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Cost Estimates of Coal Gasification for Chemicals and Motor Fuels

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1. Introduction

Solid fuels gasification technology has been understood and applied for a long time. The current directions in developing coal gasification technology are primarily related to power generation in combined systems involving steam and gas turbine implementation, which considerably increases fuel use efficiency. Compared to the first gasifying installations, the current solutions have a much higher conversion intensity and are more reliable. Integrated power generation-related gasification technology developments have created increased interest in chemical products, such as liquid motor fuels, methanol and hydrogen. At the present time, the basic reason for the increase of coal use as a raw material for chemical production is the dynamic industrial growth in countries with high economic potential that do not have their own natural gas and oil resources and have limited access to international sources of the above minerals. China is a good example of a country in this situation, and it constitutes the largest coal gasifying economy in the world. In China alone, more than 100 million tonnes of coal is gasified yearly. We expect that countries such as the USA and India will follow China in coal gasification-based production growth.

The crucial driver of gasification technology development is the necessity of a drastic reduction in CO₂ emission from anthropogenic sources, which is considered to be one of the main contributors to the greenhouse effect. Among fossil fuels, the most important CO₂ emitter is coal, which is characterised as having the highest concentration of carbon element compared to its caloric value. In the coal gasification process, carbon dioxide is removed from the processed gas by the absorption of acid components, which constitutes an inherent part of the technology. In case of chemical plant the acid gases, i.e. H₂S and CO₂ must be removed from the processed gas, regardless of the chemical facility's production profile because H₂S can damage the catalysts used during chemical synthesis, and the content of CO₂ is corrected to the expected composition of a synthesis gas. This removal step can

alleviate the need for additional CO₂ separation so that the costs associated with dehydration and compression are the primary costs remaining. These two processes are critically important to system, as they ensure safe transport of the CO₂ to the storage (sequestration) area.

In the case of integrated gasification combined cycle (IGCC, power generation), the removal of sulphur compounds (H₂S, COS) is required to protect the gas turbine, and CO₂ removal is conducted only to reduce atmospheric emissions. However, because of the high concentrations of carbon dioxide and the high-pressure of the treated gas, the removal of CO₂ from syngas (i.e., pre-combustion removal) is less expensive than if the CO₂ were separated from the flue gases (post-combustion removal). Pre-combustion CO₂ removal results in better process and economic efficiency of IGCC systems (in case of CO₂ sequestration) compared to conventional power plants based on coal combustion.

The development state of coal gasification technology

A review of the global development state of gasification technologies has been performed based on a 2010 database developed by the U.S. Department of Energy (US Department of Energy & National Energy Technology Laboratory [US DOE & NETL], 2010a). The results of this analysis have been compared in three categories characterising the current status of technology development: plants that are operational, plants that are under construction (or start-up) and, plants that are in the development phase (this category includes plants in varying degrees of implementation, including plants at the stages of planning, conceptual work and designing). When analysing the data for the various systems, plants that use natural gas as a fuel have been omitted as these plants are not considered to be gasification systems but rather are plants for the partial oxidation of natural gas. The total power of the aforementioned systems (the thermal capacity of syngas output) amounts to 15,281 MW_{th}, of which 72 % (10,936 MW_{th}) is attributed to a plant using a Shell pressure reactor that is under construction in Qatar.

The published data show that there are 116 gasification plants equipped with 342 reactors with a total power of 50,104 MW_{th} are currently operating worldwide. The total power represents the chemical energy in the gas that is produced but does not include the systems for the partial oxidation of natural gas. Seventeen systems are under construction (28 gasification reactors, 16,289 MW_{th}; coal), and 37 plants (76 reactors, 40,432 MW_{th}) are at the planning stage with systems to be implemented in the years 2011 – 2016. Since the last review in 2007, the installed power increased by 7 %, resulting in the largest recorded increase for coal gasification at 18 %. For other fuels, there was a clear decrease in the amount of gas produced, particularly for biomass and petroleum coke (-68 % and -37 %, respectively) (Table 1). The implementation of all current and planned investment projects will contribute to more than a doubling of gas production (106,825 MW_{th}).

The largest percentage of gasification systems is operating in the Asia and Oceania region (39 % of total global gas production), primarily because of extremely dynamic technology developments in China (78 % of this region). In this region, which includes China, Australia,

South Korea and Vietnam, the majority of systems are now under construction and planned for implementation in the next few years. Long-term plans exist for technology development in North America (primarily in the U.S.), the implementation of which would move this region into second place in the global production of gas from gasification (30.4 % of global gas production).

Coal, the basic feedstock for gasification, is used in gasification plants that are currently operating and accounts for 61.6 % of global gas production (Fig. 1). Petrochemical industry by-products rank second (35.8 %), and the remaining 2.6 % of gas production is attributed to petroleum coke and biomass. For plants that are under construction and planned for start-up by 2016, the role of coal as the basic fuel will be maintained, and the share of gas produced from this raw material will increase to 79 %.

The basic products of operational systems using gasification processes comprise chemicals such as ammonia, hydrogen and oxy-chemicals (46 % of world gas productions), products of Fischer-Tropsch synthesis (30 %), power (16 %) and gaseous fuels (8 %) (Fig. 2). Chemicals will also be the main products of the plants that are under construction (72 %). In the case of plants planned for implementation, the largest share will be power-generating plants (37.5 %), which is probably related to the attractiveness of power systems that are integrated with gasification, particularly in the context of the necessity for CO₂ emission reduction (Fig. 2).

Feedstock		Operating 2010	Operating 2007	Difference	%
		(operating, construction, start-up)	(operating, construction, start-up)		
Coal	MW _{th}	36,315	30,825	5,490	18
	Gasifiers	201	212	-	-
	Plants	53	45	-	-
Petroleum	MW _{th}	17,938	18,454	-516	-3
	Gasifiers	138	145	-	-
	Plants	56	59	-	-
Petcoke	MW _{th}	911	1,441	-530	-37
	Gasifiers	5	8	-	-
	Plants	3	5	-	-
Biomass/ waste	MW _{th}	373	1,174	-801	-68
	Gasifiers	9	21	-	-
	Plants	9	13	-	-
Total	MW _{th}	55,537	51,894	3,643	7
	Gasifiers	353	386	-	-
	Plants	121	122	-	-

Table 1. Comparison of the state of worldwide existing gasification technologies in the years 2007 and 2010 (US DOE & NETL, 2007, 2010a).

In the case of coal use, the most popular gasification plants are now fixed bed gasification technology, which is practically no longer developed but still accounts for 57 % of gas production due to the high manufacturing potential of the Sasol plant in South Africa.

Processes using entrained flow reactors are the most intensively developed technologies (operating plants, 43 % of gas production) as confirmed by the projects that are under construction and planned for start-up by 2016, which are nearly all related to this reactor design. Fig. 3 shows the structure of the operational plants and the coal gasifiers planned for start-up in terms of the technological solutions used.

Of the technologies used for coal gasification in entrained flow reactors (operating plants), the Shell (dry feeding) and GE/Texaco (slurry feeding) have the dominant share in gas production (77 %), followed by the ECUST (15.3 %) technology. The third place position of the use of ECUST (East China University of Science and Technology) technology in developing plants is noteworthy because of the rapid pace of the ECUST technology development. Beginning with a pilot plant (22 t/d of fuel) in 1996, the technology led to operational demonstration plants in the years 2001 – 2005 (750 and 1,150 t/d of coal) and 17 commercial gasifiers that were implemented by 2010 (capacity of up to 2,000 t/d of coal) (Liu, 2010).

The highest percentage of plants planned for start-up that are under construction and in the development phase will use the Shell gasification technology (26.7 %; 11,913.2 MW_{th}) followed by ECUST (20.8 %), Udhe PRENFLO (16.8 %), Siemens (14.7 %), ConocoPhillips E-Gas (11.3 %), GE Energy/Texaco (5.3 %) and MHI (3.7 %) (Fig. 3). The fluidised bed coal gasification reactor technologies GTI U-GAS and TRIG (KBR Transport Gasifier) will be developed apart from the entrained flow technologies.

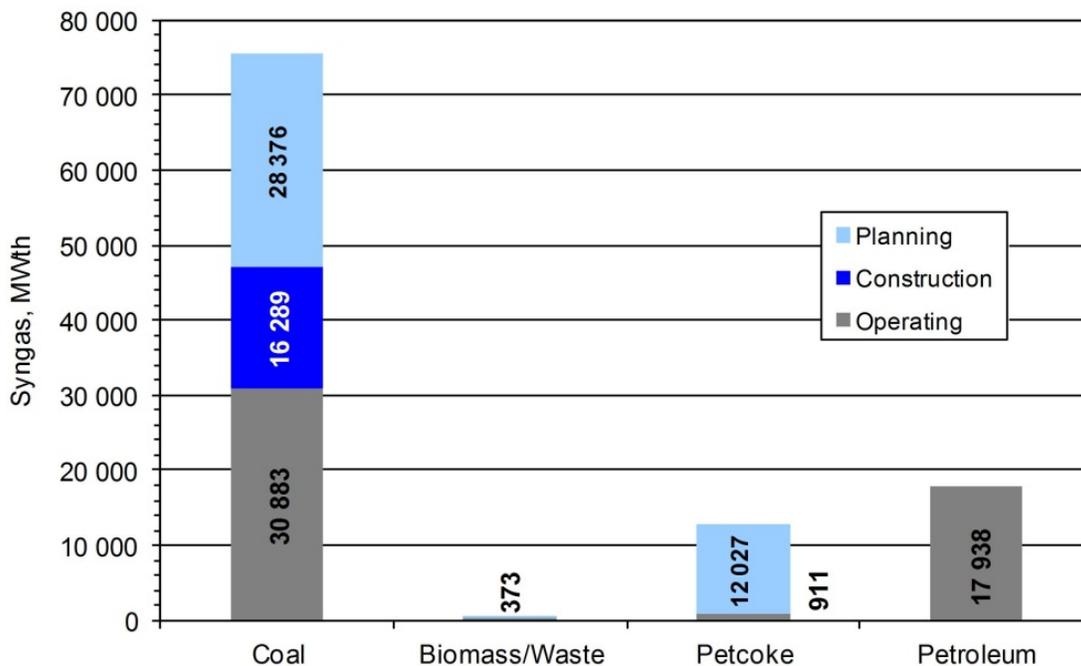


Figure 1. Total capacity of gasifiers versus fuel used (current and forecast by 2016).

