Techno-Economic Analysis of Direct Coal-Biomass to Liquids (CBTL) Plants with Shale Gas Utilization and CO₂ Capture and Storage (CCS)

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Abstract

In this paper, techno-economic analysis of direct coal biomass to liquids (CBTL) plants is performed using Aspen Process Economic Analyzer (APEA) using high fidelity process models developed in Aspen Plus for four different configurations of direct CBTL plants. Results from the economic model are validated with the data in the open literature, if available. Sensitivity studies are conducted to evaluate the impacts of key investment parameters, design parameters, and potential government-subsidized credits on the main economic measures including net present value (NPV), internal rate of return (IRR), break-even oil price (BEOP) and equivalent oil price (EOP). Using the North America 2015 pricing basis in APEA, this study shows that the BEOP of direct CBTL processes ranges from \$56.9/bbl to \$80.5/bbl for large scale (50k bbl/day) plants and from \$77.3/bbl to \$97.5/bbl for small scale (10k bbl/day) plants. It is observed that integrating a carbon capture and storage (CCS) unit to the direct CBTL process can increase the BEOP by about 10%, while utilization of the cheap and abundant shale gas (especially in the continental US) can make the direct liquefaction processes considerably more attractive than the indirect CBTL processes.

Key Words

Coal, Biomass, Shale gas, Direct liquefaction, CO₂ capture and storage, Techno-economic analysis

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1. Introduction

Direct coal to liquids (DCL) process is technically feasible for producing alternative transportation fuels from coal, which has larger inventory than petroleum crude. With appropriate product upgrading, synthetic fuels produced from the DCL process can be directly used in the current motor engines. However, similar to other coal liquefaction processes, the DCL process has not been widely commercialized mainly because of economic uncertainty, high CO₂ emission, and high capital cost [1,2]. It is reported that the capital investment in and CO₂ emission from the Shenhua DCL plant with coal-derived hydrogen and a capacity of 16,300 bbl/day are about \$1.46 billion (reported in 2008) and 0.48 tonne CO₂ per barrel liquids [1-5]. The environmental footprint of coal liquefaction processes can usually be reduced by co-processing coal and biomass and applying CO₂ capture and storage (CCS) technology but at the cost of higher capital and operating investments [3,6-10]. Furthermore, instead of coal gasification, hydrogen can be produced from less expensive and more H₂-rich sources. For example, the recent boom of shale gas in the United States has opened a door for leveraging the shale gas as a source of H₂ [1,11,12]. Therefore, there is a need to reevaluate the potential of the DCL process. There is barely any study in the current open literature on the DCL technology at the systems level with shale gas and biomass utilization integrated with CO₂ capture and compression. Most published studies have focused on the indirect coal liquefaction (ICL) technology, mainly because the ICL technology is more mature and has adequate industrial experience, including several coal or gas to liquids plants recently constructed by Shell Qatar, Sasol, ExxonMobil and Yitai. [13-15] Due to the lack of commercial experience and literature in the area of process synthesis and systematic analysis of direct coal-biomass to liquids (CBTL) processes with or without shale gas utilization and CCS, development of plant-wide process and economic models will be useful for analyzing the feasibility of the modified DCL technology and for conducting further studies, such as optimization and life-cycle assessment.

In the DCL process, coal is directly converted to liquid fuels by adding hydrogen, as shown in Reaction (1) [1,2]. Because of low H/C ratio in the coal, even though very less CO₂ is generated in the direct liquefaction reactor, a significant amount of external hydrogen is required, generation of which leads to high amount of CO₂ release [1,3]. This hydrogen is needed not only to convert coal to partially refined syncrude in the liquefaction reactor, but also to bring the H/C ratio up to about 2 in the syncrude upgrading step. In the Shenhua DCL plants, the hydrogen demand is satisfied by coal gasification route with large capital investment and high level CO₂ emission [1,2,5,16]. Other than coal gasification, hydrogen can also be produced by partial oxidizing liquefaction residues or a mixture of residues, coal and biomass, and reforming natural gas or shale gas [1,17]. Residues partial oxidization (POX) can reduce the demand of external fuels for hydrogen production, while steam methane reforming (SMR) can

produce hydrogen with less CO₂ emission and from cheaper energy sources. According to our previous studies, even though both POX and SMR technologies are considered for the DCL process, the CO₂ emission could only be reduced by about 20%, which is still significant compared with traditionally petroleum industries [1].

Similar to the studies conducted to the ICL processes [3,6-10], further reduction of CO₂ emission can be achieved by adding a small amount of biomass into the feedstock and applying CCS technology to capture the CO₂ produced in plant without substantially modifying current DCL technologies. It is widely accepted that biomass is a carbon-neutral energy source [18,19]. In addition, utilization of biomass in the DCL process can reduce the external hydrogen demand in the liquefaction reactors because of its higher H/C ratio, and therefore reduce the CO₂ emission from the hydrogen production plant [1,20]. Several experimental studies have been conducted for co-processing coal and biomass using direct liquefaction processes [10,20,21], but those processes have been barely modeled at either equipment-level or system-level. CCS technologies have been widely studied for the ICL [22], integrated gasification combined cycle (IGCC) [23], pulverized coal combustion and natural gas combined cycle (NGCC) processes [24], but there is still lack of such studies for the DCL processes.

In our previous study, detailed plant-wide models have been developed in Aspen Plus for direct CBTL plants of different configurations by considering different sources for hydrogen and different extent of CCS [1]. In that study, the focus was on conversion efficiency and CO₂ emission but not on economic performance. To analyze the commercial feasibility of those novel processes, techno-economic studies are required in addition to the technical analysis. It has been reported by several researchers that DCL processes may have better economic performance than ICL processes due to their higher thermal efficiency [1,3,25], while, as per Robinson et al., the economic performance of DCL and ICL processes are similar [4]. However, there is hardly any techno-economic study of the DCL technology conducted by using rigorous process and economic models especially while considering CO₂ capture, biomass co-processing and different H₂ sources. Most of the techno-economic studies in the open literature have been conducted for the ICL processes, IGCC plants and coal-firing power plants rather than DCL processes [6,7,18,24,26-29]. Due to the difference in the conversion mechanisms, CO₂ emission sources, and process configurations, the impact and penalty of adding biomass and CCS are expected to be different between ICL and DCL technologies. Therefore a rigorous techno-economic study of the direct liquefaction processes with CCS and different H₂ sources is very much desired.

As mentioned before, the carbon footprint of the energy conversion processes can be reduced by adding biomass and applying CCS technologies at the cost of higher operating cost and capital investment [3,6-10]. In order to promote biomass utilization and development and commercialization of CCS technologies, government subsidies, such as tax benefits, carbon tax and other environmental credits, are being offered in a number of countries or in various regions within a country [30]. For example, the federal government in the USA provides residential renewable energy tax credit to household using solar, wind, geothermal, or some other renewable energy sources [31]. In addition, the price of electricity produced in a power plant with one of the Renewable Energy Certificates can be about \$2/MWh to \$15/MWh higher than the average market prices [32]. Carbon credit or carbon tax is also being strongly suggested as an incentive to reduce CO₂ emission. A carbon tax is a specific price the government charges for carbon content in fuels [33,34]. The captured CO₂ can be traded like a product in the carbon-constrained market subject to the carbon tax. In the future if the carbon tax is imposed or incentives for use of renewables such as biomass are provided, profitability of the CBTL plants with CCS can improve. Therefore, it is desired to analyze the impact of these incentives on the profitability of the CBTL plants.

To summarize, the existing literature lacks rigorous techno-economic analysis of the DCL technology that can greatly benefit from biomass utilization, development of novel CCS technologies, and utilization of the cheap and abundant shale gas (especially in the continental US), as well as potential environmental credits. With these motivations, a techno-economic study is conducted using multiple-software, including Aspen Plus, APEA, Exchanger Design and Rating (EDR), and MS Excel with the 2015 pricing basis for North America. Detailed steady-state modeling approach in Aspen Plus can be found in our previous publication [1]. Contributions of this paper can be summarized as follows:

(1) development of economic models for the direct CBTL plants in APEA based on the process models developed in Aspen Plus, (2) sensitivity studies to analyze the impact of key design parameters (i.e. biomass/coal ratio, hydrogen sources, extent of CCS) and investment parameters (i.e. prices of coal, biomass and shale gas, project life, plant contingency and plant capacity) on key economic measures, such as net present value (NPV), internal rate of return (IRR), break-even oil price (BEOP) and equivalent oil price (EOP), (3) evaluation of potential environmental credits (i.e. carbon credit, renewable energy certification, government-subsidized tax credit) due to use of biomass and CCS, (4) comparison between indirect and direct liquefaction processes.

