer the Dow processes as a result of the Dow merger. A combined UOP/Dow response was received.

General Information

For above-mentioned IEA GHG – FWI study (2003), UOP provided for each case a set of information which allowed FW to fully evaluate the performance and investment costs of the AGRU and how this section meets the technical and economic targets of the entire IGCC plant. This information has been provided under a non-disclosure agreement between FW and UOP. As a consequence, this report includes only the data that UOP allows to be disclosed to IEA without a non-disclosure agreement between IEA and UOP. The workup of the data presented here is though based on a full set of data provided by UOP to FW.

Note that for both gasification cases, UOP, who now offer the DOW MDEA process, carried out an internal assessment on which process would be most applicable, so released data for the chosen solvent (Selexol) and not for DOW MDEA.

4.1.1 GEE Case

Process Description

For this case UOP believes that, due to the high syngas pressure (56 barg), and the extremely high CO_2/H_2S ratio (183/1), only an optimised Selexol Process is able to achieve an acceptable Claus Plant acid gas. With this high ratio, even a double amine configuration (AGR plus Acid Gas Enrichment (AGE)) cannot meet the minimum H_2S concentration of Acid Gas (15-20% vol). In addition, the high steam requirement of the amine process would entail a drastic reduction of the Steam Turbine power production.

Two configurations are possible, both based on a single train configuration equipped with a refrigeration package, one enhancing the acid gas H_2S concentration by using part of the Nitrogen produced by the ASU, the other one adopting a more complicated and electric power consuming process scheme.

A technical/economical evaluation performed in the previous study (IEA GHG – Gasification Power generation study – 2003) by FWI indicated that the most suitable option, taking into account the different impacts on the Investment Costs and on the Operating Costs of the two options, would be the option with Nitrogen use. This option allows reducing both the investments and operating costs. However, it was later known that high N_2 concentration in the product CO_2 stream has a negative impact for CO_2 storage, particularly if CO_2 is used for enhanced oil recovery. Therefore Option 2, without Nitrogen stripping, was finally selected.

Equipment Sizes

UOP has provided FW with a full equipment list for each case for the purpose of cost estimation, but, due to reasons of secrecy, this information cannot be released any further without the third parties signing a secrecy agreement with UOP.

Utility Consumptions

LP Steam, t/h	70.3	
Cooling Water, m^3/h	2966	(ΔT 12 °C)
Purge Water, m^3/h	0.3	
Electric Power, kW	32,100	(refrigeration Package: 32%)
Solvent Make-up, m^3/yr	120	

Material Balance

	Untreated Gas	Recycle Gas SRU	Treated Gas Exp.	CO_2 to Compr.	Acid Gas
kmol/h					
CO_2	14,992.4	569.4	1,512.2	13,695.3	354.3
H_2S+COS	82.1	3.1	0.1	1.3	83.8
H_2O	70.8	1.6	3.7	43.0	30.6
N_2	253.5	75.6	251.4	77.7	0.01
CO	1,058.6	0.2	1,035.1	23.6	0.2
H_2	20,516.7	17.4	20,277.9	254.5	1.8
Ar	294.5	0.8	288.3	6.9	0.05
Others	8.0	0	7.2	0.3	0.5
Total Flow, kmol/h	37,276.6	668.1	23,375.9	14,102.6	471.3
Pressure, bar g	57.2	28.3	56.2	(1)	1.8
Temperature, °C	38	38	35.7	(1)	48.9

Note (1): CO_2 stream is the combination of three different streams delivered at Unit B.L. at different conditions.

The proposed process reaches an H_2S+COS concentration of the treated gas exiting the unit of 4 ppm. This result is due to the integration of the CO_2 removal section with the H_2S removal section, which corresponds to a large circulation of the solvent. The CO_2 removal rate is more than 91% as required, allowing reaching an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and the CO_2 capture are achieved with a large power consumption.

The acid gas H_2S concentration is 19% dry basis, suitable to feed the oxygen blown Claus process.
Together with CO_2, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 254.5 kmol/h of Hydrogen, corresponding to 1.8% vol and to an overall thermal power of 17.2 MWt, i.e. an equivalent electric power of approx. 5.6 MWe, if fired in Gas Turbine.
- A very low quantity of H_2S, corresponding to a concentration of 92 ppmdv.