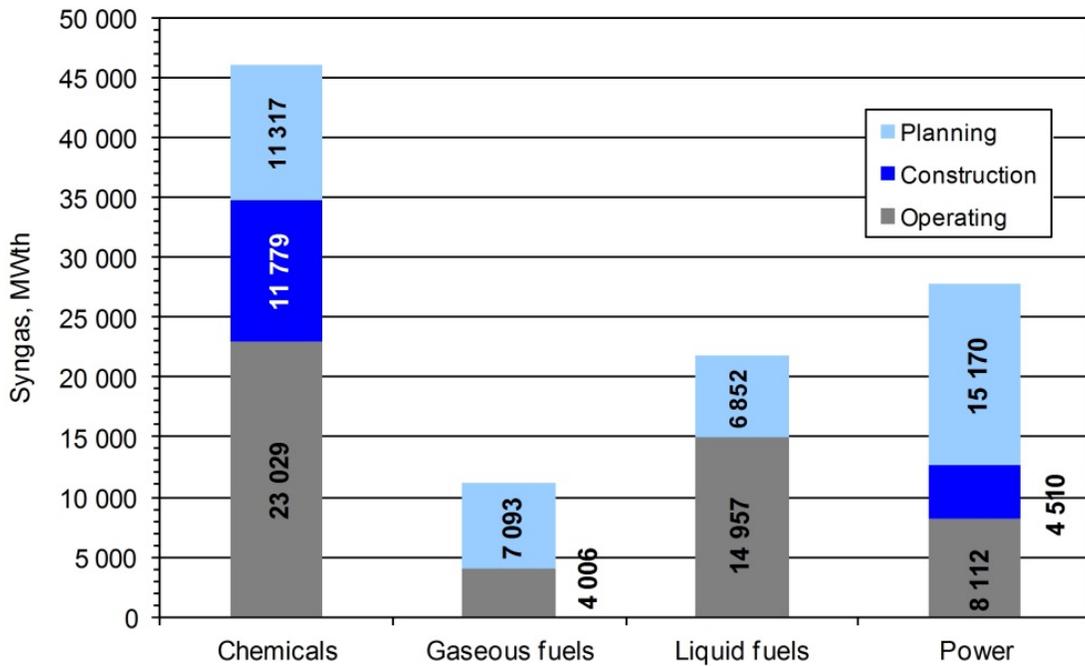


Figure 2. Total capacity of gasifiers versus product manufactured (current and forecast by 2016).

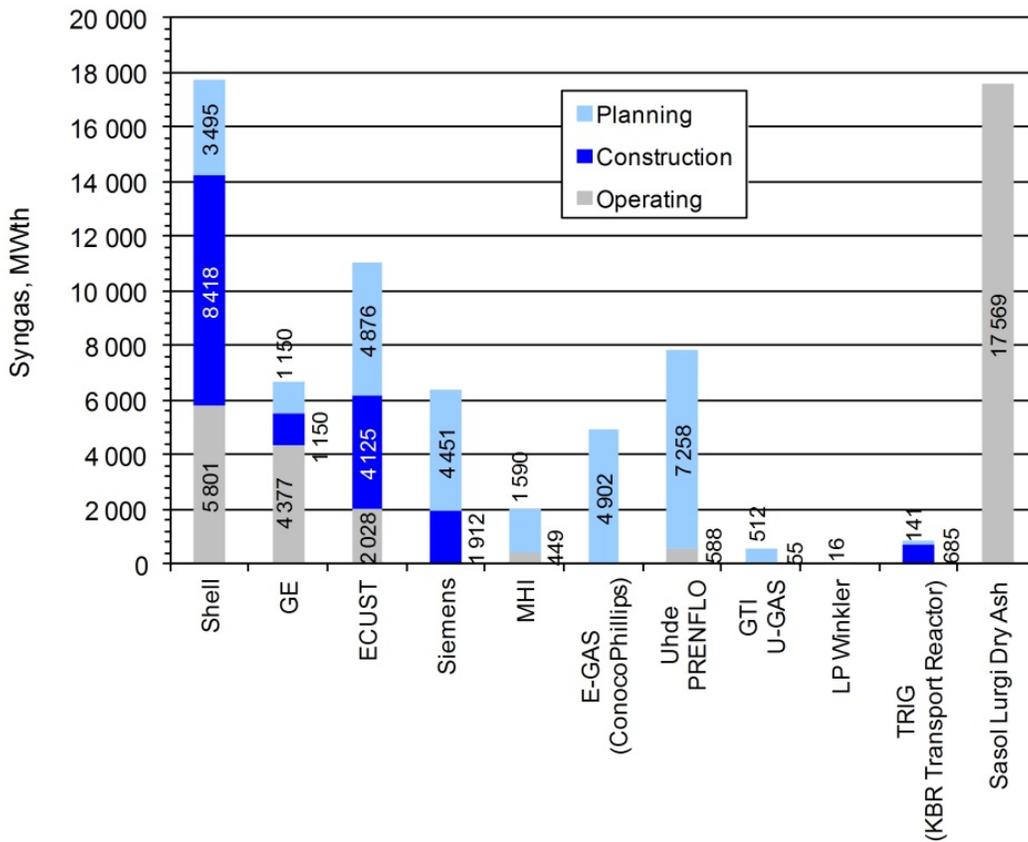


Figure 3. Total capacity of reactors using coal as the main fuel, breakdown by technological groups (current and forecast by 2016).

2. Technological option

The review of the global development state of gasification technologies shows that gasification systems will be used for syngas production in power generation systems (IGCC) and particularly in chemical synthesis to obtain liquid and gaseous fuels including methanol and hydrogen. The analysis of the above processes is the subject of this study. Four cases for coal gasification applications involving chemical synthesis and electricity generation have been analysed and discussed in detail. The options include: liquid fuel production, hydrogen generation, methanol production (options I-III) (Dreszer & Mikulska, 2009), and power and syngas production (Polygeneration Plant, option IV) (Chmielniak et al., 2008; Energoprojekt-Katowice S.A & Institute for Chemical Processing of Coal [EPK & IChPW], 2008).

Option I – A system of six gasifiers, which requires an annual coal consumption of approximately 5,600,000 t/y. The adopted scale of coal processing results from preliminary cost-effectiveness studies for liquid fuel production from coal, which have shown that the operation of a production plant starts to be profitable only at a production level exceeding 1 million tonnes of liquid fuels, which corresponds to the adopted scale of coal consumption. The plant products have been defined as technical propane-butane (LPG), diesel oil and a semi-product for the diesel oil that is not further processed into final commercial products.

Option II and III – one gasifier system. The adopted scale makes it possible to accomplish the following:

- cover the demand for hydrogen on the scale of a single standard chemical plant fertiliser production train (no network for high-volume hydrogen distribution was assumed) (option II).
- produce methanol from the gas originating from coal gasification on the scale of 500,000 t/y (option III).

Option IV – a system of two gasifiers operating in parallel technological trains to produce syngas (for methanol synthesis) and power (IGCC). Due to their identical capacity, gasifiers operating in an integrated system can provide mutual back-up functionality for each other, increasing the annual availability of syngas or electricity production units, depending on the adopted production programme. The scale of production allows to manufacture of approximately 500,000 t/y of methanol what ensures the profitability of its production.

Each of the analysed options consists of a syngas generation unit, i.e., a coal gasification system including units for converting and cleaning syngas.

Gasification technologies in the entrained flow reactors play an essential role in the production of syngas from coal and are offered by a number of providers. The final choice of gasification technology must therefore be made using a separate analysis based on detailed

data from the technology providers, including investment and operational cost and the assessment of coal suitability for processing.

GE/Texaco technology has been selected for the analysis of the considered cases for the following factors:

- mature technology / solution used for the longest recorded period,
- one of the largest shares in the coal gasification sector (33 %, operational plants),
- absence of inert gases in syngas, which constitute a redundant ballast in chemical synthesis and result in an increase in equipment size needed due to the increased gas volume in the circuit.

The disadvantage of this technology is the lower energy efficiency of the gasification process compared to technologies using dry coal feeding. However, it has been assumed that to assess different fuel production systems based on coal gasification using conceptual studies, it will be less risky to assume process guidelines for GE/Texaco technology with a coal-water slurry feeding system.

