

CHAPTER IV

CASE STUDY 1: HYDROGENATION OF CO₂ INTO METHANOL

4.1 Base Case Design

4.1.1 Process Simulation

4.1.1.1 *Initializations and Feeds*

The first necessary element is the collection of information about the emissions and off-gases that will be implemented as the feed source for the process. The CO₂ sources for the conversion process are found from published information on the emissions produced by a variety of sectors. Meanwhile, information for hydrogen sources is gathered from different technologies in which both conventional and alternative energy resources are used.

4.1.1.1.1 *CO₂ Sources*

As mentioned above, the highest CO₂ emitted comes from the combustion of fossil fuels to support heat, power for transportation and especially for electricity (Le Quéré *et al.*, 2013). Among fossil sources, coal is the world's most abundant supply to meet about 23 percent of the total world primary energy demand, and 38 percent of global electricity requirement (Hammond *et al.*, 2011). Moreover, coal accounts for the biggest share of CO₂ emission. Therefore, to fulfill increasingly severe environmental regulations, the implementation of CO₂ capture process for a coal power plant has to be conducted.

Three main techniques for CO₂ capture from coal power plants are currently developed: post-combustion capture, pre-combustion capture, and oxy-fuel combustion capture. Most researches are focused on the post-combustion process since the advantages of retrofitting and flexibilities are obvious. This method is utilized to capture CO₂ from flue gas of a conventional combustion chamber so that the combustion process is kept unchanged. Therefore, the CO₂ of flue gas from a coal-fired power plant is used as the feed in this work. The detail of CO₂ capture is not within the scope of this thesis; a work from Amann (2007) is used for information on the required energy, needed materials and the cost of implementing a CO₂ capture. MEA (30 percent mass concentration) is employed to capture CO₂ of flue

gases from a subcritical coal power plant with desulfurization (capacity of 537 MWe). After compressing to compensate the pressure drop, flue gases are introduced to the bottom of the absorption column. The lean solvent is introduced at the top of the column to react with CO₂. At the top of the column, clean gas is recovered. The rich solvent from the bottom of the absorption column is preheated by the regenerated stream before being sent to the regeneration column. The regeneration involves a reboiler and a condenser while low pressure steam is used in the reboiler to reverse the reaction between the amine and CO₂. The regenerated stream is sent back to the absorption column after preheating the rich solvent. The pure CO₂ (99.9 percent purity) is recovered at the top of the regeneration column and then dehydrated and compressed. The typical capture process is presented in Figure 4.1. A detail of this process is also showed in Table 4.1.

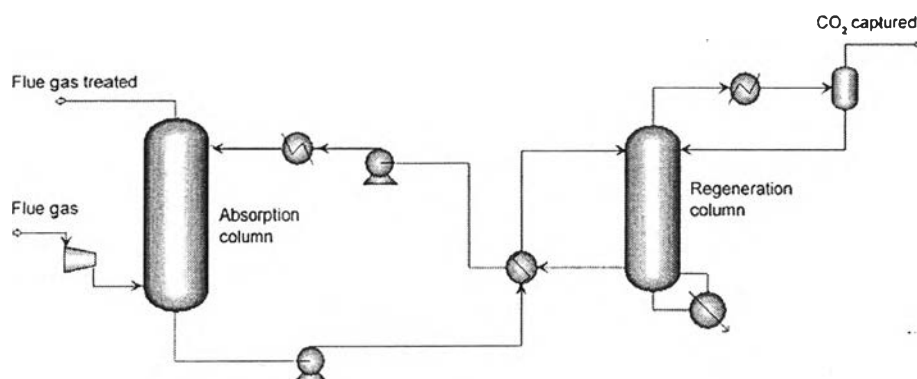


Figure 4.1 Flowsheet of the CO₂ capture unit.