4.1.2 <u>Shell Case</u>

Process Description

For this case, the untreated gas is at low pressure (26 barg), but the CO_2/H_2S ratio is very high (206/1). UOP believes that again a selective amine has no chance of meeting the minimum H_2S concentration suitable for the SRU.
Two configurations are possible, one enhancing the acid gas H_2S concentration by using part of the nitrogen produced by the ASU, the other one adopting a more complicated and electric power consuming process scheme. Both options are based on a two twin trains configuration equipped with a refrigeration package.

A technical/economical evaluation performed in the previous study (IEA GHG – Gasification Power generation study – 2003) by FWI showed the different impacts on the Investment Costs and on the Operating Costs of the two options**.** Based on these evaluations, the option without Nitrogen use is finally selected, for which all the following data now refers to this case.

Equipment Sizes

UOP has provided FW with a full equipment list for each case for the purpose of cost estimation, but, due to reasons of secrecy, this information cannot be released any further without the third parties signing of a secrecy agreement with UOP.

Utility Consumptions

LP Steam, t/h	83
Cooling Water, m^3/h	4274 (ΔT 12 °C)
Purge Water, m^3/h	1.0
Electric Power, kW	32,875 (Refrigeration Package: 41%)
Solvent Make-up, m^3/yr	121

Material Balance

kmol/h	Untreated Gas	Recycle Gas SRU	Treated Gas GT	CO_2 to Compr.	Acid Gas
CO_2	13,799.2	454.8	1,426.6	12,583.4	244.0
H_2S+COS	67.0	2.4	0.0	1.2	68.4
H_2O	115.6	1.4	6.2	35.4	21.8
N_2	1,151.8	59.4	1,202.8	8.4	0.0
CO	935.6	0.2	911.8	23.8	0.2
H_2	21,026.8	13.8	20,818.0	221.4	1.0
Ar	179.0	0.6	175.6	4.0	0.0
Others	1.8	0.0	0.0	0.0	1.8
Total Flow, kmol/h	37,276.8	532.6	24,541.0	12,877.6	337.2
Pressure, bar g	26	26	25.2	(1)	0.8
Temperature, °C	38	38	34	(1)	49

Note (1): CO_2 stream is the combination of three different streams delivered at Unit B.L. at different conditions.
(2): Material balance relevant to both trains.

The proposed process reaches an H_2S+COS concentration of the treated gas exiting the unit of 3 ppm. This result is due to the integration of the CO_2 removal section with the H_2S removal section, which corresponds to a large circulation of the solvent. The CO_2 removal rate is more than 91% as required, allowing to reach an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

These excellent performances on both the H_2S removal and CO_2 capture are achieved with a large power consumption.

The acid gas H_2S concentration is more than 22 % dry basis, suitable to feed the oxygen blown Claus process.

Together with CO_2, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 221 kmol/h of Hydrogen, corresponding to 1.7% vol and to an overall thermal power of 14.9 MWth, i.e. equivalent to approx 4.8 MWe.
- a very low quantity of H_2S, corresponding to a concentration of 90 ppm.

4.2 Linde (Rectisol)

General Information

On November 2004, for a previous study made by FWI for IEA GHG, Linde provided for each case a set of information which allowed FW to fully evaluate the performance and investment costs of the AGRU and how this section meets the technical and economic targets of the entire IGCC plant. This information has been provided under a non-disclosure agreement between FW and Linde. As a consequence, this report includes only the data that Linde allows to be disclosed to IEA without a non-disclosure agreement between IEA and Linde. The workup of the data presented here is though based on a full set of data provided by Linde to FW.

The solvent used is chilled methanol (technical grade "A") with the advantages of ready availability, high stability and good solubility characteristics for CO_2 and H_2S/COS. The application of the Rectisol process is especially adequate for gases with high sour gas concentration and high pressure.
Due to the high absolute solubility of CO_2 and H_2S in methanol the solvent circulation rate is relatively small compared to other possible washing systems. This results in rather low utility consumption figures (e.g. steam, cooling water, electric energy).