3. Description of considered technological systems

3.1. Coal gasification – GE/Texaco technology

The coal-water slurry (62-68 % coal) and oxygen from the air separation system are fed through a system of valves to the feedstock injector in the top part of the reactor where gasification proceeds at a temperature of 1,260 – 1,480 °C. The hot processed gas with molten ash flows to the bottom part of the reactor, the radiant cooler, where it is cooled down to approximately 730 °C and then is taken off of the reactor to a convective cooler and a scrubber. After being cooled down to approximately 230 °C, the raw gas is directed to the gas conversion and/or cleaning systems. High-pressure (HP) steam is produced in the radiant and convective coolers. The molten ash flows down to the water bath in the bottom part of the reactor where, after solidification and cooling, it is taken off the system through a lockhoppers. Fly ash that is separated from the gas is also taken off together with slag (the ash separation from the gas occurs through a sudden change of its flow direction before leaving the reactor). After water separation, the slag is directed to a waste landfill. Separated fly ash with a carbon fraction of approximately 30 % is delivered to the coal-water slurry preparation system for recirculation to the reactor. The spraying water from the scrubber and the water from slag dewatering is returned to the scrubber after the removal of solid particles (fine slag/fly ash), and its excess is fed to a water treatment plant (US DOE & NETL, 2002).

In addition to the technology option described above, General Electric commercially offers two other configurations of gasification plants (US DOE & NETL, 2010b):

- a reactor with direct water cooling: in this system, the hot processed gas is cooled down to 260 °C through direct contact with water before leaving the reactor.

- a reactor with a radiant cooler: the processed gas leaving the gasification zone passes through a radiant cooler that produces high pressure steam where it is cooled down to approximately 800 °C and then passes through a water lock, which lowers its temperature to approximately 200 °C.

The gasification pressure was assumed to be 3, 5.6 and 7 MPa for the production systems of liquid fuels, hydrogen and methanol plants, respectively (the pressure was selected to match the process condition for F-T, hydrogen and methanol production units).

Oxygen for the gasification system is supplied from an air separation system based on cryogenic separation. Oxygen purity levels of 99.5 % for liquid fuel production and 95 % for all other cases were assumed.

3.2. System layout – Fuel and chemical production plants

Fig. 4 to 6 present the process diagrams of the considered plants based on coal gasification. The data on the technological configurations are summarised in Table 2.

3.2.1. Liquid fuel production plant

Gas from the gasification system is directed to the hydrolysis reactor where, in the presence of the catalyst, carbonyl sulphide (COS) is hydrolysed to hydrogen sulphide. Gas exiting the COS reactor is cooled to approximately 38 °C in several heat exchangers fed by boiler feed water (steam production) or cooling tower water. Entrained water (condensate) is separated and used for coal-water slurry production and for slag cooling in the gasifier. Cool gas is fed to the Selexol system, where hydrogen sulphide and carbon dioxide are removed. Hydrogen sulphide is directed to the Claus system for sulfur recovery. Clean gas is heated to approximately 313 °C, deep purified from the remaining hydrogen sulphide in the reactor, filled with zinc oxide and fed to Fischer-Tropsch synthesis reactors. Fischer-Tropsch synthesis is carried out in a slurry reactor at 250 °C under a pressure of 2 MPa in a presence of cobalt catalyst. Unreacted part of syngas is fed to the carbon dioxide separation system based on chemical absorption (MDEA) and then to the dehydration and compression system. After passing through the product separation system, the gas is then recirculated to autothermic reforming and sent back to the synthesis reactor. Separated carbon dioxide from the Selexol and amine units is compressed to 12 MPa and transported to a storage location.

3.2.2. Hydrogen production plant

Partially cleaned gas from the gasification island is directed to the Water Gas Shift (WGS) reactor where approximately 97 % of the CO is converted to CO₂ and hydrogen. Gas exiting the WGS reactor is cooled to approximately 38 °C and then fed to the Selexol unit. In the two-stage Selexol system, gas is divided into three streams: sour gas (primarily H₂S), carbon dioxide and hydrogen-rich processed gas. Sour gas from the first stage of the Selexol absorber is directed to the sulfur recovery unit (Claus, Scot). CO₂ is compressed to 12 MPa in preparation for transport and storage. Cleaned processed gas with approximately 90 % hydrogen content is fed to a PSA

(Pressure Swing Adsorption) system, where hydrogen with >99 % purity is produced. The off gas from the PSA system is combusted in a steam boiler, and then steam from the boiler and from the gasification system is used for power generation in the steam turbine.

3.2.3. Methanol production plant

Partially cleaned gas from the gasifier is divided in two streams. One of them, which accounts for approximately 65 % of the total flow, is fed to the high temperature CO shift reactor, where, at temperatures between 400 °C and 410 °C, a carbon monoxide and steam reaction occurs, generating hydrogen and carbon dioxide and producing the required hydrogen concentration in syngas, which is directed to the methanol synthesis reactor. After being cooled to approximately 250 °C, the gas is then joined with the second stream and directed to the COS hydrolysis reactor. Next, hydrogen sulphide and carbon dioxide are removed in the Selexol system from the gas after it is cooled to 38 °C. The hydrogen sulphide that is removed from the gas is then transported to the Claus system for sulfur recovery. Carbon dioxide is separated with 78 % efficiency (the separation level is assumed to meet the stoichiometric ratio required for methanol synthesis $((H_2 - CO_2)/(CO + CO_2) = \text{aprox. } 2)$) and is then compressed to 12 MPa. The composition of the syngas leaving the synthesis system enables its direct use in methanol synthesis. Syngas that is purified in the Selexol process is then joined with circulating tail gas from the synthesis unit and, after being heated to approximately 210 °C, is conducted to the adiabatic, methanol synthesis reactor. The post-reaction mixture leaving the synthesis reactor is then cooled to 38 °C while heating the gas that is being directed to the synthesis reactor, and then it is separated into a liquid methanol fraction and off gas. The liquid fraction is decompressed and transported to a degasifying tank. The raw methanol is then directed to the rectification system, where methanol of high (>99 %) purity level is obtained. Part of the tail gas is compressed and redirected to the methanol synthesis system, and after being decompressed, the remaining gas is combusted in boiler burners where steam is overheated and directed to the steam turbine. The high pressure steam generated in the gasifying system also feeds the turbine.

3.3. Polygeneration plant

A schematic diagram of the Polygeneration Plant is presented in Fig. 6. The system enables simultaneous electricity, heat and syngas generation with sequestration of the carbon dioxide formed during the production process. Joining the combined power and heat generation with syngas production enables the high efficiency of fuel primary energy conversion, low emission indicators and high economic efficiency, also in the case of CO₂ sequestration. The presented solution was developed by Institute for Chemical Processing of Coal (IChPW) and Energoprojekt-Katowice SA (EPK) for TAURON SA (power producer, Poland) and Zakłady Azotowe Kędzierzyn SA (ZAK SA, chemical works, Poland) (Chmielniak et al., 2008 ; EPK & IChPW, 2008).

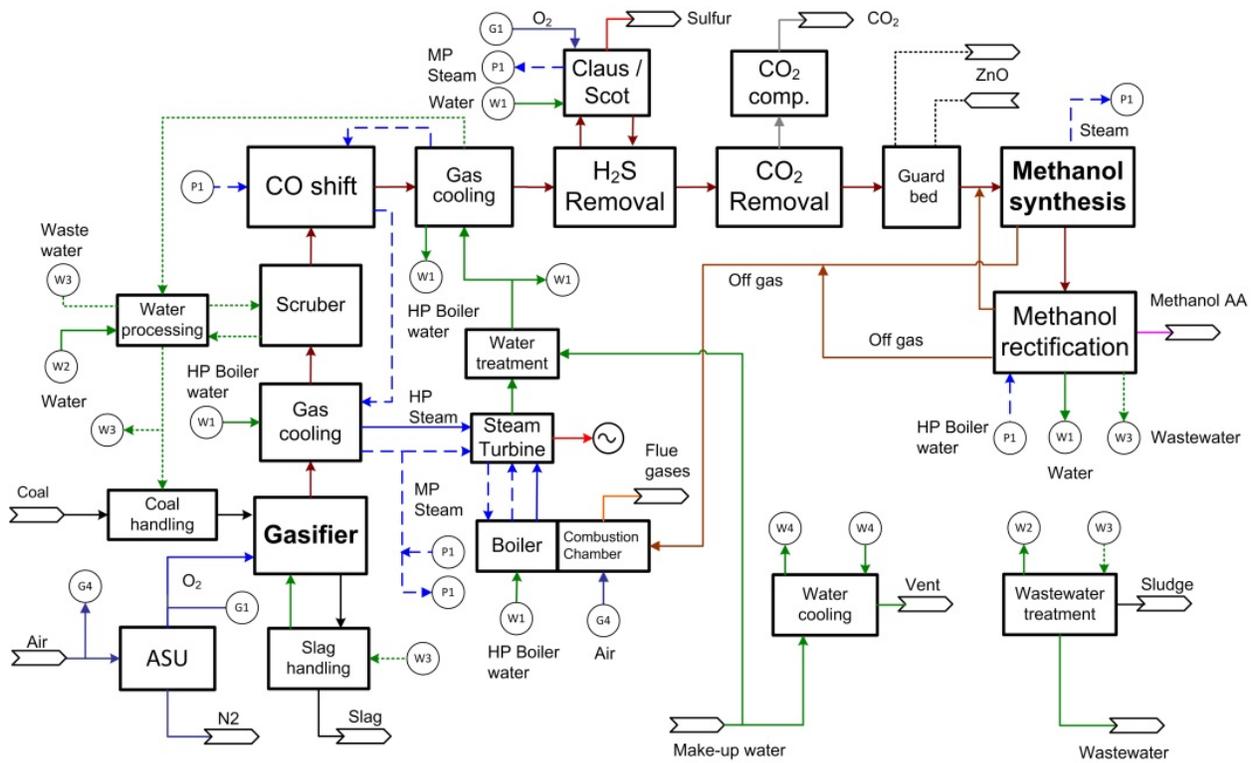


Figure 5. Process diagram of methanol production plant (option III).