Table 4.1 Tabulated information of the relevant results from CO₂ capture simulations (Amann, 2007)

Variable	Value
Flue gas amount	16806.82 kmol/hr
CO ₂ removal	85%
Flue gas compression	3.9 MWe
Steam to regeneration column	77.7 MWth
CO ₂	88 ton/hr
CO ₂ emission	0.723 kg/kWe

In this study, a medium size of methanol plant is selected as a methanol plant capacity (20000 kg/hr). This capacity allows the process to ensure the feasibility in terms of equipment size in an industrial scale. To adapt for methanol productivity (20000 kg/hr), the value of flowrate of CO₂ product is scaled down to 29249 kg/hr.

4.1.1.1.2 Hydrogen Sources

Nuclear, natural gas and coal, biomass, and other renewable sources are the candidates to produce hydrogen. Moreover, diverse technologies can be employed to produce hydrogen, including thermal (natural gas reforming, biomass, and coal gasification), electrolytic (water breaking) and photolytic (breaking water by solar energy). However, hydrogen is commonly produced from natural gas steam reforming with nearly 50 percent of global demand, following 30 percent from oil reforming. Coal gasification and water electrolysis account for 18 and 3.9 percent, respectively (Holladay *et al.*, 2009). In near future, fossil-based hydrogen production will continue to be constructed and operated. Therefore, hydrogen from steam reforming of natural gas is used as the feed in base case.

The details of hydrogen production are not within the scope of this thesis; all values needed are referred from literature. The cost to produce hydrogen by various processes is taken from papers Yumurtaci *et al.* (2004); Bartels *et al.* (2010) after updated to 2014 dollars by using Chemical Engineering Plant Cost Index, presented in Table 4.2. To adapt for methanol productivity (20000 kg/hr), the value of flowrate of hydrogen product is selected around 4022 kg/hr.

Table 4.2 Cost to produce hydrogen with different scenarios

Method	Cost (\$/kgH ₂)	Reference
<i>Methane steam reforming</i>	2.885	Bartels <i>et al.</i> (2010)
<i>Wind/electric</i>	4.14	
<i>Nuclear/steam electrolysis</i>	6.42	
<i>Solar thermal</i>	3.07	
<i>Biomass</i>	1.59	
<i>Hydroelectric</i>	1.28	Yumurtaci <i>et al.</i> (2004)

4.1.1.2 Methanol Synthesis

The process is reproduced from the work of Van-Dal *et al.* (2013) as a CO₂ conversion process. A typical methanol synthesis as shown in Figure 4.2 is selected as the base case design.

The major assumptions for simulation are as follows:

- The methanol productivity was fixed at around 20000 kg/hr with its purity higher than 99.5 percent.
- 100 percent purity of hydrogen and CO₂ were used.
- The methanol reactions and kinetic model were employed from the study of Bussche *et al.* (1996).

4.1.1.2.1 Process Flowsheet Description

CO₂ at 1 bar and 25°C is compressed to 75.7 bar through a string of compressors while hydrogen is compressed from 30 bar to 75.7 bar in a single compressor. After mixing them together, two gas streams are combined with the recycle stream. Preheated by the effluent from the reactor, the stream is fed to the fixed bed reactor at 215°C. The outflow is divided into two streams. One is employed to preheat the feed to the reactor while another is used to support the reboiler and preheat the feed to the distillation column. The two streams are blended and cooled to 40°C, which allows water and methanol to be condensed and removed from the unreacted gases in the first flash drum.

The liquid flow is expanded to 1.2 bar through valves before injected into the second flash drum to remove most residual gases. The remaining stream is heated up to 80°C and fed to the distillation column. While the bottom product of the column is water, crude methanol comes out of the top in gaseous phase. After cooled to 50°C, crude methanol is sent to the last flash drum to completely separate non-reacted gases from the top. Methanol product is achieved at the bottom of the flash drum in the liquid form.