4.2.1 GEE Case

Process Description

The high syngas pressure (56 barg) and the extremely high CO_2/H_2S ratio (183/1), make the Rectisol process suitable to meet the AGR Unit specification. The unit is based on a double train configuration.

Equipment List

Linde has provided FW with a list of equipment for each case for the purpose of cost estimate, but, due to reasons of secrecy, this information cannot be released any further without the third parties signing of a secrecy agreement with Linde.

Utility Consumptions

VLP Steam, t/h	17.0	
LP Steam, t/h	10.9	
Cooling Water, m^3/h	660	(ΔT 12 °C)
Electric Power, kW	9,900	(refrigeration Package: 35%)
Solvent Make-up, t/yr	1,410	

Material Balance

	Untreated Gas	Recycle Gas SRU	Treated gas GT	CO_2 to compress.	Acid Gas
kmol/h					
CO_2	14,992.4	313.2	1,373.0	13,783.6	149.0
H_2S+COS	82.1	3.1	0.0	0.6	84.5
H_2O	70.8	0.9	0.0	0.0	0.0
N_2	253.5	70.7	320.2	4.0	0.0
CO	1,058.6	0.1	1,034.1	25.7	0.0
H_2	20,516.7	16.3	20,480.2	52.8	0.1
Ar	294.5	0.8	286.2	9.1	0.0
Others	8.0	0.0	7.6	0.9	0.0
Total flow (kmol/h)	37,277	405	23,501	13,877	234
Pressure, bar g	57.2	57.2	55.2	(1)	1
Temperature, °C	38	38	29	(1)	34.2

Note (1): CO_2 stream is the combination of two different streams delivered at Unit B.L. at different conditions.

The proposed process reaches an H_2S+COS concentration of the treated gas exiting the Unit of 1 ppmv. The CO_2 removal rate is more than 91% as required, allowing reaching an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

The acid gas H_2S concentration is 36% dry basis, leading to a big advantage for the downstream oxygen blown Claus process.
Together with CO_2, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 52.8 kmol/h of Hydrogen, corresponding to 0.4% vol and to an overall thermal power of 3.6 MWth, i.e. equivalent to more than 1.2 MWe;
- A very low quantity of H_2S, corresponding to a concentration of 27 ppmv.

4.2.2 <u>Shell Case</u>

Equipment List

Linde has provided FW with a list of equipment for each case for the purpose of cost estimation, but, due to reasons of secrecy, this information cannot be released any further without the third parties signing of a secrecy agreement with Linde.
The unit is based on a triple train configuration.

Utility Consumptions

VLP Steam, t/h	37.2	
LP Steam, t/h	12.7	
Cooling Water, m^3/h	1360	(ΔT 12 °C)
Electric Power, kW	16,900	(refrigeration Package: 40%)
Solvent Make-up, t/yr	1,800	

Material Balance

	Untreated Gas	Recycle Gas SRU	Treated gas GT	CO_2 to compress.	Acid Gas
kmol/h					
CO_2	13,799.2	164.2	1,244.2	12,599.9	119.2
H_2S+COS	67.1	5.8	0.0	0.8	72.1
H_2O	115.6	1.4	0.0	0.0	0.0
N_2	1,151.8	55.8	1,202.4	5.2	0.0
CO	935.6	0.1	924.8	11.3	0.0
H_2	21,026.8	12.9	21,017.7	22.0	0.02
Ar	178.9	0.6	177.2	2.3	0.0
Others	1.9	0.0	1.8	0.1	0.0
Total flow (kmol/h)	37,277	241	24,568	12,642	191
Pressure, bar g	26	26	24.25	(1)	1
Temperature, °C	38	38	29	(1)	34.2

Note (1): CO_2 stream is the combination of two different streams delivered at Unit B.L. at different conditions.

The proposed process reaches an H_2S+COS concentration of the treated gas exiting the Unit of 1 ppm. The CO_2 removal rate is more than 91% as required, allowing reaching an overall CO_2 capture of 85% with respect to the carbon entering the IGCC.