2. Steady-state simulation

In this study, four different plant configurations are considered for direct CBTL processes, SMR_CCS, SMR_VT, CG_CCS and CG_VT, with the block flow diagram (BFD) presented in Fig. 1. Two sources of

hydrogen are evaluated, namely steam methane reforming (SMR) or co-gasification (CG). The CO₂ can be directly vented (VT) or compressed for sequestration (CCS). In Fig. 1, the solid lines denote the common process sections for all configurations, while the dash lines denoted the process sections varying with different configurations. The liquefaction and hydrocarbon recovery section is designed similar to the catalytic two-stage liquefaction (CTSL) technology from Hydrocarbon Technologies Inc. (HTI) [2,35-37]. Inline hydrotreating approach is introduced to reduce the energy penalty of raw syncrude stabilization step. In addition to vacuum distillation, Residual Oil Supercritical Extraction-Solids Rejection (ROSE-SR) technology is used to increase the efficiency of and liquid recovery in the solid/liquid separation process as part of the hydrocarbon recovery unit [38-40]. Isomerization, catalytic reforming, and gas oil hydrotreating are applied to upgrade syncrude to on-spec gasoline and diesel [41]. Heating duties of the furnaces are provided by utilizing the light gas oil and fuel gas produced inside the DCL plant. Remaining gas from the fuel gas header is sent to the power island for generating electricity and utility steams. Required hydrogen is produced from the syngas via water gas shift (WGS) reaction [23]. In the CG_CCS and CG_VT processes, syngas is produced by gasifying a mixture of coal and biomass from the pre-processing units and liquefaction residues from the ROSE-SR unit. In the SMR_CCS and SMR_VT processes, syngas is produced by a combination of liquefaction residues POX and shale gas SMR. The syngas from the WGS reactors is sent to the acid gas removal (AGR) unit and pressure swing adsorption (PSA) unit to produce high purity hydrogen. In the SMR_VT and CG_VT processes, captured CO₂ is vented to the atmosphere. In the SMR_CCS and CG_CCS processes considering high extent of CCS, additional CO₂ is captured from the flue gas and all captured CO₂ is sent to the CO₂ compressor for sequestration. The dual-stage Selexol technology is selected to remove CO₂ and H₂S available at high pressure from the POX/CG unit while the MDEA/PZ technology is selected to remove CO₂ available at intermediate or low pressures from the SMR and combustion units. [1]

In this study, four configurations of the direct CBTL technology, namely SMR_CCS, SMR_VT, CG_CCS and CG_VT, presented in our earlier paper are evaluated [1]. Illinois No.6 coal, wood chip, and Marcellus shale gas are used as feedstocks [1]. Most of the equipment items are simulated using standard library models available in the Aspen Plus library. For the reactors in liquefaction and product upgrading sections, yield models are developed in Excel and validated using the experimental data available in the open literature. These Excel models are integrated into the main flowsheet in Aspen Plus using User2 blocks. Table 1 summarizes the simulation approach, operating conditions, and property models of the key equipment items. Plant configuration, modeling approach and composition of products and feedstocks can be found in our previous publication. [1]

Fig. 1. General BFD of direct CBTL processes.

Table 1

Summery of the process model of direct CBTL plants.

Section/Block	Simulation Approach	Property Model/Operating Conditions	Reference
Liquefaction and hydrocarbon recovery		Peng-Robinson	
Liquefaction	Close-coupled yield model for two ebullated-bed reactors in series	1 st stage: 407 °C, 22.1 MPa 2 nd stage: 432 °C, 20.7 MPa	[1,37,38,42]
Inline hydrotreating	Yield model	370 °C, 17.2 MPa	[1,7,37]
Distillation columns	PetroFrac	Atmospheric column: 2.8 bar Vacuum column: 0.1 bar	[1,7]
ROSE-SR	Component separator for settlers and flash vessel for deashing solvent separator	Solvent/solid weight ratio: 3 1 st stage settler: 300 °C, 55 bar 2 nd stage settler: 370 °C, 54.5 bar	[1,43]
Product upgrading		Peng-Robinson	
Gas oil hydrotreating	Same as inline hydrotreater	350 °C, 180 bar, LHSV: 1 h ⁻¹	[1,7,41,44]
Isomerization	Yield model	Hydrogen/oil: 0.14 wt% Targeted RON: 83	[1,7,41,45]
Catalytic reforming	Yield model	Targeted RON: 95	[1,7,41,46]
Syngas Production		Peng-Robinson/ELECNRTL-RK	
Pre-reformer	RGibbs model	Adiabatic; Inlet: 510 °C, 27 bar	[1,47]
Steam reformer	RGibbs for reformer and Rstoic with combustion reactions for furnace	Reformer: 815 °C, 25 bar Reformer furnace: 955 °C	[1,47]
Gasification	RGibbs model	1315 °C, 56 bar	[6,17,23,48]
Water gas shift	Plug flow reactor	CO conversion: 95%	[6,23]
Acid gas removal and hydrogen recovery		ELECNRTL/PC-SAFT	
Chemical absorption	RadFrac model with rate-based stages and reaction kinetics	Absorber feed: 38 °C Regenerator: 1.7 bar	[1,5,6]
Physical absorption	RadFrac model with equilibrium stages	Solvent chilling: 2 °C	[1,6,23]

Hydrogen recovery	Polybed PSA process modeled as component separator	Adsorption: 26.2 bar Desorption: 6.9 bar	[1,6,45]
CO ₂ compression	Multistage compressor	15.3 MPa for CO ₂ pipeline	[1,6,23]
Power island		Ideal/IAPWS-95	
Combined cycle	Stage-by-stage estimation of steam turbine and Aspen Plus standard models for others	Triple-pressure HRSG with reheat: 114/25/4 bar	[1,7,23]

3. Economic analysis

APEA V8.4 is used to perform economic analysis of the direct CBTL plants. It is noted that all results presented in this work are based on the 2015 pricing basis for North America. Fig. 2 summarizes the procedure that is followed for techno-economic analysis in this study. Stream information, such as temperature, pressure and flowrate, as well as the basic equipment type is automatically specified by directly ‘exporting’ the plant-wide models developed in Aspen Plus to APEA. In APEA, the capital investment, denoted as the total project cost (TPC), can be estimated by mapping the equipment items from the Aspen Plus flowsheet to corresponding APEA project component(s), if available. These equipment items are sized using ASTM standards or other correlations available in APEA. [26] Vendor cost obtained from the open literature is used for the equipment items for which there are no suitable APEA project component and also for those for which yield models were used in Aspen Plus [26]. Section 3.1 provides a more detailed description of project component mapping, sizing and cost estimation. The main source of operating and maintenance (O&M) cost is from utilities and raw materials. Other O&M costs can be found in Section 3.2. In addition, investment parameters are specified. In APEA, economic analysis and sensitivity studies can be conducted by using the Decision Analyzer tool. If plant configuration and/or any key process design parameters listed in Table 7 changes, a new process model is developed in Aspen Plus and then ‘exported’ to APEA for economic analysis [26].

Fig. 2. Procedure for economic analysis in multi-software environment.