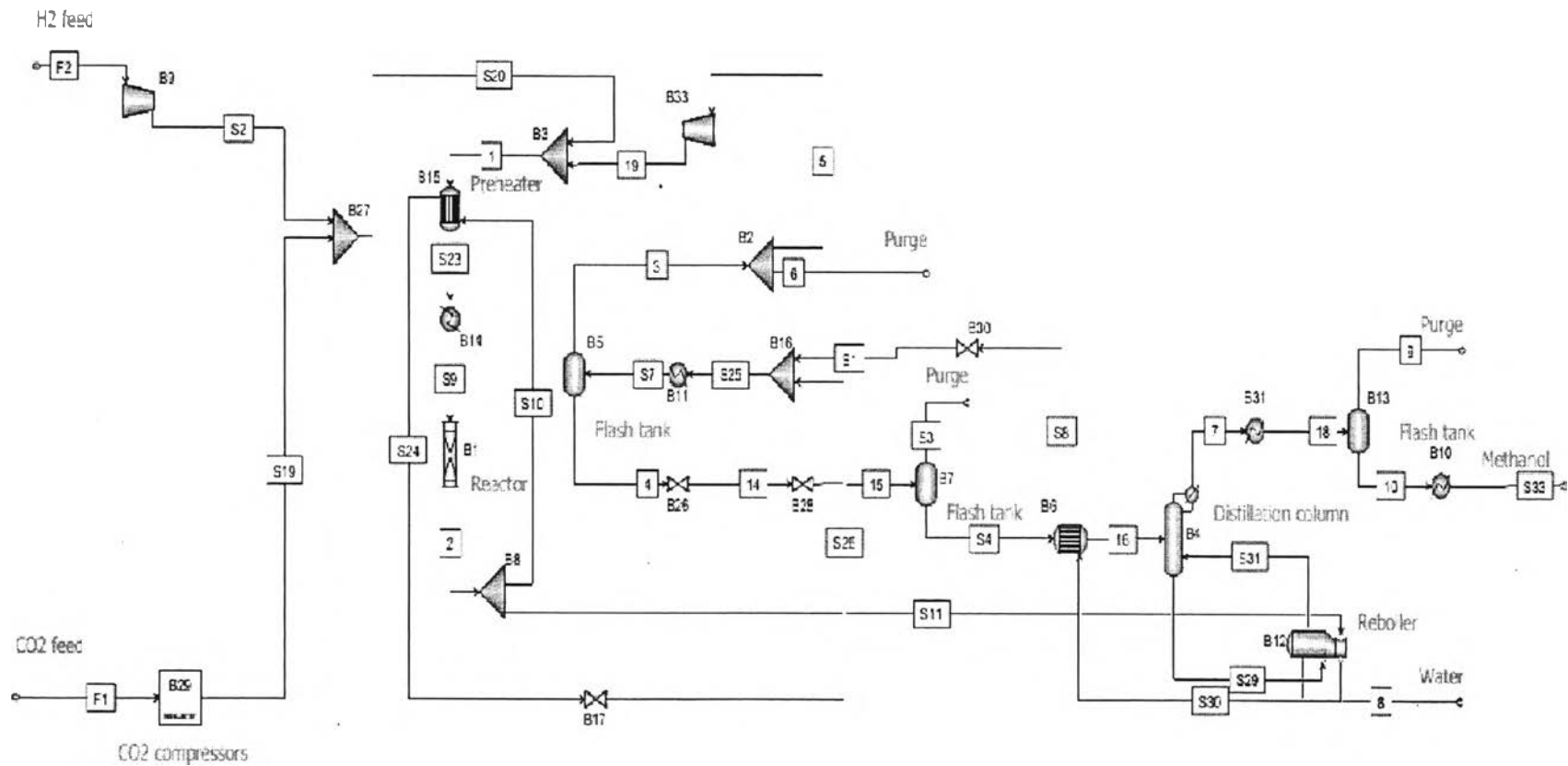


Figure 4.2 Flowsheet of the hydrogenation of CO₂ into methanol for the base case design.

4.1.1.2.2 Process Input

From the published information, simulation is used to construct the base case. It is important to note that the publication does not provide all the necessary data, and therefore some values were calculated to make the results become reasonable. All input data are presented in Table 4.3.