To demonstrate an alternative for clean coal technology, the concept of a polygeneration facility assumes possible complete elimination of atmospheric carbon dioxide emissions. Thus, the proper configuration of the IGCC system (energy island, Fig. 6) is necessary for the efficient removal of CO₂ (CO₂ removal in a chemical island is a technological need for the production of syngas). Regarding the IGCC plants that are currently under operation (without CO₂ removal), major changes include the introduction of CO shift reactors and CO₂ separation system. The CO conversion process allows to convert gasifier product (raw gas) to hydrogen-rich syngas and to concentrate most of the carbon contained in the gas in to a CO₂ stream. This allows for the removal of carbon from the syngas before the combustion process (a CO₂ stream is removed in the subsequent stages of syngas processing). Additionally, during the conversion process, the COS hydrolysis reaction takes place without requiring additional equipment (an IGCC facility without CO₂ removal requires systems for the hydrolysis of COS). CO₂ is removed from the syngas during an absorption process. Due to the high pressures under which the gasifier is typically operated, the most energy efficient method of gas separation is by physical absorption. A double stage physical absorption system is recommended for use in a gasification system when separation of CO₂ is required.

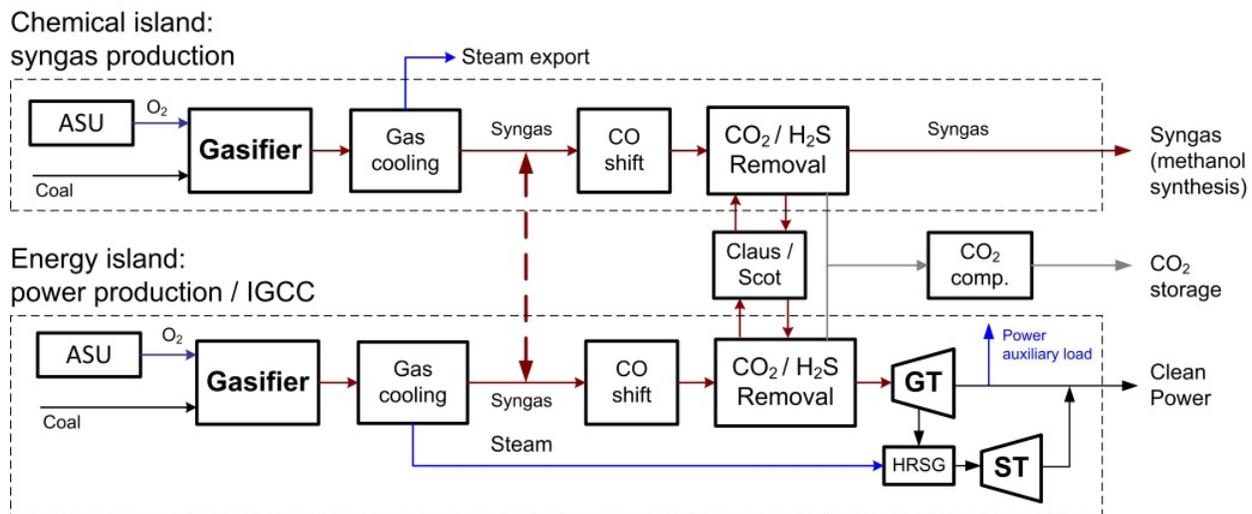


Figure 6. A scheme of Polygeneration Plant.

A Polygeneration Plant consists of three basic technological units:

- Chemical island: coal gasification system that is equipped with a gas conversion and purification system with a CO₂ separation unit and generates syngas for chemical production purposes and high pressure steam for power and heat generation. Technological configuration as in the case of the production of methanol (option III, see p. 3.2.3).
- Energy island: coal gasification system that is integrated with a combined cycle for power production (gas and steam turbines, HRSG - Heat Recovery Steam Generator) and is equipped with syngas conversion and purification systems, as well as with a pre-combustion CO₂ capture system. Configuration of gas treatment system as in the case of the production of hydrogen (option II, see p. 3.2.2) with the difference that after removal of CO₂ the gas is not enriched in hydrogen (no installation PSA) but is heated to about 240 °C and then mixed with nitrogen comes from the air separation unit¹ in order to reduce gas lower heating value (LHV) to 4.7 MJ/m³ (increase of power output of gas turbine as the result of mass flow increase and lowering of gas firing temperature for, i.a., control of NO_x emission).
- CO₂ transport and storage system.

The design of the Polygeneration Plant assumes that the system is coupled with a classic CHP plant (not shown in Fig. 6) consisting of a circulating fluidised bed boiler and steam turbine power generator. A CHP plant uses high temperature steam produced in the chemical island of the Polygeneration Plant for additional power and heat production. The energy production in the form of heat and power covers the needs of local consumers, the town heat distribution network and industrial users. It is assumed that the presented conceptual facility will replace two actual operating heat and power plants. Due to their

¹If the amount of available nitrogen is not sufficient, gas is diluted through the humidification and the third option is steam injection (US DOE & NETL, 2010b)

identical production capacity, the gasifiers working in the system may complete each other, increasing the yearly availability of syngas or power production units based on the assumed production programme.

Specification	option I	option II	option III	option IV
Product	Liquid fuels	Hydrogen	Methanol	Polygeneration Plant
ASU	Cryogenic separation			
Gasification island				
Reactor	Entrained flow, slurry feed; Technology: GEE/Texaco			
Gasification Pressure	3 MPa	5.6 MPa	7 MPa	5.6 MPa
Coal conversion	98 %	98 %	98 %	98 %
Gas cooling	Radiant and convective cooler			
Oxidiser	Oxygen 99.5 % (% vol.)	Oxygen 95 % (% vol.)	Oxygen 95 % (% vol.)	Oxygen 95 % (% vol.)
Fuel	Coal-water slurry; 63% dry solids concentration in the slurry			
Gas treatment				
CO shift	No	Yes, Sour gas shift two-stage CO conversion 97 %	Yes, Sour gas shift one-stage CO conversion ~68 % ^a	Yes, IGCC: see option II Methanol: see option III
Sulfur removal	COS hydrolysis Selexol I stage (99.7 %), ZnO sulfur polishing bed (< 1 ppb)	Selexol I stage (99.7 %)	Selexol I stage (99.7 %) COS hydrolysis (raw gas bypass)	Selexol I stage (99.7 %) COS ^b hydrolysis (raw gas bypass)
Sulfur recovery	Claus, Scot; elemental sulfur			
CO ₂ separation	Selexol II stage MDEA	Selexol II stage	Selexol II stage	Selexol II stage
Liquid fuel	F-T synthesis slurry reactor	-	-	-
Hydrogen	-	PSA, 85 %	-	-
Methanol	-	-	Adiabatic, fixed bed reactor	Adiabatic, fixed bed reactor ^c
Power	Steam turbine excess heat, gas (hydrocarbon recovery unit) combustion,	Steam turbine excess heat, tail gas (PSA) combustion	Steam turbine excess heat, tail gas (methanol synthesis) combustion	Combined cycle IGCC, gas turbine, HRSG, steam turbine

^a as the result of CO Shift and by pass of the raw gas; ^b methanol line; ^c Polygeneration Plant produce syngas with composition enabling its direct use in methanol synthesis.

Table 2. Data on the process configuration of fuel production plants and a Polygeneration Plant (Dreszer & Mikulska, 2009; EPK & IChPW, 2008).

3.4. Transport and storage of CO₂

Separated carbon dioxide is compressed in a multi-stage, intercooled compressor. During the compression, the CO₂ stream is dehydrated with triethylene glycol that is introduced into the compressed stream of CO₂. Dry CO₂ is directed to an intermediate tank and then transported by a pipeline to underground storage units. The condensate from CO₂ drying is directed to a water purification system.

4. Results of process calculations

Coal: For analysis, three hard coals produced in Poland² were selected:

- "Ziemowit" and "Piastr" coal mines: option I
- "Bogdanka" coal mines: option II
- "Janina" coal mines: options II, III and IV

For gasification, the chosen coals have acceptable water and ash contents and sufficient caloric value and ash fusion temperature. We should highlight, however, that the gasification of coals with lower quality parameters, such as high ash content, leads to gasification efficiency decrease and may cause technical problems in the slag feed system. The assessed properties of coal are presented in Table 3.