Table 2 lists the prices of raw materials, labor and products in 2015 basis. The prices of raw material and products are mainly obtained from the US Energy Information Administration (EIA) website. The crude oil price (COP) is the refiner acquisition cost of crude oil in the PADD1 area (the east coast of US). In this study, NPV and IRR are calculated assuming the wholesale prices of gasoline and diesel are COP plus the refinery margin, \$0.333/gallon for gasoline and \$0.371/gallon for diesel [26]. Break-even oil price (BEOP) is the COP making the process NPV zero, while equivalent oil price (EOP) is defined as the COP making the process IRR be 12%. The carbon credit is defined as carbon in the additional CO₂ captured by the CCS facilities compared with the petroleum baseline. In the PADD1 area, the CO₂ emission from the petroleum refineries is about 45 kg CO₂/bbl crude oil, which is equivalent to about 8.12 kg CO₂/GJ fuel [49]. It is assumed that if the CBTL facility is located in a place that is subject to carbon tax and if CO₂ emission of the CBTL plant with CCS is lesser than 8.12 kg CO₂/GJ fuel, then the additional CO₂ that is captured and sequestered can be leveraged to improve the plant economics. In the base case, the price of carbon credit is set to be zero as carbon tax is still fairly uncommon in most locations around the world. Table 3 lists the investment parameters for the base case scenario. Here, process contingency is set to be 24% because of the novelty of the direct CBTL plants. The length of start-up period is set to be 40 days because of the process complicity.

Table 2

Prices of raw material, labor, and product (base case).

Cost (\$/unit)		Cost (\$/unit)	
Coal (\$/tonne)	34.0	Supervisor (\$/hr)	80
Wood chip (\$/dry tonne)	61.5	Crude oil price (\$/bbl)	60
Shale gas (\$/GJ)	2.25	Electricity (\$/MWh)	50
Operator (\$/hr)	50	Carbon credit (\$/tonne)	0

Table 3

Investment parameters (base case).

Parameter	Value	Parameter	Value
Start date of engineering	2015	Utility escalation (%/year)	1
Contingency percent	24%	Working capital percentage (%/FCI)	12
Number of years for analysis	30	Operating charges (% of labor costs)	25
Tax rate	40%	Plant overhead	50%
Interest rate/desired rate of return	10%	General & administrative expenses	8%
Project capital escalation (%/year)	1	Length of start-up period (weeks)	40
Products escalation (%/year)	1	Operating hours per period	8000
Raw material escalation (%/year)	1	Construction time	2.5 yr

3.1. Estimation of the total capital cost (TPC)

In this study, all heat exchangers are designed in Aspen EDR; all columns are designed in Aspen Plus using available options for tray/packing sizing; other equipment items such as vessels, pumps and compressors are sized in APEA. The materials of the construction (MOC) for all project components are selected based on the operating temperature, composition of the service stream (i.e. H₂, H₂S partial pressure), and common industry practice [5,37,50-52]. For all pumps and reciprocating compressors, spares are considered. The equipment items and sections for which costs cannot be estimated by the Icarus database especially the reactors and product upgrading units, are mapped as ‘quoted’ equipment item in APEA[®] with cost correlations shown in Eqs. (1) and (2). In those correlations, *DIP* is the direct permanent investment including ISBL and OSBL cost; *C₀* is the base cost; *S₀* is the base capacity; *S* is the actual capacity, *sf* is the scaling factor; and *n* is the number of trains. Eq. (1) is used for the ‘quoted’ equipment with missing information, where the OSBL cost is estimated by considering the balance of plant (BOP) percentage. Eq. (2) is applied for other ‘quoted’ equipment items. Parameters in the cost correlations with 2015 pricing basis are obtained from the open literature or derived using the data available in the open literature, as shown in Table 4. Various steps for capital cost estimation in multiple-software environment are described here only for the gas oil hydrotreating unit as an example. The configuration of the gas oil hydrotreating unit is shown in Fig. 5 of our previous publication [1], while Table 5 lists the corresponding models in APEA, required numbers of items and spares, if any, sizing approach and MOC for each equipment. Complete equipment lists and detailed specifications for all standard equipment items can be found in Appendix A of this paper.

$$DIP = (1 + BOP)C_0(S/S_0)^{sf}n^{0.9} \quad (1)$$

$$ISBL \text{ cost} = C_0(S/S_0)^{sf}n^{0.9} \quad (2)$$

Table 4

Parameters for Eqs. (1) and (2) for quoted equipment items (2015 pricing basis).

Equipment	<i>C₀</i> (MM\$) ⁽¹⁾	<i>S₀</i>	<i>S_{max}</i>	<i>S₀</i> basis	Units	<i>sf</i>	BOP	Reference
Gasifier	137.09	2464	2616	dry feed	tonne/day	0.67	added	[27]
WGS reactor	3.16	2556	2600	output	tonne/day	0.65	no	[27]
Isomerization	1.00	13.06	2720	feed	tonne/day	0.62	no	[53]
Catalytic reforming	5.39	36.99	8160	feed	tonne/day	0.6	no	[53]
Air separation unit	57.90	1839	2500	O ₂	tonne/day	0.5	added	[27]
Coal pre-processing	57.50	2464	2616	dry feed	tonne/day	0.67	added	[27]
Biomass pre-processing	27.98	2000		dry feed	tonne/day	0.67	added	[27]
CO ₂ compressor	31.81	11256		CO ₂	tonne/day	0.75	no	[24]
PSA H ₂ recovery	0.84	944		H ₂	Nm ³ /h	0.55	no	[53]

Claus unit	24.23	125	S	tonne/day	0.67	no	[27]	
Steam methane reformer	62.10	26.1	35	feed	kg/s	0.67	no	[54]
Shale gas pre reformer	12.30	26.10		feed	kg/s	0.67	no	[55]
ROSE-SR unit	66.70	50800		feed	bbl/day	0.67	no	[52]
Liquefaction reactor	94.79	587.79		feed	tonne/hr	0.67	no	[52]

(1) The costs of quoted equipment are escalated using the Chemical Engineering Plant Cost Index (CEPCI).

Table 5

Detailed component specifications for the gas oil hydrotreating unit in APEA.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC ⁽¹⁾
Reactors & vessels				
gas oil hydrotreater	2/0	VT MULTI WALL	APEA	A387F (SS347)
Hot high pressure flash	1/0	VT CYLINDER	APEA	A387D
Cold high pressure flash	1/0	VT CYLINDER	APEA	A516
Low pressure flash	1/0	VT CYLINDER	APEA	A516
Stabilizer condenser drum	1/0	HT HORIZ DRUM	APEA	A516
Main distillation condenser drum	1/0	HT HORIZ DRUM	APEA	A516
Distillation columns				
Main distillation tower	1/0	TW TRAYED	Aspen Plus	A516/A285C
Stabilizer tower	1/0	TW TRAYED	Aspen Plus	A516/A258C
Compressors, pumps & turbines				
Main distillation reflux pump	1/1	CP CENTRIF	APEA	CS casing
Stabilizer reflux pump	1/1	CP CENTRIF	APEA	CS casing
Makeup H ₂ compressor	1/1	GC RECIP MOTR	APEA	SS casing
Recycle H ₂ compressor	1/0	GC CENTRIF	APEA	SS316 casing
Stabilizer feed pump	1/1	CP CENTRIF	APEA	CS casing
Gas oil feed pump	1/1	CP CENTRIF	APEA	CS casing
Furnaces, boiler & heat exchangers				
Main distillation reboiler	1/0	RB U TUBE	EDR	A213C/A387B
Main distillation condenser	1/0	HE FIXED T S	EDR	A214/A516
Main distillation pumparound	2/0	HE FLOAT HEAD	EDR	A214/A516
Stabilizer reboiler	1/0	RB U TUBE	EDR	A214/A516
Stabilizer condenser	1/0	HE FIXED T S	EDR	A214/A516
Gas pre-heater	1/0	HE FLOAT HEAD	EDR	A214/A516
Liquid pre-heater	1/0	HE FLOAT HEAD	EDR	A213D/A387D
Feed furnace	1/0	FU BOX	APEA	347S
LP steam generator	3/0	HE WASTE HEAT	EDR	CS
Other coolers	6/0	HE FLOAT HEAD	EDR	A214/A516

(1) () denotes cladding material

3.2. Estimation of the operating and maintenance (O&M) cost

Other than the raw material costs, costs of utility, operating labor, catalysts and chemicals also have significant contributions on the O&M cost of a chemical plant. In this study, the raw material cost can be

easily estimated based on the material and energy balance given steady state simulation. Process fuels, steam and electricity are generated internally from the fuel gas header and the combined cycle power island. As the circulating water system is designed using Analyzer Utility Modules (AUM), process water is the only external utility considered in this economic model. The costs of catalysts and chemicals are listed in Table 6 for all four plant configurations. In APEA, the initial loading of catalysts and chemicals is specified as ‘quoted’ equipment, while costs for replacing catalysts and chemicals are specified under raw materials. For the water gas shift, Claus, isomerization and catalytic reforming units, the initial catalyst cost is included in the equipment cost. The catalyst in the liquefaction unit is replaced continuously. The catalyst in the catalytic reforming unit is replaced continuously to maintain the desired catalysts activity. [53] Other catalysts are replaced every five to ten years, depending on the catalyst life. Replacement costs of those catalysts are amortized when treated as raw materials. The number of operators is calculated based on the economic analysis given by Bechtel/Amoco [52].