Table 4.3 Input data of the methanol synthesis

Feed Characteristics	
Variable	Value
H ₂ /CO ₂	3
Flow rate of H ₂ (kg/hr)	4022
Flow rate of CO ₂ (kg/hr)	29249
Methanol Reactor Conditions	
Inlet temperature (°C)	215
Inlet pressure (bar)	75.7
Reaction type	Exothermic
Reactor model in Aspen Plus	Plug flow reactor, kinetic model applied (Bussche <i>et al.</i> , 1996), adiabatic reactor
Methanol Reactor Design	
Number of reactor tubes	2659
Tube diameter (m)	0.05
Tube length (m)	12.2
Bed voidage	0.4
Catalyst particle diameter (m)	0.0055
Catalyst density (kgcat/m ³ cat)	1775
Separation Unit Designs	
The First Flash Tank	
Temperature (°C)	40
Pressure (bar)	73
The Second Flash Tank	
Temperature (°C)	27.5
Pressure (bar)	1.2
The Third Flash Tank	
Temperature (°C)	50
Pressure (bar)	1

Distillation Column	
Number of stages	26
1 st stage pressure (bar)	1
Pressure drop (bar)	0.1
Reflux ratio	1.2

4.1.1.3 Simulation Results

The inlets of the process are the CO₂ and hydrogen feed. The outlet flows are three purge gas streams from the flashes, water as the bottom product of the distillation column and methanol as the bottom product of the third flash. Mass balance and methanol product characteristics results are presented in Table 4.4 and 4.5, respectively.

Table 4.4 Mass balance

Inputs		Outputs	
	kg/hr		kg/hr
CO ₂ Feed	29249	Purge 1	448
Hydrogen Feed	4022	Purge 2	1210
		Purge 3	47
		Water	11396
		Methanol	20170
Sum	33271	Sum	33271

Table 4.5 Product characteristics

Variable	Value
Flow rate of methanol product (kg/hr)	20170
Purity (wt%)	99.5
Mass Fraction	
CO ₂	0.002
H ₂	949 PPB
Water	0.002
CO	788 PPB
Methanol	0.995

4.1.2 Sustainability Analysis

As this process is aimed to reduce CO₂ amount in atmosphere, the net CO₂ emission must be evaluated. To attain net CO₂ reduction, the process has to use more CO₂ than it discharges (such as from purge gas) or produces (primarily through energy requirements). The net CO₂ emission is described by the following equation.

$$n_{CO_2,net} = n_{CO_2,purge} + n_{CO_2,utilities} - n_{CO_2,feed} \quad (4.1)$$

This value is negative when the output amount of CO₂ from the process is lower than the input.

Table 4.6 Net CO₂ emission for only methanol synthesis

	Value	Unit
Direct (Purge)		
Purge 1	235.7	kgCO ₂ /hr
Purge 2	950.0	kgCO ₂ /hr
Purge 3	22.4	kgCO ₂ /hr
Total Direct	1208.2	kgCO ₂ /hr
Indirect (Utilities)		
Electricity	4240.2	kgCO ₂ /hr
Hot oil	0.0	kgCO ₂ /hr
Total Indirect	4240.2	kgCO ₂ /hr
Feed	29249.0	kgCO ₂ /hr
Net CO₂ Emission	-23800.6	kgCO ₂ /hr
	-1180.01	kgCO ₂ /tMeOH

Table 4.6 shows the net CO₂ emission from only methanol synthesis. It illustrates that indirect emission accounts for the largest part of CO₂ emission in methanol synthesis with more 78 percent while only 22 percent comes from purge gases. In addition, the process reveals the possibility of CO₂ reduction by achieving net negative CO₂ emission.

However, Quadrelli *et al.* (2011) demonstrated that the hydrogen production accounts for the biggest share of CO₂ emission in hydrogenation route. In addition, the CO₂ capture to obtain concentrated CO₂ emits CO₂. Therefore, to evaluate

the process comprehensively in terms of sustainability, three processes including hydrogen production, CO₂ capture and methanol synthesis have to be combined together as presented in Table 4.7.

Obviously, even though the process is used to reduce CO₂ emissions, operation of the conversion process leads to a non-negligible emission of CO₂. If only considering the methanol synthesis, the process provides the feasible way to achieve the sustainable design. Despite the promise of reducing CO₂ emission, when the total process is taken into account, hydrogen production is a critical parameter influencing the sustainable result. As a result, alternatives for hydrogen production instead of steam reforming of natural gas have to be assessed.