The acid gas H_2S concentration is 38% dry basis, leading to a big advantage for the downstream oxygen blown Claus process.
Together with CO_2, the following quantities of other components are sent to the final CO_2 destination, after compression:

- 22 kmol/h of Hydrogen, corresponding to 0.2% vol and to an overall thermal power of 1.5 MWth, i.e. equivalent to more than 0.5 MWe;
- A very low quantity of H_2S, corresponding to a concentration of 20 ppmdv.

5.0 RESULTS COMPARISON

5.1 Scheme Performance

Due to different characteristics of the processes, the schemes have different performances in terms of acid gas compositions when compared. The syngas fed to the AGR section is the same in composition and flowrate in order to allow a fair comparison.
The following tables show these differences between the selected AGR technologies and for each alternative gasification technologies.

5.1.1 Clean Syngas

H_2S + COS concentrations (ppmv) (target specification < 40ppmv):

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	4.3	1
Shell case	3.3	1

CO_2 removal (mol%) (specification 91%):

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	91.3	91.9
Shell case	91.2	91.3

5.1.2 Acid Gas (H_2S Rich)

H_2S concentration (target specification: >15-20 mol%):

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	17.8	36.1
Shell case	20.3	37.7

5.1.3 Acid Gas (CO_2 Rich or Combined Cases)

H_2S+COS concentration (ppmv) (specification 100ppm max):

	UOP (Selexol)	Linde (Rectisol)
GEE case	92.2	43.2
Shell case	93.2	63.3

5.2 Equipment List

UOP and Linde have provided FW with an equipment list for each case for the purpose of cost estimate, but, due to reasons of secrecy, this information cannot be released any further without the third parties signing of a secrecy agreement with UOP and Linde.

For this reason the data relevant to equipment cannot be shown for each case in this study.

Capital costs are compared within section 5.4.

5.3 Utility Consumptions

The following tables summarise the utility consumptions (all trains) for the various technologies for each case as appropriate:

5.3.1 Steam

All flows in t/h

	UOP (Selexol)		Linde (Rectisol)	
	VLP steam	LP steam	VLP steam	LP steam
GEE case	-	70.3	17.0	10.9
Shell case	-	83.0	37.2	12.7

5.3.2 Power

All consumptions in kW.

AGR section:

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	32,100	9,900
Shell case	32,875	16,900

CO_2 compression section:

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	38,115	52,070
Shell case	32,975	52,035

Total Power consumption:

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	70,215	61,970
Shell case	65,850	68,935

5.3.3 Cooling Water

All flows in m^3/hr (12°C temperature rise).

	UOP (Selexol)	**Linde** (Rectisol)
GEE case	2,966	660
Shell case	4,274	1,360

5.4 Capital Costs

The following are the capital costs for each case; UOP Selexol and Linde Rectisol are compared taking into account also their impact on the downstream units (SRU and Tail Gas Treatment, CO_2 compression and drying and Hydrogen production).
The investment cost for Selexol (both GEE and Shell cases) has been developed using FWI proprietary software. The computerized system allows to estimate complete units starting from preliminary technical information.
The investment cost for Rectisol (both GEE and Shell cases) are derived from FWI in house data, obtained for other studies by an investment cost calculation software based on dimensions of main equipment, properly escalated to year 2006 and adjusted taking into account syngas flowrate, pressure, purity and quantity of CO_2 and H_2S removed.

- **Option 1 – Based on Selexol washing;**
- **Option 2 – Based on Rectisol washing.**

	GEE CASE		SHELL CASE	
CAPEX	**Option 1 Selexol**	**Option 2 Rectisol**	**Option 1 Selexol**	**Option 2 Rectisol**
AGR Investment Cost, €	86,944,380	162,295,393	135,727,260	237,019,794
SRU&TGT Investment Cost, €	51,205,380	24,716,340	44,019,960	23,521,620
CO_2 compr. Investment Cost, €	39,840,640	48,826,400	35,779,520	47,700,800
PSA Plant Investment Cost, €	17,861,520	17,977,800	18,169,320	18,282,180
TOTAL, €	195,851,920	253,815,933	233,696,060	326,524,394
DIFFERENCE, €	**-57,964,013**		**-92,828,334**	

For both gasification technologies, the CAPEX comparison is in favour of Option 1 – Selexol (saving respectively 58.0 MM€ and 92.8 MM€ in GEE and Shell case).