Table 6

Cost of catalyst and chemicals in the direct CBTL plants with CCS (10,000 bbl/day).

	Unit Cost ⁽¹⁾	Initial (k\$)/Replacement (\$/hr) ⁽²⁾				Cost source
	(\$/unit)	SMR_CCS	SMR_VT	CG_CCS	CG_VT	
Catalysts						
Liquefaction	\$4.00/kg	661/461	661/461	661/461	661/461	[52]
Water gas shift	\$16774/m ³	0/75	0/75	0/75	0/75	[22,24]
Claus unit	\$4414/m ³	0/15	0/15	0/15	0/15	[24]
Steam reforming	\$22930/m ³	868/33	868/33	0/0	0/0	[54]
Hydrotreating	\$34.17/kg	6754/282	6754/282	6754/282	6754/282	[53]
Isomerization	\$4414/m ³	0/4.5	0/4.5	0/4.5	0/4.5	[53]
Chemicals						
Selexol solvent	\$3804/m ³	98/2.0	98/2.0	433/9.2	433/9.2	[22,24]
Amine solvent	\$2.16/kg	1355/17	301/3.8	350/4.4	0/0	[22,24]
ROSE-SR solvent	\$3/gallon	54/1.6	54/1.6	54/1.6	54/1.6	[52]
Total		9790/891	8736/878	8252/853	7902/848	

(1) Costs listed are the original value published in different years.

(2) The costs of catalyst and chemicals are escalated with the average Producer Price Index. [56,57]

4. Results and Discussion

The specified key design parameters for the base case scenario and the results obtained from the steady-state process models are provided in Table 7 for the four direct CBTL configurations. While the process model which is the basis for the material and energy balances has been validated in in our previous work [1], estimation of capital cost also needs to be compared with the data available in the open literature if feasible. Section 4.1 shows that comparison for capital costs. A number of sensitivity studies are

conducted as reported in Sections 4.2-4.4 by considering different hydrogen sources, key design parameters listed in Table 7, raw material price listed in Table 2, and investment parameters listed in Table 3. In Section 4.5, impacts of potential environmental credits are evaluated. Finally, the direct CBTL plant with various configurations is compared with the indirect CBTL plant in Section 4.6.

Table 7

Key design parameters and plant performance measures (base case).

Key design parameters	Value	Plant performance	SMR_CCS	SMR_VT	CG_CCS	CG_VT
Plant capacity (bbl/day)	10000	Coal/biomass (tonne/hr)	100/9	100/9	151/14	151/14
Biomass type	Wood chip	Shale gas (tonne/hr)	22	22	0	0
Biomass/coal (wt/wt, dry)	8/92	Gasoline/diesel (bbl/day)	2433/7557	2433/7557	2433/7557	2433/7557
Low pressure CO ₂ capture	MDEA/PZ	Net power (MW)	52.4	77.8	84.5	111.9
Extent of CCS if considered	High ⁽¹⁾	Efficiency (% , HHV)	64.3	66.5	60.0	62.1

(1) 90% of carbon in the raw materials is either converted to gasoline and diesel or stored in captured CO₂.

4.1. Comparison of the total project cost (TPC) estimation

In the limited techno-economic studies conducted for direct liquefaction processes, coal is the only feedstock considered; hydrogen is usually supplied by coal gasification; and no CCS facility is considered [4,52]. In this study, the liquefaction reactor feed only contains 8 wt% of biomass in the base case scenario, which is not expected to have significant impact on the TPC estimation. Hence, the capital cost estimation of the CG_VT process is validated by comparing with the estimates available in the open literature for the DCL plant with different capacities. The estimated costs of the SMR unit and CO₂ compression units are compared with the natural gas to liquids plant and the power plant separately and are found to have good match [20,24,54,55]. The Decision Analyzer tool in APEA is applied to change the plant capacity from our base case model for fair comparison. For some equipment items, parallel trains have to be considered, because of issues such as hardware constraints, high radial variation, etc. Table 8 summarizes the results of the comparison, while Table 9 provides detailed comparison of each plant section for the large scale case. Our estimations are found to be similar to the data reported by Shenhua [4] which is one of the only existing commercial scale DCL plants in the world, but slightly higher than the data reported by Bechtel/Amoco [52], mainly because the gasification cost estimated by Bechtel/Amoco in 1992 was lower than the data reported by NETL and others [22,24,27], even after it is escalated by CEPCI.

Table 8

Validation of capital cost estimation.

Process	CG_VT ⁽¹⁾	CG_VT ⁽²⁾
Reference	[4]	[46]
Capacity (bbl/day)	16300	61943
Biomass (wt %)	0	0
Total project cost (MM\$, 2015)		
Estimated	2024	6853
Reported	2086	6115

(1) The original capital cost is \$1.46 billion for a DCL facility in China in 2008 [4]. This value is adjusted by the reported location factor for China and escalated by CEPCI [58,59].

(2) The original capital cost is \$3.87 billion with 1991 pricing basis. The capital investment of the gasification unit reported by Bechtel/Amoco is lower than most recent estimation reported by NETL [22,24,52].

Table 9

Detailed comparison of equipment cost estimation (MM\$, 61943 bbl/day).

	Estimated	Reported		Estimated	Reported
Feed drying and handling	103.8	115.2	Hydrogen production ⁽³⁾	250.4	129.3
Liquefaction ⁽¹⁾	416.2	455.2	Air separation unit	138.2	165.0
Product upgrading ⁽²⁾	92.6	47.6	Sulfur recovery	46.0	24.1
Hydrogen purification	105.3	96.8	Total equipment cost	1178.1	1053.7
ROSE-SR	25.6	20.6	Total project cost	6711.1	6115.0

(1) Required solvent/feed ratio for liquefaction has been reduced since Bechtel/Amoco did their estimation in 1992.

(2) Naphtha upgrading was not considered in Bechtel/Amoco's design but in our design

(3) The equipment cost for gasification estimated by Bechtel/Amoco is lower than the data published in other resources [22,24,27,52].

4.2. Profitability analysis of four plant configurations (base case)

With the economic parameters listed in Table 2 and 3 as well as the material and energy balance shown in Table 7, the major economic measures of the base case are calculated and reported in Table 10. It is noticed that none of the four investigated configurations of the direct CBTL plants can make profit or have positive NPV due to the current low crude oil price (COP). However, the direct CBTL plants may start to payback once COP surpasses the reported BEOP, and be competitive with traditional petroleum industries once COP surpasses the reported EOP. The results also shows that the capital investments of the CG_CCS and CG_VT processes are much higher than those of the SMR_CCS and SMR_VT processes, because of the high capital cost and low hydrogen production efficiency of the gasification unit in comparison to the shale gas steam reforming unit [1,3]. As a result, the BEOP and EOP of the SMR_CCS and SMR_VT processes are higher than those of the CG_CCS and CG_VT processes, which

indicate that the direct CBTL plants will be more profitable if hydrogen is produced from low cost shale gas. Additionally, the relative penalty of CCS based on BEOP is about 10.2% if hydrogen produced from shale gas SMR and residual POX and 8.8% if hydrogen is produced from coal/biomass/residues CG, because CO₂ produced from gasification unit is at higher partial pressure and therefore easier to be captured and compressed for sequestration [1].

Table 10

Major economic measures (10,000 bbl/day, base case).