Table 4.7 Net CO₂ emission for the total process

	Value	Unit
Hydrogen Production		
Net CO ₂ emission	48663.0	kgCO ₂ /hr
CO₂ Capture		
Net CO ₂ emission	4671.6	kgCO ₂ /hr
Methanol Synthesis		
Net CO ₂ emission	-23800.6	kgCO ₂ /hr
Total		
Net CO ₂ emission	29534.0	kgCO ₂ /hr
	1464.3	kgCO ₂ /tMeOH

4.1.3 Economic Evaluation

Since the implementation of the final design is likely based on economic factors, an economic analysis is necessary to evaluate whether the process is feasible in terms of economy or not. This section of the report supports information of economic issue for the base case design which was calculated by using ECON software. The details to calculate the capital and production cost are presented in Appendix B.1.1.

This process design was modeled and simulated by the Aspen Plus 8.6 as mentioned earlier. The capacity of this plant is nearly 20000 kg/hr or around 162 kt/year. The plant operates 333 days/year or approximately 8000 hours/year.

4.1.3.1 Capital Cost of Base Case Design

The outcome of the Total Capital Investment (TCI) calculations for the base case design is 55.46 MM\$. The greatest share of 61 percent is from direct cost section, followed by indirect cost and working capital section comprising 24 and 15 percent, respectively as seen in Figure 4.3.

Clearly, Figure 4.3 shows that the direct costs are what takes the largest piece in the capital investment, thus it is interesting to see what constitutes the direct costs.

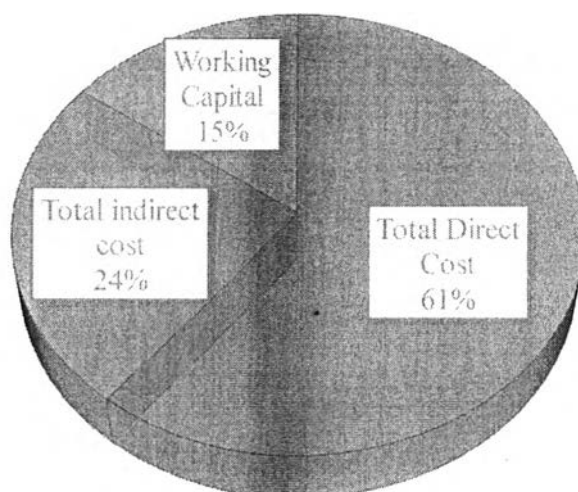


Figure 4.3 Breakdown of the total capital investment.

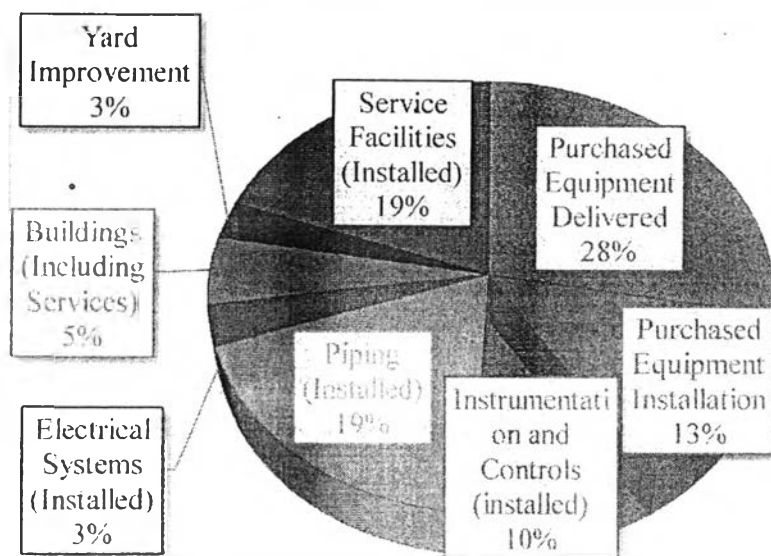


Figure 4.4 Breakdown of the direct cost.