Lp.	Parameter	Coal mine			
		"Janina"	"Bogdanka"	"Piastr"	"Ziemowit"
Proximate analysis					
1.	W ^{ar} , %	19.1	11.3	13-16	14.8
2.	W ^{ad} , %	8.6	5.5	4-6	7.3
3.	A ^{ad} , %	19.8	21.0	20-25	20.1
4.	V ^{ad} , %	28.4	27.1	30.2	28.5
5.	C ^{adfix} , %	43.2	46.4	43.2	44.0
6.	Q ^{ar} _i , MJ/kg	18.16	21.28	18.0-20.0	19.83
Ultimate analysis, %					
1.	C ^{ad}	54.00	59.45	55.26	56.01
2.	H ^{ad}	4.04	3.47	3.56	3.50
3.	N ^{ad}	0.94	1.26	0.82	0.69
	S ^{ad}	2.00	1.07	0.91	0.93
4.	O ^{ad}	10.62	8.20	14.32	11.40
Ash fusion temperatures, °C					
1.	Initial deformation temp. (IT)	920	900	910	910
2.	Softening temperature (ST)	1,260	1,220	1,250	1,310
3.	Hemispherical temp. (HT)	1,340	1,500	1,360	1,490
4.	Fluid temperature (FT)	1,360	1,500	1,360	1,500

^{ar} as received, ^{ad} air dried

Table 3. Properties of selected coals for analysis of coal gasification for liquid and gaseous fuel production.

² The dominant share - 67% of coal production in the EU27 (Lorenz, 2008).

Availability: a total yearly working time of 85 % has been assumed for all of the options, which is equal to 7,446 hours/year.

Gasifier: process calculations were made for gas generated in the gasifier using GE/Texaco technology. It was assumed that the gasification process would be carried out in a gasification reactor with 125 t/h of raw coal processing capacity. This value meets the processing capacity of operating and newly built entrained flow gasifiers, which are in the range of 100-130 t/h of coal. In typical gasification systems using GE/TEXACO technology, both radiant and convective coolers produce high pressure saturated steam. In the analysed cases, it was proposed that in the radiant cooler, the produced steam is overheated in a convective heat exchanger and then fed directly to a steam turbine for power generation.

Preparation of CO₂ for transport and storage: separated carbon dioxide is compressed to the pressure required for transport conditions, i.e., approximately 12 MPa, and then is transported to storage sites for underground storage.

Process calculation: for the considered technological options, mass balances have been determined on the basis of a calculation made in the ChemCAD v.6.0.2 process simulator for steady state conditions. For liquid fuel production by Fischer-Tropsch synthesis, process calculations were made using data from (US Department of Energy [US DOE], 1999).

4.1. Results of calculations

The summarised results of the process calculations are shown in Table 4.

Parameter	unit	option I	option II		option III
		"Ziemowit"/ "Piast" coal	"Janina" coal	"Bogdanka" coal	"Janina" coal
Coal input	t/h	750	125	125	125
Thermal input	MWth	4,131	631	739	631
F-T liquid production	kg/h	146,200	-	-	-
Methanol	kg/h	-	-	-	62,138
Hydrogen	kg/h	-	10,941	12,197	-
Gross power output	kWe	349,920	73,470	80,040	71,965
Auxiliary load	kWe	366,957	69,204	79,864	72,778
Net power output	kWe	-17,037	4,266	176	-813
Production efficiency	%	N.A.	57.7 hydrogen LHV	54.8 hydrogen LHV	54.6 methanol LHV
CO ₂ sequestration (total)	kg/h	883,660	188,448	220,039	210,462
geological	kg/h	883,660	188,448	220,039	125,022
chemical	kg/h	N.A.	-	-	85,440
CO ₂ capture ^a	kg/h	62 ^b	86	86	96
CO ₂ emission ^c	kg/h	40,800 (56,866)	25,800 (21,777)	29,620 (29,454)	4,143 (4,910)

^a including geological and chemical sequestration, ^b chemical sequestration not included, ^c including the necessary purchase of electricity (943 kg CO₂/MWh) (Finkenrath, 2011)

Table 4. Results of the process calculations (option I-III).

After consuming 750 t/h (5.6 million t/y) of raw coal, a plant produces 146.2 t/h of Fisher-Tropsch synthesis products, including 14.6 t/h of LPG (liquefied petroleum gas), 25.3 t/h of diesel and 106.3 t/h of components for diesel production. In addition, sulfur (6.6 t/h) and carbon dioxide (883.7 t/h) are also produced in the system. The off gas from the F-T processes and the steam generated in the system are used to produce electricity (electric power: 350 MW_e). The electricity produced covers approximately 95 % of the system needs; to balance the power consumption, an additional 127 GWh (17 MW_e) of electric energy is needed.

In the case of hydrogen production, which depends on coal, the plant produces 10.9 and 12.2 t/h of hydrogen from "Janina" and "Bogdanka" coal, respectively. The application of lower quality coal decreased the hydrogen production by approximately 11 %. The gross electricity production also decreased, but due to the growing auxiliary needs in the case of "Bogdanka" coal, which has a greater oxygen demand, a facility using lower quality fuel produces more net energy. In both cases, the electricity production covers the needs of the system. The system also produces sulfur (2.2 and 1.1 t/h) and carbon dioxide (188 and 220 t/h). The efficiency of hydrogen production is 58 % and 55 % (based on LHV) for "Janina" and "Bogdanka" coal, respectively.

A methanol production plant produces 62 t/h of methanol with a high grade purity level. The efficiency of methanol production is approximately 55 % (based on LHV). The energy generated in the system nearly covers the system needs (approximately 99 %). The sulfur production amount is 2.2 t/h. For all of the analysed options, methanol production is characterised by the lowest CO₂ emissions to the atmosphere and, consequently, the highest efficiency CO₂ removal (96 %).

This is because "chemical sequestration" takes place in the methanol production process and part of the CO₂ formed during coal gasification and the conversion of synthesis gas is "stored" in the final product, i.e., methanol.

Case IV involving the Polygeneration Plant is described and analysed in a later section of the paper.

5. Investment expenses

To calculate the investment expenses, an exponential investment assessment method was used based on the following function:

$$C_1 = C_0 \left(\frac{S_1}{S_0} \right)^f \quad (1)$$

where: C_1 is the calculated investment for the system component, C_0 is the reference investment cost, S_1 is the scale of the system component, S_0 is the base scale parameter and f is the scaling exponent.

The base scales and scaling exponents for the components of the production facilities based on coal gasification are shown in Table 5.

Capital expenditures specified for the base year were calculated for the current year using the method of indices according to equation (2):

$$C_2 = C_1 \left(\frac{I_2}{I_1} \right) \quad (2)$$

where: C_2 is the current investment. C_1 is the base investment. I_2 is the current index value and I_1 is the base index value.

The indices used in this study were from the M&S (Marshall & Swift Equipment Cost Index) and CEPCI indices (Chemical Engineering Plant Cost Index) as published in Chemical Engineering. Having assessed the main equipment investments (machines, instruments, devices), the factor analysis has been used by adding relevant coefficients to the coordinates positions and obtaining fixed assets investment estimation results. For total fixed assets investment estimation, the following equation has been used:

$$C_n = E + \sum_{i=1}^m f_i E \quad (3)$$

where: C_n is the fixed assets investment, E is the equipment purchasing costs, and f_i are the coefficients for instruments and devices, fittings, foundations assembly cost, etc.

Plant component	Scaling parameter	Base scale	Exponent
Coal handling	Coal feed	100 t/h	0.67
Gasifier	Coal thermal input	697 MW _{th}	0.67
Oxygen plant – ASU	O ₂ flow	76.6 t/h	0.50
O ₂ compression	Compression power	10 MW _e	0.67
N ₂ compression	Compression power	10 MW _e	0.67
Selexol –H ₂ S removal	Sulfur feed	3.4 t/h	0.67
Selexol –CO ₂ removal	CO ₂ removed	327 Mg/h	0.67
CO ₂ drying and compression	Compression power	13 MW _e	0.67
CO Shift (WGS)	Thermal input	1,377 MW _{th}	0.67
Claus. SCOT	Sulfur feed	3.4 t/h	0.67
Boiler	Heat transfer surface	225, 000 m ²	0.67
Steam turbine	Turbine output	136 MW _e	0.67
Gas turbine	Turbine output	266 MW _e	–
FT synthesis reactor	Thermal output	100 MW _{th}	1.00
FT product upgrading	FT product production	286 m ³ /h	0.7
MeOH synthesis reactor– w/o recirculation	Syngas flow	2.89 kmol/s	0.65
MeOH synthesis reactor – w/ recirculation	Syngas flow	10.81 kmol/s	0.65
MeOH separation and purification	Methanol production	4.66 kg/s	0.29
PSA – hydrogen separation	Hydrogen production	0.294 kmol/s	0.74
CO ₂ removal	CO ₂ flow	3, 280 mol/h	0.60

Table 5. Base scales and scaling exponents for coal conversion system equipment investments.

The investment costs were calculated assuming expenditures presented as "overnight costs" on the basis of the second quarter of the year 2006 and taking into account an investment cost growth of approximately 60 % by mid-2008. To determine the escalation of capital costs a 30 % increase in the cost of engineering services (60 % share in cost increase) and a 100 % increase in steel price³ (40 % share in cost increase) were assumed.