Process	SMR_CCS	SMR_VT	CG_CCS	CG_VT
Total project cost (MM\$)	1162	1080	1464	1387
Net present value (MM\$)	-408.6	-263.8	-591.7	-453.0
Internal rate of return (%)	6.0	7.3	5.2	6.2
Break-even oil price (\$/bbl)	86.1	77.3	97.5	88.9
Equivalent oil price (\$/bbl)	101.0	91.5	115.5	107.0

4.3. Effect of the economic parameters and plant capacities

Fig. 3-6 provide the results due to $\pm 25\%$ changes in the major plant economic inputs for all four configurations of the direct CBTL plant with a 10,000 bbl/day capacity. The results shows that the BEOP is between \$83.4/bbl to \$92.2/bbl for the SMR_CCS process, between \$74.5/bbl to \$82.9/bbl for the SMR_VT process, between \$93.4/bbl to \$104.7/bbl for the CG_CCS process, and between \$84.7/bbl to \$96.0/bbl for the CG_VT process. Fig. 7 shows the effect of plant capacity in comparison between the small-scale operation (10,000 bbl/day, base case) and the large-scale operation (50,000 bbl/day) for all four configurations. As the plant capacity increases, multiple trains may be required for different process sections. For example, three parallel trains are required by the liquefaction and hydrocarbon recovery section, when the plant capacity reaches 50,000 bbl/day. [5,52] The results indicate that the BEOP of the SMR_VT process decreases to \$56.9/bbl with high capacity, which is less than the COP of the second quarter of 2015. However, the BEOP of the CG_CCS and the CG_VT processes is still much higher than the COP even with a high plant capacity, because multiple trains are required by the gasification unit, one of the most expensive process sections.

Fig. 3. Sensitivity studies of the SMR_CCS process (10,000 bbl/day).

Fig. 4. Sensitivity studies for the SMR_VT process (10,000 bbl/day).

Fig. 5. Sensitivity studies of the CG_CCS process (10,000 bbl/day).

Fig. 6. Sensitivity studies of the CG_VT process (10,000 bbl/day).

Fig. 7. Effect of the plant capacity (10,000 and 50,000 bbl/day).

4.4. Effect of the biomass to coal ratio and the extent of CCS

In this study, two levels of biomass to coal weight ratio, 8/92 and 20/80, and two levels of CCS are considered and compared with the direct CBTL cases without CCS. For the case with low extent of CCS, all CO₂ removed in the hydrogen production and purification unit is sent to the CO₂ compression section preparing for CO₂ pipeline. For the case with high extent of CCS, additional CO₂ is captured from the low pressure sources, such as flue gas, and sent to the CO₂ compression section with the CO₂ captured from the hydrogen plant. The results are shown in Fig. 8 and Table 11. Fig. 8 indicates that the penalty of CCS increases with the increase in the extent of CCS and the decrease in the biomass to coal ratio. Table 11 indicates that the CO₂ emission can be significantly reduced even with the low extent of CCS, where no additional CO₂ capture is required. As a result, the BEOP and TPC do not increase considerably if only

low extent of CCS is considered. On the other hand, the penalty of CCS per unit of CO₂ capture in the cases with high extent of CCS is higher than that in the cases with low extent of CCS, because not only additional CO₂ needs to be captured but that the additional CO₂ needs to be captured from the low pressure sources, significantly increasing the operating cost and capital investment. The results also indicate that the overall cost and the penalty due to CCS decrease with the increase in the biomass content in the feedstock. Due to the higher H/C ratio in the biomass than coal, the hydrogen requirement in the liquefaction reactors gets reduced. As a consequence, the throughput of the hydrogen plant and associated CO₂ emission also gets reduced with the increase in the biomass content. To summarize, application of the CCS technology will increase the BEOP of the two direct liquefaction processes by about \$8.7/bbl (SMR_CCS) and \$8.55/bbl (CG_CCS), while addition of biomass will reduce the BEOP by about \$1.0/bbl.

It is noticed that even with the high extent of CCS and even after taking into account the CO₂ credit due to use of biomass, the SMR_CCS and CG_CCS processes with a biomass to coal ratio of 8/92 still have a higher carbon footprint than the petroleum refineries (about 8.12 kg CO₂/GJ product). However, if the biomass to coal ratio increases to 20/80, the CO₂ emission from both SMR_CCS and CG_CCS process with high extent of CCS is lower than the petroleum refinery.

Table 11

Performance and economics of the direct CBTL plants with different extent of CCS.

Biomass/coal (wt/wt)	8/92			20/80		
Extent of CCS	High	Low	No	High	Low	No
Hydrogen produced from shale gas steam reforming and residues partial oxidation (SMR)						
Total project cost (MM\$)	1162	1112	1080	1123	1044	1024
Internal rate of return (%)	6.0	7.0	7.3	5.9	6.9	7.3
CO ₂ emission (kg CO ₂ /GJ product)	12.0	26.3	53.2	11.5	21.9	47.2
CO ₂ emission with biomass credit ⁽¹⁾ (kg CO ₂ /GJ product)	9.4	23.7	50.6	5.6	16.0	41.3
Thermal efficiency (HHV, %)	64.3	66.0	66.5	67.6	68.7	69.2
Hydrogen produced from coal/biomass/residues co-gasification (CG)						
Total project cost (MM\$)	1464	1409	1387	1411	1366	1343
Internal rate of return (%)	5.2	5.9	6.2	5.1	5.9	6.1
CO ₂ emission (kg CO ₂ /GJ product)	14.3	27.3	77.4	13.8	23.5	72.7
CO ₂ emission with biomass credit ⁽¹⁾ (kg CO ₂ /GJ product)	9.9	22.9	73.0	3.2	12.9	62.1
Thermal efficiency (HHC, %)	60.0	61.2	62.1	62.1	63.1	63.6

(1) When biomass credit is accounted, CO₂ produced from biomass is deducted from CO₂ emission, which is the molar flowrate of biomass × carbon percentage in the biomass obtained from the ultimate analysis × (1- carbon

efficiency of the process) $\times$ the molecular weight of CO₂. The ultimate analysis of biomass and calculation of process carbon efficiency can be found in our previous publications [1].

Fig. 8. Effect of the extent of CCS.

4.5. Effect of potential environmental credits

In this section, three potential environmental credits are discussed for the SMR_CCS and CG_CCS processes with high-level CCS and a biomass to coal ratio of 20/80. For each potential environmental credit, two levels are considered, as shown in Table 12. Here, carbon credit is defined as the additional CO₂ captured from the processes with the baseline of petroleum refineries, which can be traded as a product in a carbon-constrained market. If the renewable/alternative energy certification is considered, the electricity can be sold at a premium. Here, we assume that the electric power generated from biomass qualifies for this credit, which is defined as the total power generated in the combined cycle island multiplied by the biomass HHV percentage in the feedstock. In addition, the federal government may apply lower tax rate to promote the development of renewable or alternative fuel related technologies, denoted as government-subsidized tax credit. The results in Table 13 show that the maximum reduction in BEOP is about \$7.1/bbl for the SMR_CCS process and \$8.8/bbl for the CG_CCS process if the proposed environmental credits are considered for the cases with a biomass to coal ratio of 20/80 while considering the value of all design and economic parameters the same as the base case. Combined with the sensitivity study shown in Section 4.4, the BEOP of the small scale SMR_CCS and CG_CCS can be reduced to \$77.9/bbl and \$87.6/bbl at the best case scenario. It is observed that the contribution from the carbon credit and renewable energy certification is not significant because the relatively low biomass percentage in the feed and also due to very high capital and operating costs of the DCL technology. Due to the same reason, the contribution of these two credits is smaller in the SMR_CCS process than that in the CG_CCS process.

Table 12

Definitions and levels of potential environmental credits.

Potential environmental credits	Description	High	Low	No
Carbon credit (\$/tonne carbon)	Additional CO ₂ captured	30	15	0
Renewable energy certification (\$/MWh)	Electricity from biomass	60	55	50
Government-subsidized tax credit (%)	Incentive tax rate for alternative fuel	30	35	40

Table 13

Potential environmental credits for the direct CBTL plants (10,000 bbl/day).