5.5 Operating Costs & Option Selection

The operating costs have been evaluated on the following basis:

- Hours of operation: 7446 h/year;
- Years of operation: 6 years (payback target);
- VLP cost: 9 €/t;
- LP cost: 11 €/t;
- Power cost: 0.06 €/kWh;
- H_2 selling price: 0.095 €/Nm^3.

	GEE		Shell	
OPEX h/y 7446	Selexol €/y	Rectisol €/y	Selexol €/y	Rectisol €/y
Acid Gas Removal				
VLP	0.0	1,139,000	0.0	2,493,000
LP	5,779,000	896,000	6,823,000	1,044,000
Power (1)	14,446,000	4,446,000	14,839,000	7,598,000
Solvent losses	811,000	423,000	818,000	540,000
SRU & TGT				
Power (2)	1,072,000	849,000	313,000	371,000
CO2 compression				
Power (1)	17,028,000	23,264,000	14,731,000	23,246,000
Total operating costs	39,136,000	31,017,000	37,524,000	35,292,000
H2 production				
Hydrogen sold	-137,327,000	-138,697,000	-140,984,000	-142,339,000
Delta opex		9,489,000		3,587,000

(1) Including cooling water pump
(2) Only Recycle tail gas compressor considered
(3) Minus prior to figure means that is not a cost, but a revenue

For both gasification technologies, the OPEX comparison is in favour of Option 2 – Rectisol (saving respectively 9.5 MM€/y and 3.6 MM€/y in GEE and Shell case).

From the comparison of OPEX and CAPEX, the pay back time for Rectisol in GEE case is approx 6 years, while for Shell case is more than 20 years. This is due both to the investment cost and to the operating costs: for the two items, the Rectisol technology in Shell case appears penalised by high investment costs and low difference in operating costs with respect to Selexol case.
For GEE case, the two configurations are almost similar and the pay back time is close to the years of operations.
For these reasons the Selexol based AGR is preferred both for GEE and for Shell gasification technology.

CLIENT	:	IEA GREENHOUSE GAS R&D PROGRAM
PROJECT NAME	:	HYDROGEN AND ELECTRICITY CO-PRODUCTION
DOCUMENT NAME	:	HYDROGEN STORAGE

ISSUED BY	:	L. VALOTA
CHECKED BY	:	P. COTONE
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Date	Revised Pages	Issued by	Checked by	Approved by
April 2007	Draft	L.Valota	P.Cotone	S.Arienti
July 2007	Rev. 1	L.Valota	P.Cotone	S.Arienti

HYDROGEN STORAGE

INDEX

1.0 INTRODUCTION

Hydrogen is currently used on a large scale in ammonia plants and modern petroleum refineries. In the future it may also be used an as energy carrier for vehicles, distributed heat and fuel cells power generation. Moreover hydrogen can be stored without relevant technical problems, for example in above ground and underground storages.

The long term goal is to produce hydrogen from renewable energy sources but in a near term the cheapest way to produce hydrogen with low CO_2 emissions is expected to be by use of fossil fuels with CO_2 capture and storage.

Hydrogen can be produced in stand-alone plants but it may be advantageous to co-produce hydrogen and electricity, following the energy consumption. Thus, in order to constantly match the demand, hydrogen storage has to be considered.

In this attachment, different storage technologies are described, focusing on advantages and disadvantages for storage for large amounts of hydrogen. Estimation of the costs is also provided for different storage options. Finally relevant data from state-of-the-art hydrogen storage experiences are provided and an explanation of the criterion of choice is presented.