The costs of instruments and devices include the initial equipment plus chemical substances and catalysts. Unpredictable expenses include process costs and project risk.

To calculate investment costs for CO₂ transport and storage 40 km (option I) and 100 km (options II and III) pipelines were assumed.

The investment estimation was conducted with the same accuracy as the pre-feasibility study. i.e., ± 30 %. The investment estimation results are presented in Tables 6 and 7.

Investment component	Thousands \$ (1 \$ =2.2531 PLN; 2008)		
	option I	option II	option III
Instruments and devices supply ^a	1,766,211	390,751	400,737
Instruments and devices assembly ^b	671,160	148,507	152,279
Instrumentation and control equipment	105,973	23,434	24,056
Electric installation	162,491	35,950	36,883
Construction works	264,931	58,630	60,095
Land development	105,973	23,434	24,056
Total direct investments	3,076,740	680,707	698,105
Design and supervision	370,904	82,065	84,151
Total direct and indirect investments	3 447,644	762,771	782,256
Unpredictable expenses	635,836	140,695	144,246
Total investment in Fixed capital	4 083,480	903,466	926,501
Start-up	68,953	15,268	15,623
Total investments	4,152,433	918,734	942,124
Total investments, Thousands \$/TPD (Investments 10 ³ \$/coal input in tonne per day)	230.7	306.2	314.0

^a – includes auxiliary equipment, ^b – includes foundations and piping

Table 6. The investment estimation results for the technological part of the considered plants.

³ Steel Business Briefing Ltd, september 2008

Description	Thousands \$ (1 \$ =2.2531 PLN; 2008)		
	option I	option II	option III
CO ₂ pipeline construction	146,082	113,419	76,932
CO ₂ storing facility	47,601	17,309	11,784
Total	193,683	130,729	88,716

Table 7. Total CO₂ transport and storage related investments.

6. Financial and economic analysis

The base year for financial and economic analysis is assumed to be 2008 (Q4). The analyses have been prepared using fixed prices, without consideration for inflation prognoses or other changes that may constitute factors influencing future prices of the elements involved in the production process. Any prognoses for the coal, gaseous and oil based fuel processing sector bears considerable risk, which convinced us to use actual prices (base year) and keep the relationships between individual assisting factor prices in our analysis. All of the prices used in the calculations are net with VAT excluded. In the calculation, the unit prices were estimated according to the prudence rule for both sales income and for enterprise working cost, which creates a safety margin in terms of possible price fluctuations and other unexpected expenses. At the time of analysis was performed 1 \$ =2.2531 PLN and 1 € = 3.438 PLN. The limit value of the internal rate of return assumed at 6.4 and 8.2 % respectively for the models FCFF (Free Cash Flow to Equity) and FCFA (Free Cash Flow to Firm). The analysis was performed using the UNIDO method (COMFAR III Expert software).

Regarding the foreseen changes in compulsory CO₂ emission allowances starting in 2013, the efficiency calculation is based on three development scenarios:

- basic, assumes project functioning in the present conditions with no regulations on CO₂ (no necessity to buy rights) – hereinafter referred to as scenario 1.
- reference, where a plant owner buys 100 % of the CO₂ emission rights at a price of 39 €/t – hereinafter referred to as scenario 2.
- prospective, assumes the necessity of building CO₂ transport and storage facilities. In this scenario, we include the costs of purchasing and assembling systems for carbon dioxide sequestration, which enable the majority of emitted carbon dioxide to be stored in designated geological structures. For the remaining CO₂ emitted to the atmosphere, there is a requirement to purchase 100 % of the emission rights at a price of 39 €/t – hereinafter referred to as scenario 3.

Assumptions for the calculation are summarised in Table 8 and Table 9 show the adopted total operational costs for the chemicals, the transport and storage of CO₂ and

environmental protection costs (waste disposal, emission fees: NO_x, SO₂, dust, CO₂). The results of the economic analysis are presented in Fig. 7.

The liquid fuel production does not reach the required return rate of the invested capital in the predicted scenarios. The reasons for this situation are the large initial investments for building the plant and production start-up. In case of scenarios which assume the necessity to purchase CO₂ emissions, and especially in the scenario 2 weak financial result is due to the large amounts of CO₂ formed in relation to the manufactured product which is about 6 t/t.

Hydrogen generation enables invested capital return in both analysed cases ("Janina" coal and "Bogdanka" coal); however, considering the possibility of CO₂ emission rights fee implementation, it will be necessary to build additional carbon dioxide transport and storing facilities. Whenever a project lacks these structures, there is no profitability (results – scenario 2).

The methanol production option produced the best results among all of the options analysed for scenarios 2 and 3. This is related to the lower CO₂ amount that is emitted (option II) or designed for sequestration (option III) compared to the hydrogen generation options. It is associated with the "chemical sequestration" i.e. the use of CO₂ for methanol synthesis.

A lack of economic effectiveness in scenario 2 for options I and II and, at a lower rate, for option III with respect to scenario 3 confirms the desirability of CO₂ sequestration (capture, transport and storage), particularly from the perspective of the probability of 100 % emission rights duty after 2012.

The results of the calculations of DPBT (Dynamic Pay Back Time) for the FCFE models allow us to make the following conclusions:

- liquid fuel production does not allow a return on investment expenditures in the assumed lifetime of the installation (30 years).
- for the hydrogen generation project, the discounted pay back times are the following: "Janina" coal: scenario 1 – 9 years from the operation start-up, scenario 3 – 13 years from the operation start-up; "Bogdanka" coal: scenario 1 – 8 years from the operation start-up, scenario 3 – 12 years from the operation start-up.
- methanol generation enables the achievement of financial results that guarantee invested capital return within 9 years from the operation start-up in scenario 1 and 10 years from the start-up in scenario 3.

Project profitability and liquidity assessment

In scenarios 1 and 3, the projects generate positive financial results, which constitute the basis for project stability and for getting the surplus necessary for invested capital return. Scenario 1 assures slightly higher profitability; nevertheless, we may potentially face CO₂ emission rights purchasing after 2012. For option I, the financial performance is insufficient to ensure a return on the invested capital.

Specification	Unit	Cost/ price	Comment
Unitary prices			
Liquefied petroleum gas (LPG)	\$/t	1,556	The basis for the technical propane unit price calculation was its market price, less the excise tax (2008 Polish market).
Diesel	\$/t	936	Fuel oil wholesale price (2008 Polish market).
Component for diesel production	\$/t	749	Price was determined by the fuel oil wholesale price, decreased by 20 % for the value added for its final processing.
Sulfur	\$/t	266	Sulfur prices grew considerably from 2007-2008 from 50 to 500 \$, which made us choose a safe price level considering possible speculative fluctuations. Additionally, price decreases caused by an oversupply in the market are usually small in this product segment.
Hydrogen	\$/t	3,106	Costs of hydrogen production from natural gas (NG) were calculated according to the equation presented in (Stiegel & Ramezan, 2006). The NG suppliers' price parameters have been used in this equation, using the prices for large buyers. As the equation structure primarily considers investment amortisation values, which drastically grew during 2007 and 2008, the results have been increased by 30 % for investment growth compensation. Chemical business specialists were consulted on the calculation methodology and estimated total production cost.
Power	\$/MWh	89	The power sales price, has been accepted as competitive in comparison with prices offered by the CHP plants to the industry, (2008 Polish market).
Methanol	\$/t	596	Average price on the European market for the 2007-2008 period.
Unitary costs			
Coal	\$/GJ	3.99	Market price (2008 Polish market)
Power	\$/MWh	111	See above
CO ₂ emission cost	€/t	39	Related to data published directly by the European Commission (SEC(2008) 85/3)
Water	\$/t	0.11	-
Solid gasification product	\$/t	-	For the prudence rule, the solid product is given away for free, which eliminates the costs of its treatment and disposal.

Table 8. Unitary costs and prices.

Specification	Thousands \$		
	option I	option IIa (IIb)	option III
Chemical substances	2,663	444 (444)	444
CO ₂ pipeline operation cost (scenario 3)	4,674	3,629	2,462
CO ₂ storage operation costs (scenario 3)	2,799	1,156	858
Emission fees (scenarios 1 and 2)	1,062	239 (275)	166
Emission fees (scenario 3)	421	115 (130)	74
Waste disposal	2,219	444 (444)	444

Table 9. Operational costs related to chemical consumption, CO₂ transport and storage and environmental protection.