Difference in BEOP (\$/bbl)	SMR_CCS		CG_CCS	
Level of the credits	High	Low	High	Low
Carbon credit	-0.2	-0.1	-0.3	-0.2
Renewable energy certification	-0.2	-0.1	-0.5	-0.3
Government-subsidized tax credit	-6.7	-3.1	-8.0	-3.7

4.6. Direct versus indirect CBTL plant

The economic performance of chemical processes depends not only on the technology but also on the plant throughput and plant location. An increase in the throughput can reduce the break-even price of the product but results in higher risk due to the large initial capital investment. [60] The plant location can strongly affect the cost of labor, therefore affecting the cost of construction, operation and maintenance, cost of materials (i.e. feedstocks, materials of construction, catalysts and chemicals) and cost of land. The plant location can also affect investment parameters, such as the tax rate, interest rate, plant overhead, etc. Therefore for comparing the DCL technology with the ICL technology, economic analysis of both technologies is conducted with the 2015 pricing basis for North America for the same plant throughput.

A detailed process and economic model of the indirect CBTL plant based on the Fischer-Tropsch (FT) technology was developed and validated in our previous studies using 2014 pricing basis [6,7,26]. For fair comparison, previous economic model developed for indirect CBTL plant with CCS (FT_CCS) and without CCS (FT_VT) is updated to the 2015 pricing basis and the same economic parameter listed in Table 2 and Table 3 of this paper except plant contingency. It is noted that 8% of biomass and 10,000 bbl/day capacity are considered for all cases. Because of the difference in sources of CO₂ and their CO₂ partial pressure, the extent of CO₂ capture is different between the indirect and direct technologies for the cases with the low extent of CCS [1,26]. Hence, only the cases with the high extent of CCS and the cases

without CCS are considered in this section for fair comparison. For the indirect CBTL plants, the plant contingency is set to be 18%, because the technology is more proven and there are more industrial operating experiences than the direct CBTL plants. Additionally, the TPC estimation of indirect CBTL plants matches well with the industrial data, once 18% plant contingency is applied.

The results are shown in Fig. 9. Thermal efficiencies of the direct liquefaction plants with or without shale gas utilization are higher than that of the indirect liquefaction technology as expected [1,3,6,7,26]. However, the capital cost (TPC) of the indirect liquefaction technology (FT_CCS and FT_VT) is higher than that of the direct liquefaction technology with shale gas utilization (SMR_CCS and SMR_VT) but lower than that of the direct liquefaction technology without shale gas utilization (CG_CCS and CG_VT), mainly because of the large hydrogen demand in the direct liquefaction technology and the difference in efficiency and capital cost of the two hydrogen production technologies that are investigated here. The overall effect of process efficiency and the capital investment is reflected in the BEOP and EOP. Fig. 9 also shows that the BEOP and EOP of the CG_CCS and CG_VT processes are slightly higher than those of the FT_CCS and FT_VT processes, while those of the SMR_CCS and SMR_VT processes are much lower than those of the FT_CCS and FT_VT processes. It indicates that the direct CBTL plants are comparatively less competitive than the indirect CBTL plants even with a higher thermal efficiency, if the hydrogen required in the direct CBTL plants is entirely produced from the expensive and less efficient gasification route. If the hydrogen is produced through a more efficient and less expensive route, for example shale gas steam reforming, then the direct CBTL plants are more competitive than the indirect CBTL plants. It is also observed that if the shale gas price is higher, the economic performance of the SMR_CCS and SMR_VT processes may be worse than that of the FT_CCS and FT_VT processes. Table 14 shows that the BEOP of the SMR_CCS and SMR_VT processes becomes the same as the FT_CCS and FT_VT processes when the price of the shale gas increases to \$3.70/GJ or \$5.38/GJ, respectively. If the shale gas price keeps increasing, the indirect liquefaction process eventually becomes a better option.

The results also show that the CCS penalty of indirect CBTL plants is lower than both direct CBTL plants, because additional CO₂ needs to be captured in the direct CBTL plant to achieve high level of CCS [1], while the difference between the FT_CCS and FT_VT processes is only in the CO₂ compression unit [6,7,26]. As mentioned before, the plant contingency is specified to be 24% because the limited commercial experience of the direct CBTL plant. If the plant contingency is set to be 18%, the same as the indirect CBTL plant, the BEOP of the CG_VT processes reduced to \$85.2/bbl lower than that of the FT_VT process as shown in Fig. 6, because of reduced capital investment. However, the BEOP of the CG_CCS is still higher than that of the FT_CCS process, because of the higher CCS penalty.

Table 14

SMR processes versus FT processes (10,000 bbl/day).

	SMR_CCS		FT_CCS	SMR_VT		FT_VT
Shale gas price (\$/GJ)	2.25	3.70		2.25	5.38	
Break-even oil price (\$/bbl)	86.1	90.7	90.7	77.3	86.4	86.4

Fig. 9. Indirect and direct CBTL plants (10,000 bbl/day).

5. Conclusions

In this work, a techno-economic study is conducted for direct CBTL plants with shale gas utilization and CCS. BEOP is evaluated in all sensitivity studies by analyzing impacts of technology selection, economic inputs and design parameters, while other key economic measures, including NPV, IRR and EOP, are reported for the base case studies for four different plant configurations. The results shows that only the large-scale SMR_VT process, where shale gas is used for hydrogen production and the CO₂ is vented, can be profitable in the current market conditions due to the extremely low crude oil price. The BEOP of small scale direct CBTL plants without CCS is about \$77.3/bbl if hydrogen is produced from shale gas (SMR_VT) and \$88.9/bbl if hydrogen is produced from coal and biomass (CG_VT). Application of the CCS technology will increase the BEOP of the two direct liquefaction processes by about \$8.7/bbl (SMR_CCS) and \$8.55/bbl (CG_CCS), while addition of biomass will reduce the BEOP by about \$1.0/bbl, if any potential government-subsidized environmental credit is not considered. The maximum reduction in BEOP is about \$7.1/bbl for the SMR_CCS process and \$8.8/bbl for the CG_CCS process if

the proposed environmental credits are all considered. By comparing with the indirect CBTL plant with CCS, it is observed that the direct CBTL plant with CCS is economically superior if hydrogen required for liquefaction is produced from a cheap, H₂-rich source that is shale gas. The economic performance of the SMR_CCS and SMR_VT processes highly depends on the shale gas price. If CCS is considered, the BEOP from the indirect and direct technologies become the same when the shale gas price increases by about 60% compared to the base case. If CCS is not considered, the shale gas price has to increase by about 140% in comparison to the base case for the BEOP from both technologies to be the same. It should be noted that the purpose of this work was to develop the bases for process synthesis, steady-state modeling, and capital cost estimation, and to obtain a reasonable feasible start point for the further economic-based optimization and life cycle assessment of direct CBTL processes. The authors strongly believe that further improvement in the economics of the DCL technology can be obtained through plant-wide optimization, a task that the authors look forward to undertake in the near future.

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Appendix A. Equipment List

Table A.1 to Table A.6 list all standard equipment in the SMR_CCS process with a capacity of 10,000 bbl/day, and their material of construction (MOC), modeling and sizing approach. It is noted that all 'quoted' equipment is not listed in this section.