2.0 HYDROGEN STORAGE OPTIONS

The main options for storing hydrogen are as a compressed gas (above ground or underground), as a liquid or in metal hydrides. Metal hydride option has not been considered since it's not suitable to large quantities of hydrogen. Above ground storage (compressed gas tanks) and underground storage (geological), although they are based on the same principles, have been separately analysed due their strong technological differences. Finally a series of considerations on hydrogen pipeline storage is detailed in para. 2.4

2.1 Above ground compressed gas storage

Compressed gas storage of hydrogen is the simplest storage solution and the most traditional way. The only equipment required is a compressor and a pressure vessel. The main advantages are simplicity, practically indefinite

storage time and no purity limits on hydrogen. The main problem with compressed gas storage is the low storage density, which depends on the storage pressure thus on tank materials. Low-pressure spherical tanks can hold as much as 1,300 kg of hydrogen at 12-16 bar [1]. High-pressure storage vessels have maximum operating pressures of 200-300 bar [2]. European countries tend to use low pressure cylindrical tanks with a maximum operating pressure of 50 bar and storage capacities of 115-400 kg of hydrogen [2]
A review of existing plants shows capital costs that vary from $1,250 to $4,160 per kg of hydrogen storage [3] (updated at 2007).
In many cases, small tanks are rented by the gas supplier for a couple thousand dollars per month [2]. Operating costs are around 1.04 $/kg excluding the compressor energy [3].

2.2 Liquefied gas storage

Liquid hydrogen has been used as a fuel in space technology for several years. It has a low density and has less potential risks in terms of storage pressure compared with the compressed gas. However, the hydrogen liquefies at -252.9°C and thus the storage vessels require cryogenic systems and sophisticated insulation techniques.

Liquefaction is done by cooling a gas to form a liquid. Liquefaction processes use a combination of compressors, heat exchangers, expansion engines, and throttle valves to achieve the desired cooling [4]. The simplest liquefaction process is the Linde cycle or Joule-Thompson expansion cycle. In this process, the gas is compressed at ambient pressure, then cooled in a heat exchanger, before passing through a throttle valve where it undergoes an isenthalpic Joule-Thompson expansion, producing some liquid. This liquid is removed and stored while the cool gas is returned to the compressor via the heat exchanger [4]

A major concern in liquid hydrogen storage is minimizing hydrogen losses from liquid boil-off. Because liquid hydrogen is stored as a cryogenic liquid that is at its boiling point, any heat transfer to the liquid causes some hydrogen to evaporate, causing boil-off and hydrogen leakage.

Thus, even if liquid hydrogen has the highest storage density of any method, it also requires an insulated storage container and an energy intensive liquefaction process.

Operating costs are 2.4 $/kg while capital cost varies from 2000 to 40,000 $/kg in dependence of the cryogenic system and of the size and type of the tank [3] (updated at 2007).

2.3 Underground gas storage

Underground caverns have been used for years for methane storage and they have already been proven to be a cheap and relatively easy method for large-scale storage of gas. Depending on the geology of an area, underground storage of hydrogen gas is possible [2]. Moreover, underground storage of helium, which diffuses faster than hydrogen, has been practiced successfully in Texas [1].

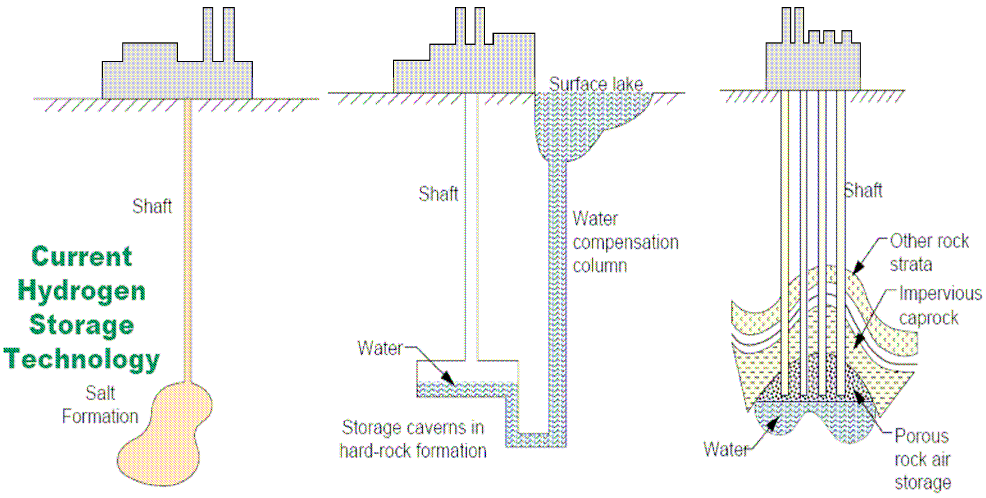
There are three types of underground hydrogen storage facilities and these are strictly related to the geological configuration of the site (Fig 1) [5]