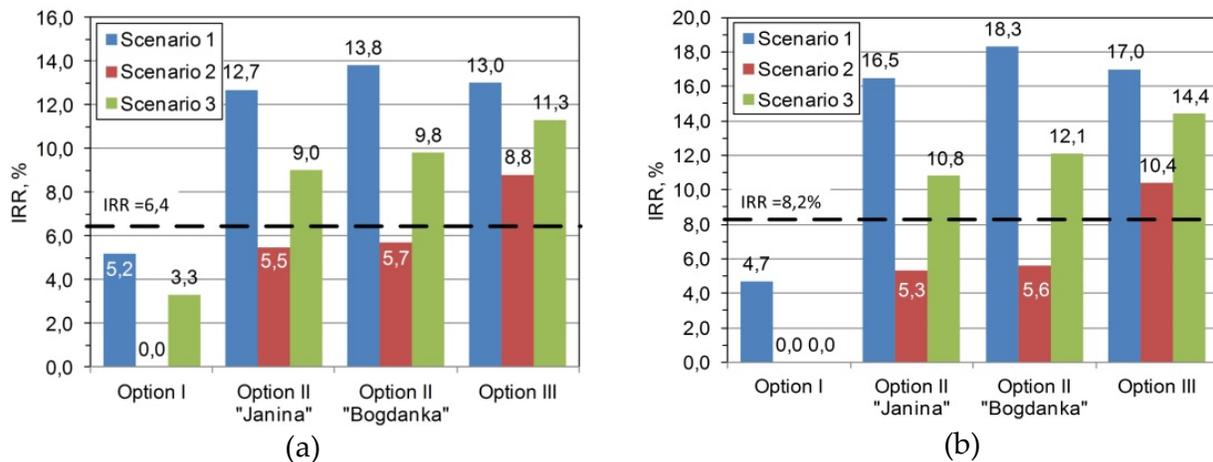


Figure 7. Internal return rate according to FCFF (A) and FCFA (B)

Risk assessment – Project sensibility

The project sensibility has been examined for all options in scenarios 1 and 3 (Fig. 8).

The following parameters have been subject to analysis:

- coal purchase prices: $\pm 10\%$ and their 20% increase.
- investments: $\pm 10\%$ and 20% and 30% growth.
- basic product sales price in all of the options: $\pm 10\%$ and 20%.
- CO₂ emission rights: 10% and 20% growth.

The results of the calculations enable us to formulate the following conclusions:

- coal prices changed in a given area do not implicate large deviations from the calculated efficiency indicators. A basic fuel price increase of 20% does not cause any loss of liquidity in options II and III using both scenarios. For option I, a 15% coal price drop in scenario 1 and a 40% drop in scenario 3 is necessary to obtain a minimum level of profitability,
- an investment level growth of 30% causes a loss of efficiency in option II using scenario 3. Achieving efficiency for option I is related to a necessity to reduce investments by 25% in scenario 3 and by approximately 10% in scenario 1,

- hydrogen sales prices drop by 20% will cause loss of efficiency in scenario 3. For the production of methanol, the lower limit for price level is 23% below the price which was assumed for the calculations. Achieving efficiency measures for option I is related to a necessity to raise sales prices by 15% and 27% respectively for scenario 1 and 3.
- thanks to a CO₂ transport and storage system, the project is not excessively price sensitive in terms of emission rights purchasing in scenario 3. Even with 20 % growth, the project efficiency is preserved. Option III is characterised by the smallest fluctuation and lowest carbon dioxide emission indicator.

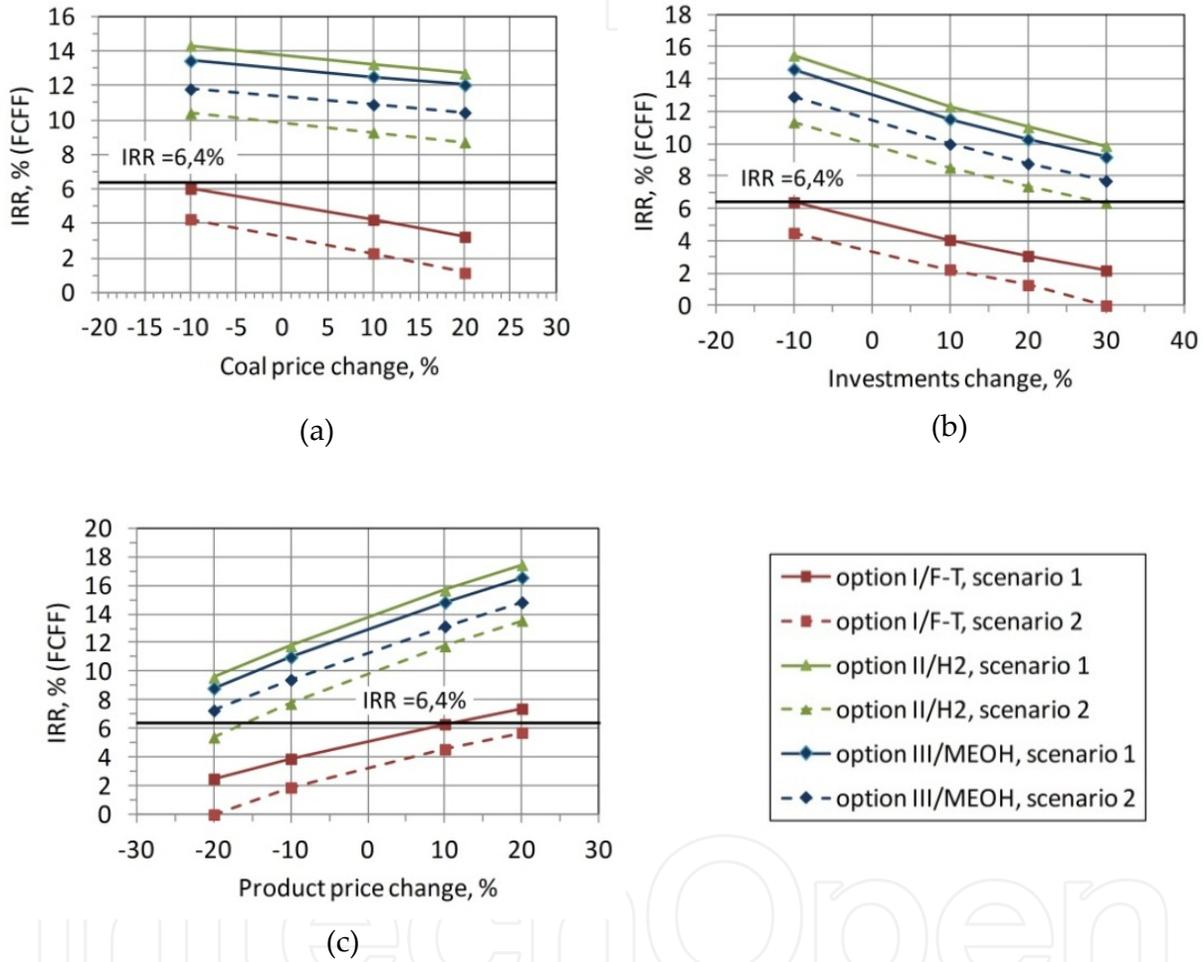


Figure 8. Sensitivity analysis A) coal price, B) investments, C) product price.

Additionally, for the defined scenarios using the considered options, the basic product minimum prices have been determined to assure profitability limit achievement. i.e., IRR equal 6.4 % (Table 10.). The prices of the analysed gasification products have been referred to the oil and natural gas prices in the following manner:

- option I – a motor oil semi-finished component is the basic product of the system. The unit price of the semi-finished motor oil component that was used in this analysis was the motor oil wholesale price decreased by 20 % (see Table 8). We have assumed that the motor oil semi-finished component will be equivalent to crude oil.

- option II – Hydrogen is the basic product of the system. The basis for the hydrogen generation cost calculation was the price of natural gas. The costs of hydrogen production from natural gas (NG) were calculated according to the equation presented in (Stiegel & Ramezan, 2006).
- option III – Methanol is the basic product of the system has been compared to the equivalent natural gas prices on the basis of available projects and consultations with Polish Chemical Industry Chamber experts. We should, however, highlight that methanol prices in the market are subject to considerable fluctuations, which are not always caused by natural gas prices changes. The calculations above may be burdened with methodological error that is difficult to define.

The results of calculations show minimum oil and natural gas prices, which assure the profitability of products included in individual options and according to assumed scenarios. The methanol production project has the best relationship in this matter as hydrogen production marketability is more dependent on natural gas prices. For coal-based liquid fuel production (motor oil semi-finished component) to be attractive with the different scenarios considered, oil prices must exceed 87 \$/bbl.

Option	unit	Scenario 1	Scenario 2	Scenario 3
Liquid fuel production	\$/t	832	1,338	948
	\$/bbl oil equivalent	87	140	99
Hydrogen production	\$/t	2,173 (2,090)	3,220 (3,192)	2,699 (2,617)
	\$/1000 m ³ NG equivalent	364 (350)	553 (549)	459 (444)
Methanol production	\$/t	418	539	455
	\$/1000 m ³ NG equivalent	373	481	406

Table 10. Minimum selling prices of manufactured products assuming minimum profitability (IRR = 6.4 %).

7. Polygeneration plant

Polygeneration systems mediate the simultaneous production of chemicals and electricity from syngas. The purpose of these systems is to make maximum use of the chemical energy of coal by maximising the total energy efficiency of the transformation of primary fuel into useful products while minimising the capital expenditure and operating costs. Syngas may be used independently to produce chemicals and electricity, most advantageously in IGCC (integrated gasification combined cycle) systems.

Polygeneration usually include electricity production that is integrated with the generation of hydrogen, methanol or the products of Fischer-Tropsch synthesis. The principal advantages of a polygeneration system include:

- increased economic flexibility (two or more products);
- lower production costs due to more efficient use of syngas and of the technological heat produced in the course of the production process.

The integration of the processes of power and chemical production in a polygeneration system allows the achievement of high rates of fuel conversion, low emission rates and high economic efficiency, as in the case of CO₂ sequestration.