Table A.1

Equipment list of the liquefaction and hydrocarbon recovery unit.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC ⁽¹⁾
Reactors & vessels				
Inline hydrotreater	1/0	VT MULTI WALL	APEA	A387D
Slurry tank	2/0	AT MIX	APEA	A516
Slurry surge tank	1/0	VT CYLINDER	APEA	A285C
Slurry surge tank vent scrubber	1/0	VT CYLINDER	APEA	A516
High pressure high temp flash	1/0	VT CYLINDER	APEA	A387F (SS347)
Low pressure oil separator	1/0	TW TRAYED	Aspen Plus	A387D/A387D
High pressure cold flash	1/0	VT CYLINDER	APEA	A387B
Low pressure warm flash	1/0	VT CYLINDER	APEA	A516
Low pressure cold flash	1/0	VT CYLINDER	APEA	CS
High pressure warm flash	1/0	VT CYLINDER	APEA	A387D

Atmosphere still feed separator	1/0	VT CYLINDER	APEA	A387B
Wash water drum	1/0	HT HORIZ DRUM	APEA	CS
Sour water drum	1/0	HT HORIZ DRUM	APEA	CS
Recycle solvent tank	1/0	VT CYLINDER	APEA	A387D
Atmosphere still condenser drum	1/0	HT HORIZ DRUM	APEA	A516
Stabilizer condenser drum	1/0	HT HORIZ DRUM	APEA	A516
Distillation columns				
Atmosphere still tower	1/0	TW TRAYED	Aspen Plus	316L/316L
Atmosphere gas oil stripper	1/0	TW TRAYED	Aspen Plus	316L/316L
Atmosphere naphtha stripper	1/0	TW TRAYED	Aspen Plus	316L/316L
Stabilizer tower	1/0	TW TRAYED	Aspen Plus	A516/A258C
Vacuum still tower	1/0	TW TRAYED	Aspen Plus	SS410/SS410
Compressors, pumps & turbines				
Atmospheric still reflux pump	1/1	CP CENTRIF	APEA	SS casing
Stabilizer reflux pump	1/1	CP CENTRIF	APEA	CS casing
Slurry tank bottom pump	1/1	CP CENTRIF	APEA	SS316 casing
High pressure slurry feed pump	1/1	P RECIP MOTR	APEA	SS316 casing
Make up H ₂ compressor	1/1	GC RECIP MOTR	APEA	SS casing
Recycle H ₂ compressor	1/0	GC CENTRIF	APEA	SS316 casing
Stabilizer feed pump	1/0	CP CENTRIF	APEA	CS casing
Stabilizer feed compressor	1/1	GC CENTRIF	APEA	CS casing
ROSE-SR unit feed pump	1/1	CP CENTRIF	APEA	SS316 casing
Atmospheric still bottom pump	1/1	CP CENTRIF	APEA	SS316 casing
Atmospheric still feed pump	1/1	CP CENTRIF	APEA	SS316 casing
Gas oil product pump	1/1	CP CENTRIF	APEA	SS casing
Sour water pump	1/1	CP CENTRIF	APEA	SS casing
LVGO pumparound	1/1	CP CENTRIF	APEA	SS casing
HVGO pumparound	1/1	CP CENTRIF	APEA	SS casing
VGO product pump	1/1	CP CENTRIF	APEA	SS316 casing
Furnaces, boiler & heat exchangers				
Atmosphere still condenser	1/0	HE FIXED T S	EDR	A214/A516
Atmosphere still feed furnace	1/0	FU VERTICAL	APEA	347S
Stabilizer condenser	1/0	HE FIXED T S	EDR	A214/A516
Stabilizer reboiler	1/0	RB U TUBE	EDR	A214/A516
Slurry feed heat exchanger	2/0	HE FLOAT HEAD	EDR	316S/SS316
Slurry feed heat exchanger	4/0	HE FLOAT HEAD	EDR	316S/A387D (SS316)
Slurry feed furnace	1/0	FU BOX	APEA	347S
H ₂ pre heating	1/0	HE FLOAT HEAD	EDR	I825/SS304
H ₂ pre heating	1/0	HE FLOAT HEAD	EDR	321S/A387D
H ₂ feed furnace	1/0	FU BOX	APEA	347S
Recycle H ₂ heat exchanger	3/0	HE FLOAT HEAD	EDR	304LS/304L
Product heat exchanger	1/0	HE FLOAT HEAD	EDR	316LS/A387D (SS316)
IP steam generator	1/0	HE WASTE HEAT	EDR	CS

Water cooler	7/0	HE FLOAT HEAD	EDR	A214/A516
Product heat exchanger	1/0	HE FLOAT HEAD	EDR	A213C/A387C
LP steam generator	3/0	HE WASTE HEAT	EDR	CS

(1) () denotes cladding material

Table A.2

Equipment list of the syngas production unit.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC
Reactors & vessels				
Slag separator	1/0	VT CYLINDER	APEA	SS304
Scrubber	1/0	VT CYLINDER	APEA	SS304
Syngas KO drum	2/0	HT HORIZ DRUM	APEA	A516
flue gas KO drum	2/0	HT HORIZ DRUM	APEA	A516
Compressors, pumps & turbines				
Boiler feed water pump	1/1	CP CENTRIF	APEA	CS casing
Shale gas compressor	1/0	GC CENTRIF	APEA	CS casing
Furnaces, boiler & heat exchangers				
Boiler feed water heater	2/0	HE AIR COOLER	EDR	A214
Low pressure steam generator	2/0	HE WASTE HEAT	EDR	CS
Shale gas pre heater	1/0	HE FLOAT HEAD	EDR	A214/A516
Steam reformer pre heater	1/0	HE FLOAT HEAD	EDR	316S/SS316
Medium pressure steam generator	3/0	HE WASTE HEAT	EDR	CS
High pressure steam generator	2/0	HE WASTE HEAT	EDR	CS
Low pressure steam economizer	1/0	HE AIR COOLER	EDR	A214
High pressure steam economizer	1/0	HE AIR COOLER	EDR	A214
High pressure steam superheater	1/0	HE AIR COOLER	EDR	A214
Other coolers	3/0	HE FLOAT HEAD	EDR	A214/A516

Table A.3

Equipment list of the Selexol (AGR) unit.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC
Reactors & vessels				
High pressure flash	1/0	VT CYLINDER	APEA	A516
Medium pressure flash	1/0	VT CYLINDER	APEA	A516
Low pressure flash	1/0	VT CYLINDER	APEA	A516
H ₂ recovery drum	1/0	VT CYLINDER	APEA	A516
H ₂ S concentrator	1/0	VT CYLINDER	APEA	A516
H ₂ S stripper condenser drum	1/0	VT CYLINDER	APEA	A516
Selexol stripper condenser drum	1/0	HT HORIZ DRUM	APEA	A516
Lean solvent vessel	1/0	VT CYLINDER	APEA	A516

Distillation columns				
CO ₂ absorber	1/0	TW TRSYED	Aspen Plus	A516/A285C
H ₂ S absorber	1/0	TW TRSYED	Aspen Plus	A516/A285C
Selexol stripper tower	1/0	TW TRSYED	Aspen Plus	A516/A285C
Compressors, pumps & turbines				
NH ₃ compressor	1/0	GC CENTRIF	APEA	SS304 casing
H ₂ recovery compressor	1/0	GC CENTRIF	APEA	CS casing
Stripped gas compressor	1/0	GC CENTRIF	APEA	CS casing
Lean solvent pump	2/2	CP CENTRIF	APEA	CS casing
Recycle solvent pump	1/1	CP CENTRIF	APEA	CS casing
Selexol stripper reflux pump	1/1	CP CENTRIF	APEA	CS casing
Rich solvent pump	1/1	CP CENTRIF	APEA	CS casing
Furnaces, boiler & heat exchangers				
Selexol stripper condenser	1/0	HE FIXED T S	EDR	A214/A516
Selexol stripper reboiler	1/0	RB U TUBE	EDR	A214/A516
Recycle solvent cooler	1/0	HE FLOAT HEAD	EDR	A214/A516
H ₂ S absorber solvent cooler	1/0	HE FLOAT HEAD	EDR	A214/A516
Lean solvent cooler	1/0	HE FLOAT HEAD	EDR	A214/A516
Syngas cooler	1/0	HE FLOAT HEAD	EDR	A214/A516
Other coolers	4/0	HE FLOAT HEAD	EDR	A214/A516

487

488 **Table A.4**

489 Equipment list of the amine unit.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC ⁽¹⁾
Reactors & vessels				
MDEA/PZ storage tank	1/0	VT CYLINDER	APEA	A516
GT flue gas condenser	1/0	VT CYLINDER	APEA	A516
MDEA storage tank	1/0	VT CYLINDER	APEA	A516
CO ₂ Stripper condenser drum	2	HT HORIZ DRUM	APEA	A516
H ₂ S Stripper condenser drum	1	HT HORIZ DRUM	APEA	A516
Distillation columns				
High pressure absorber	1/0	TW PACKED	Aspen Plus	A516 (SS304)/M107YC
GT flue gas absorber	2/0	TW PACKED	Aspen Plus	A516 (SS304)/M107YC
SMR flue gas absorber	1/0	TW PACKED	Aspen Plus	A516 (SS304)/M107YC
CO ₂ Stripper tower	2/0	TW TRSYED	Aspen Plus	304L/M107YC
H ₂ S Absorber	1/0	TW PACKED	Aspen Plus	A516 (SS304)/1.0PPR
H ₂ S Stripper tower	1/0	TW TRSYED	Aspen Plus	304L/1.0PPR
Compressors, pumps & turbines				
Flue gas blower	2/0	FN CENTRIF	APEA	CS
CO ₂ Stripper reflux pump	2/2	CP CENTRIF	APEA	SS316 casing