Manmade caverns. Underground caverns are mined with access to the surface with wells. The most common type of cavern is in salt domes, often found in the form of layers that can be hundreds of meters thick. The principle consists in dissolving the salt with fresh water and removing the brine via a single well, which then serves for gas injection and withdrawal. The storage capacity for a given cavity volume (several hundred thousand cubic metres) is proportional to the maximum operating pressure, which depends on the depth. They offer several advantages: high deliverability, high degree of availability, short filling period, total recovery of cushion gas with brine injection. When gas is stored, the gas pressure depends upon the inventory of the cavern [12,14].

Pressure-compensated manmade caverns. Underground caverns are mined with access to the surface with wells. In addition, a surface lake connected to the bottom of the manmade cavern is created. The water pressure from the surface lake results in a constant pressure in the cavern that is equal to the hydraulic head of the water. The compressed gas is stored and delivered at a constant pressure. This option requires a rock that does not dissolve in water.

Porous rock with cap rock. In many parts, porous rock exists with an impermeable cap rock above it that forms a natural trap for gases (inverted "U" shape). Wells are drilled into the porous rock, and injected gas pushes out whatever other fluids exist in the porous rock. Much of the world's natural gas is found in this type of geological trap. Because the natural gas has been trapped for tens of millions of years, nature has demonstrated that the cap rock is extremely impermeable to fluids.

Fig 1: Different technologies for underground hydrogen storage



The pressure in underground caverns depends on the kind of storage and the geological site. Anyway it has to be lower than the overburden pressure due to the load of the rock column (around 0.226 bar/m)[6]. Giving all the different geologies and situations, a reasonable estimation of the operating pressure could be from a value of 70 bar up to 180 bar.

One concern with large storage vessels (especially underground storage) is the cushion gas that remains in the empty vessel at the end of the discharge cycle. In small containers this may not be a problem, but in larger tanks this can be as much as 50% of the working volume, or several hundred thousand kilograms of gas. Some storage schemes pump brine into the area to displace the hydrogen, but this increases the operating and capital costs [7,2,1].

Another concern is the purity of hydrogen coming out the geological store. Several gases may be present underground, such as H_2S and CH_4, that can contaminate the hydrogen. Even if this topic is currently under research, data provided by recent underground experiences shows that this problem becomes significant only in case of porous rock with cap rock storage, while it is not relevant to the other types of underground storage.

The losses caused by the leaks in underground caverns are about 1 - 3% of the total volume per year [8,9].

Several underground storages of hydrogen are in operation, providing a sufficient experience on this technology. The city of Kiel in Germany has been storing town gas containing 60–65% of hydrogen in a gas cavern at a depth of 1330 m since 1971 [8, 10]. England and France both have long-term experience in the field of underground hydrogen storage. The British chemical concern ICI stores hydrogen in three brine compensated salt caverns in Teeside, England. The hydrogen is stored at pressures up to 5 MPa in these up to 366 m deep caverns. From 1957 until 1974, Gaz de France stored town gas with 50% hydrogen content in a 330 Mm^3 aquifer storage.

Praxair is constructing a hydrogen storage facility to enable peak shaving, which will be the first of its kind in the industrial gases industry. Located in Liberty County, Texas, the facility will utilize an underground storage cavern. Last but not least, one of the most significant worldwide experiences is the storage operated by ConocoPhillips at Clemens Terminal, Texas [11]. Here, caverns are used as hydrogen buffering for hydrotreaters for a close refinery. It’s composed of two caverns in a salt layer, operating 850 m underground at 150 bar and with a temperature of about 37°C. A solution mined method has been used using fresh water (6 volumes of water removes 1 volume of salt). The limitation in depth is due to the pressure required to pump brine out the cavern to recover hydrogen. It has a total physical volume of 580,000 m^3. The maximum fill rate is around 2,960 kg/h while the maximum discharge rate is around 4,960 kg/h. The cavern leaks less than 1.2 g/minute. It requires maintenance every 6 months to the valves; every 10 years it has to empty and maintenance has to be performed for 6 weeks. The system is provided with two reciprocating, two stages, compressors. It has been regulated by Texas RR Commission.