Table 11 presents the basic process data of the considered system. Consuming about 257 t/h of coal, the production of syngas amount to 85.1 t/h. This is enough to obtain 63.4 t/h of methanol. Net power and heat (in form of HP steam) output is 142 MW_e and 130 MW_{th} respectively. Geological sequestration of CO₂ will be 311 t/h. The amount of CO₂ stored in the chemical end product (methanol) will be 87 t/h.

Tables 12 and 13 show the investment costs and the minimal energy and synthesis gas prices to ensure the viability of a project (NPV >= 0 and IRR >= 7 %. with an amortisation period of 20 years). The calculation results are presented separately for the main technological units and for the whole facility of the Polygeneration Plant for the base case (CO₂ emissions within the scope of given emission rights) and in the case of CO₂ sequestration (separation; transport and storage of CO₂; fees for the remaining CO₂ emissions 39 €/t).

Fig. 9 shows the results of the calculations of produced synthesis gas prices against electricity prices (NPV >= 0 and IRR >= 7 %. with an amortisation period of 20 years) and the area of the economic attractiveness of the project.

Parameter	unit	Option IV Poygeneration Plant "Janina" coal
Coal input	t/h	257
Thermal input	MW _{th}	1,296
Syngas production	kg/h	85,079
equivalent methanol production	kg/h	63,400
Gross power output	kW _e	282,700
Auxiliary load	kW _e	140,591
Net power output	kW _e	142,109
Thermal output ^a	kW _{th}	130,000
Production efficiency (mixed)	%	57.6 ^b
Syngas production efficiency	%	73.2
Power production efficiency (IGCC)	%	31.4
CO ₂ sequestration	kg/h	397,811
geological	kg/h	310,636
chemical	kg/h	87,175
CO ₂ capture ^c	kg/h	88
CO ₂ emission	kg/h	38,802

^a high pressure steam from chemical module (see Fig. 6), ^b including syngas (chemical enthalpy), heat (HP steam) and power production, ^c including geological and chemical sequestration

Table 11. Results of process calculations for option IV.

The calculation results clearly indicate the attractiveness of the polygeneration process. The combination of electricity generation and synthesis gas production for the presented technological configuration (use of gasification technologies for energy production and syngas) causes a significant reduction in the minimum price of energy in comparison to the IGCC system (production of electricity) to 49 and 21 \$/MWh without and with CO₂ sequestration, respectively (Tables 12 and 13).

Specification	Unit	IGCC (power island)	Syngas production unit (chemical island)	Poligeneration Plant
Investments	mIn \$ 10 ³ \$/TPD ^a	1,105 358.3	670 217.3	1,776 287.9
Price limits:				
Power	\$/MWh	131	111 ^b	82
Syngas	\$/1000 m ³		144	202 ^c

^a Investments 10³ \$/coal input in tonne per day, ^b the approved purchase price of electricity reflects the price level of december 2008, ^c Adopted the maximum price of synthesis gas (Q4 2008), considered to be commercially attractive (the price of the synthesis gas produced from natural gas).

Table 12. Investments and price limits for manufactured products (power and syngas); Polygeneration Plant without CCS.

Specification	Unit	IGCC (power island)	Syngas production unit (chemical island)	Poligeneration Plant
Investments:	mIn \$ 10 ³ \$/TPD ^a	1,256 368	804 221.5	2,060 294.8
Including:				
CO ₂ Transport and Storage	mIn \$	121	121	242
Price limits:				
Power	\$/MWh	191	111 ^b	170
Syngas	\$/1000 m ³		167	202 ^c

^a Investments 10³ \$/coal input in tonne per day - technological part only without CO₂ Transport and Storage, ^b and ^c see table 13.

Table 13. Investments and price limits for manufactured products (power and syngas) Case: CO₂ sequestration.

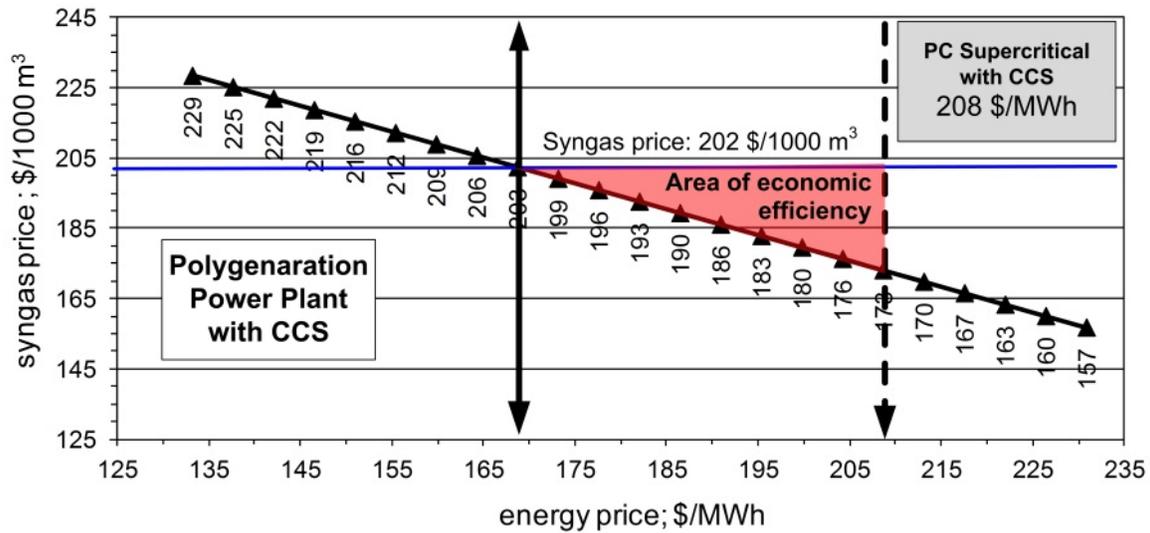


Figure 9. Estimated cost of synthesis gas in relation to the price of electricity and the area of economic efficiency of the Polygeneration Plant.

The CO₂ sequestration benefits of the proposed solution are also visible when comparing the Polygeneration Plant with a Supercritical Power Plant based on coal combustion. A comparison of the energy price limits for both cases at the same production level shows that with polygeneration we obtain lower energy prices by 38 \$/MWh (energy price forecast for the supercritical coal unit with CCS amounts to 208 \$/MWh). This underlines the attractiveness of the presented solution and the need to develop the proposed concept under appropriate technological conditions with the existence of a recipient for the produced synthesis gas as an alternative to traditional solutions.

8. Conclusion

The analysis concerned the installations for gaseous and liquid fuel production based on coal gasification using commercially available technologies of coal gasification, gas cleaning and conversion and chemical synthesis.

Systems for liquid fuels, hydrogen and methanol production were analysed in detail assuming three scenarios: basic (with no necessity to buy rights for CO₂ emission), reference (purchase 100 % of CO₂ emission rights at a price of 39 €/t), and prospective (assuming construction of CO₂ transport and storage facilities).

The analysis of the examined cases shows that with the adopted assumptions, the most favourable option is definitely the production of methanol, which shows economic effectiveness in all of the scenarios and, in the case of scenarios 2 and 3, gives the best results among the options analysed. The reason for this superiority among other options is related to low CO₂ emission, associated with the “chemical sequestration” i.e. the use of CO₂ for methanol synthesis.

The economic attractiveness of the production of hydrogen is significantly more dependent on natural gas prices. Hydrogen production is economically feasible only in scenarios 1 (base) and 3 (prospective). Developments in this direction and, consequently, the hydrogen economy seem to be limited due to a lack of cost-effective storage technology and transport infrastructure. At present, hydrogen from coal can effectively be used in chemical plants for the production of ammonia and fertiliser by substitution of the hydrogen produced from natural gas.

The coal to liquid fuels process based on Fischer-Tropsch synthesis is attractive only when exceed 87, 140 and 100 \$/bbl for scenarios 1, 2 and 3, respectively.

Among the analysed technological options, the production of liquid fuels from coal using FT synthesis is definitely the least attractive and, on the basis of the obtained results, is not recommended as a potential direction for the application of coal gasification technology.

However, the idea of the production of liquid fuels from coal is still attractive, and the production of liquid fuels from coal using methanol seems to be a reasonable option. Methanol is used directly as motor fuel or is added to liquid motor fuels to improve their operational performance (methyl tertiary butyl ether, MTBE). Moreover, technologies for the production of motor fuels from methanol (MTG - methanol to gasoline and MTO/MOGLD – methanol to olefines/Mobile olefines to gasoline and destilate) are being intensively developed and are commercially available at the industrial scale.

A lack of economic effectiveness in scenario 2 for options I and II and, at lower rates, for option III with respect to scenario 3, confirm the desirability of CO₂ sequestration (capture, transport and storage), particularly from the perspective of the necessity to purchase CO₂ emission rights after the year 2012.

The analysis of the Polygeneration Plant clearly shows the attractiveness of the solutions and the need to develop the proposed concept in appropriate technological conditions with the existence of a recipient for the synthesis gas produced as an alternative to traditional solutions. The realisation of this production process would give the possibility of significant reductions in the price of electricity generated, even in the case of CO₂ sequestration, compared to traditional technologies, including IGCC, while maintaining cost-effective production of synthesis gas for chemical applications. Also important from the economic point of view is installation flexibility in terms of the final product. i.e., the ability to design a production profile according to market demand for the manufactured products.

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