H ₂ S Stripper reflux pump	1/1	CP CENTRIF	APEA	SS316 casing
GT rich solvent pump	2/2	CP CENTRIF	APEA	CS casing
SMR rich solvent pump	1/2	CP CENTRIF	APEA	CS casing
MDEA/PZ lean solvent pump	1/1	CP CENTRIF	APEA	CS casing
MDEA lean solvent pump	1/1	CP CENTRIF	APEA	CS casing
Furnaces, boiler & heat exchangers				
High pressure absorber pumparound	1/0	HE FLOAT HEAD	EDR	A214/A516
GT absorber pumparound	2/0	HE FLOAT HEAD	EDR	A214/A516
SMR absorber pumparound	1/0	HE FLOAT HEAD	EDR	A214/A516
CO ₂ Stripper condenser	2/0	HE FIXED T S	EDR	T150A/SS316
CO ₂ Stripper reboiler	2/0	RB U TUBE	EDR	316LW/SS316
H ₂ S Stripper condenser	1/0	HE FIXED T S	EDR	T150A/SS316
H ₂ S Stripper reboiler	1/0	RB U TUBE	EDR	316LW/SS316
GT flue gas cooler	1/0	HE FLOAT HEAD	EDR	A214/A516
MDEA/PZ lean/rich exchanger	1/0	HE PLAT FRAM	EDR	SS316
Lean solvent cooler	2/0	HE FLOAT HEAD	EDR	A214/A516
MDEA lean/rich exchanger	1/0	HE PLAT FRAM	EDR	SS316

(1) () denotes cladding material

Table A.5

Equipment list of the hydrocarbon upgrading unit.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC ⁽¹⁾
Reactors & vessels				
Gasoline storage tank	3/0	VT STORAGE	APEA	A516
Diesel storage tank	6/0	VT STORAGE	APEA	A516
Gas oil hydrotreater	2/0	VT MULTI WALL	APEA	A387F (SS347)
Hot high pressure flash	1/0	VT CYLINDER	APEA	A387D
Cold high pressure flash	1/0	VT CYLINDER	APEA	A516
Low pressure flash	1/0	VT CYLINDER	APEA	A516
Stabilizer condenser drum	2/0	HT HORIZ DRUM	APEA	A516
Main distillation condenser drum	1/0	HT HORIZ DRUM	APEA	A516
Distillation columns				
Main distillation tower	1/0	TW TRAYED	Aspen Plus	A516/A285C
Stabilizer tower	1/0	TW TRAYED	Aspen Plus	A516/A258C
Compressors, pumps & turbines				
Main distillation reflux pump	1/1	CP CENTRIF	APEA	CS casing
Stabilizer reflux pump	1/1	CP CENTRIF	APEA	CS casing
Makeup H ₂ compressor	1/1	GC RECIP MOTR	APEA	SS casing
Recycle H ₂ compressor	1/0	GC CENTRIF	APEA	SS316 casing
Stabilizer feed pump	1/1	CP CENTRIF	APEA	CS casing
Gas oil feed pump	1/1	CP CENTRIF	APEA	CS casing
Furnaces, boiler & heat exchangers				

Diesel pumparound	1/0	HE FLOAT HEAD	EDR	A214/A516
Gas oil pumparound	1/0	HE FLOAT HEAD	EDR	A214/A516
Main distillation condenser	1/0	HE FIXED T S	EDR	A214/A516
Main distillation feed furnace	1/0	FU VERTICAL	APEA	A213C
Stabilizer condenser	1/0	HE FIXED T S	EDA	A214/A516
Stabilizer reboiler	1/0	RB U TUBE	EDA	A214/A516
H ₂ pre heater	1/0	HE FLOAT HEAD	APEA	A214/A516
Feed H ₂ furnace	1/0	FU BOX	APEA	347S
Gas oil feed pre heater	1/0	HE FLOAT HEAD	EDR	A213D/A387D
Low pressure steam generator	3/0	HE WASTE HEAT	EDA	CS
Heavy diesel cooler	1/0	HE FLOAT HEAD	EDR	A214/A516
Light gas oil cooler	1/0	HE FLOAT HEAD	EDR	A214/A517
Other coolers	4/0	HE FLOAT HEAD	EDR	A214/A516

(1) () denotes cladding material

Table A.6

Equipment list of the combined cycle power island.

Equipment	#Required/Spares	Model in APEA	Sizing	MOC
Reactors & vessels				
High pressure steam blowdown	1/0	HT HORIZ DRUM	APEA	A516
Compressors, pumps & turbines				
High pressure BFW pump	1/1	CP CENTRIF	APEA	CS casing
Medium pressure BFW pump	1/1	CP CENTRIF	APEA	CS casing
Low pressure BFW pump	1/1	CP CENTRIF	APEA	CS casing
Condenser pump	1/1	VP MECH BOOST	APEA	CS casing
Steam turbine	1/0	EG TURBO GEN	APEA	CS casing
Fuel gas compressor	1/0	GC CENTRIF	APEA	CS casing
Gas turbine	1/0	EG TURBO GEN	APEA	CS casing
Furnaces, boiler & heat exchangers				
High pressure pre economizer	1/0	HE AIR COOLER	EDA	A214
High pressure BFW economizer	1/0	HE AIR COOLER	EDA	A214
High pressure steam superheater	1/0	HE AIR COOLER	EDA	A214
Medium pressure steam reheater	1/0	HE AIR COOLER	EDA	A214
Boiler feed water heater	1/0	HE AIR COOLER	EDA	A214
Air ejector	1/0	HE FLOAT HEAD	EDR	A214/A516
Steam packing exhaustor	1/0	HE FLOAT HEAD	EDR	A214/A516
High pressure steam generator	1/0	HE WASTE HEAT	APEA	CS
Low pressure steam generator	1/0	HE WASTE HEAT	APEA	CS
Surface condenser	1/0	C BAROMETRIC	APEA	CS

Nomenclature

498	AGR = acid gas removal	520	HHV = higher heating value
499	APEA = Aspen Process Economic Analyzer	521	HRSR = heat recovery and steam generation
500	AUM = Analyzer Utility Modules	522	HTI = Hydrocarbon Technologies Inc.
501	BEOP = break-even oil price	523	ICL = indirect coal liquefaction
502	BFD = block flow diagram	524	IGCC = integrated gasification combined
503	BFW = boiler feed water	525	cycle
504	BOP = balance of plant percentage	526	IRR = internal rate of return
505	CBTL = coal-biomass to liquids	527	ISBL = inside battery limit
506	CCS = CO ₂ capture and storage	528	LHSV = Liquid hourly space velocity
507	CEPCI = Chemical Engineering Plant Cost	529	MOC = materials of the construction
508	Index	530	NGCC = natural gas combined cycle
509	COP = crude oil price	531	NPV = net present value
510	CTSL = catalytic two-stage liquefaction	532	O&M = operating and maintenance
511	CG = co-gasification	533	OSBL = outside battery limit
512	DCL = direct coal liquefaction	534	POX = partial oxidation
513	DIP = direct permanent investment	535	PSA = pressure swing adsorption
514	EDR = Exchanger Design and Rating	536	RON = research octane number
515	EIA = Energy Information Administration	537	ROSE-SR = Residual Oil Supercritical
516	EOP = equivalent oil price	538	Extraction-Solids Rejection
517	FCI = fixed capital investment	539	TPC = total project cost
518	FT = Fischer-Tropsch	540	SMR = steam methane reforming
519	GT = gas turbine	541	WGS = water gas shift
542	Reference		

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