Several studies have been recently performed in order to evaluate the quantity of hydrogen that can be reasonably stored underground in different geological sites. One of the most important has been focused on UK geological conditions [13].

There is still a debate on hydrogen underground costs. The ConocoPhillips experience shows a cost of 0.80-1.60 $/kg [11] while other studies set the cost from 5 $/kg to 40 $/kg [3] (updated at 2007).

2.4 Pipeline storage

Hydrogen pipelines have been operated safely over scores of years. Most of them were built not for hydrogen usage, and have been recently converted in order to follow the increasing interest of the petroleum industry in hydrogen derived products. Pipeline storage of hydrogen is based, exactly like happens for natural gas, on increasing the pressure in order to allow more gas in the line. Even if for other gasses the high pressures do not present a relevant technical issue, for hydrogen an increase of pressure could present some materials problems [15-19].
In view of the importance of the topic, different companies and research institutes, are releasing technical considerations that can result in these conclusions:

- Existing steel pipelines are subject to hydrogen embrittlement and are inadequate for widespread H_2 high pressure distribution.
- Current joining technology (welding) for steel pipelines is major cost factor and can exacerbate hydrogen embrittlement issues.
- New H_2 pipelines will require large capital investments for materials, installation, and right-of-way costs
- H_2 leakage and permeation pose significant challenges for designing pipeline equipment, materials, seals, valves and fittings
- H_2 delivery infrastructure will rely heavily on sensors and robust designs and engineering.

Thus, at this time, hydrogen storage in pipelines is not a reliable option for the purpose of this study.

3.0 Conclusions

Several parameters have to be considered in order to choose the best storage option. In particular it depends on the application that the hydrogen is stored for, the required energy density, the storage period (on daily basis or seasonal), maintenances, safety issues and finally costs.

Based on these considerations, the following general conclusions can be made:

- Underground storage is convenient for large quantities of gas, long-term storage;
- Aboveground compressed gas is suitable only for small quantities of gas and short period due to their very high costs;
- Liquid hydrogen has specific applications related to its high degree of safety and low storage density but requires expensive cryogenic facilities.

For the scope of this study the underground storage is the best solution in relation to the very large volumes of hydrogen to be stored for long periods.

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UNDERGROUND STORAGE GLOSSARY

Total gas storage capacity is the maximum volume of gas that can be stored in an underground storage facility by design and is determined by the physical characteristics of the reservoir and installed equipment.

Total gas in storage is the volume of storage in the underground facility at a particular time.

Base gas (or cushion gas) is the volume of gas intended as permanent inventory in a storage reservoir to maintain adequate pressure and deliverability rates throughout the withdrawal season.

Working gas capacity refers to total gas storage capacity minus base gas.

Working gas is the volume of gas in the reservoir above the level of base gas. Working gas is available to the marketplace.

Deliverability is most often expressed as a measure of the amount of gas that can be delivered (withdrawn) from a storage facility on a daily basis. Also referred to as the deliverability rate, withdrawal rate, or withdrawal capacity, deliverability is usually expressed in terms of millions of cubic feet per day (MMcf/day). The deliverability of a given storage facility is variable, and depends on factors such as the amount of gas in the reservoir at any particular time, the pressure within the reservoir, compression capability available to the reservoir, the configuration and capabilities of surface facilities associated with the reservoir, and other factors. In general, a facility's deliverability rate varies directly with the total amount of gas in the reservoir: it is at its highest when the reservoir is most full and declines as working gas is withdrawn.

Injection capacity (or rate) is the complement of the deliverability or withdrawal rate–it is the amount of gas that can be injected into a storage facility on a daily basis. As with deliverability, injection capacity is usually expressed in MMcf/day, although dekatherms/day is also used. The injection capacity of a storage facility is also variable, and is dependent on factors comparable to those that determine deliverability. By contrast, the injection rate varies inversely with the total amount of gas in storage: it is at its lowest when the reservoir is most full and increases as working gas is withdrawn.