As seen from Figure 4.4, the equipment costs (purchased equipment delivered) contribute the largest weight on the direct costs; hence, it is the most influence on the TCI. Table B.1.6 summarizes sizing and purchased cost of each equipment. The breakdown of equipment costs is illustrated in Figure 4.5 to gain further insight.

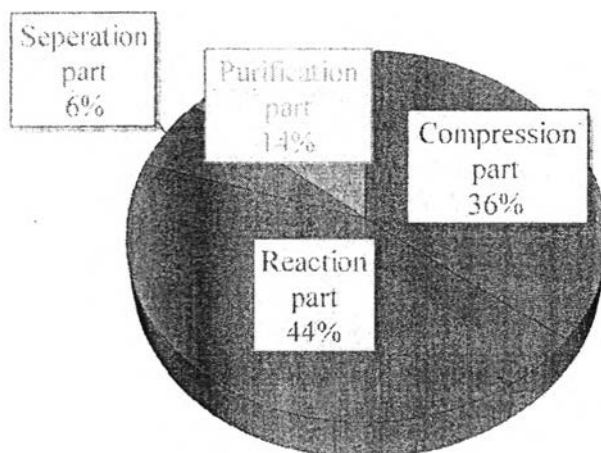


Figure 4.5 Contribution to equipment costs of each area of the process.

As shown in Figure 4.5, 44 percent of the equipment cost comes from the reaction part, followed by the compression and purification part with 36 and 14 percent, respectively.

4.1.3.2 Production Cost of Base Case Design

The result of the total production cost (without depreciation) calculations for the base case design is 128.29 MM\$/year. The greatest share of around 87 percent comes from the variable cost section, followed by the general expense, the plant overhead and fixed charges section constituting 9, 3 and 1 percent, respectively. The breakdown of the total product cost can be seen in Figure 4.6.

Obviously, the variable cost is the highest portion for the total production cost which mainly results from raw materials and utility cost as presented in Figure 4.7.

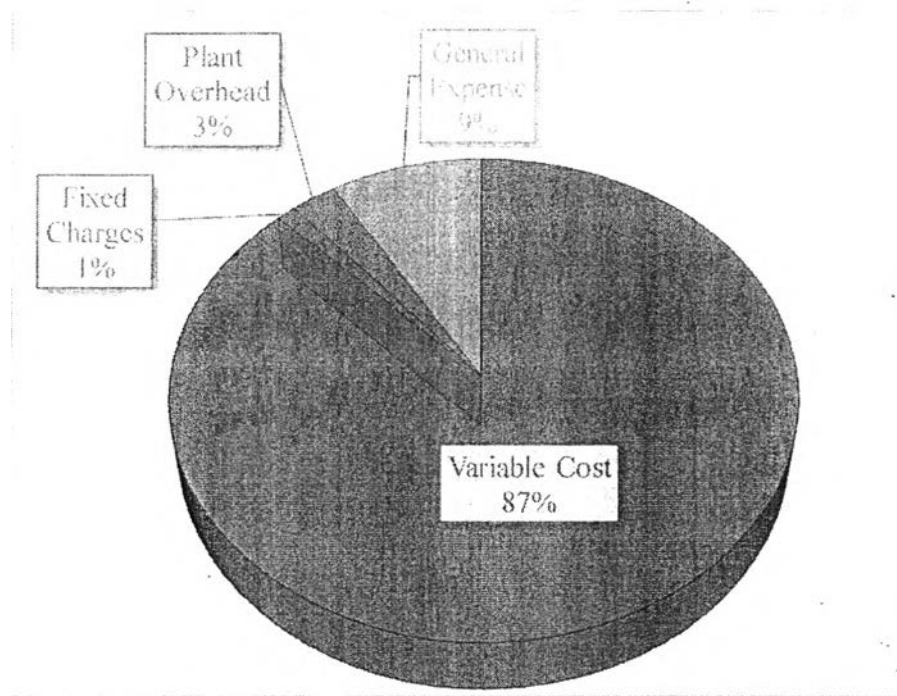


Figure 4.6 Breakdown of the total production cost.

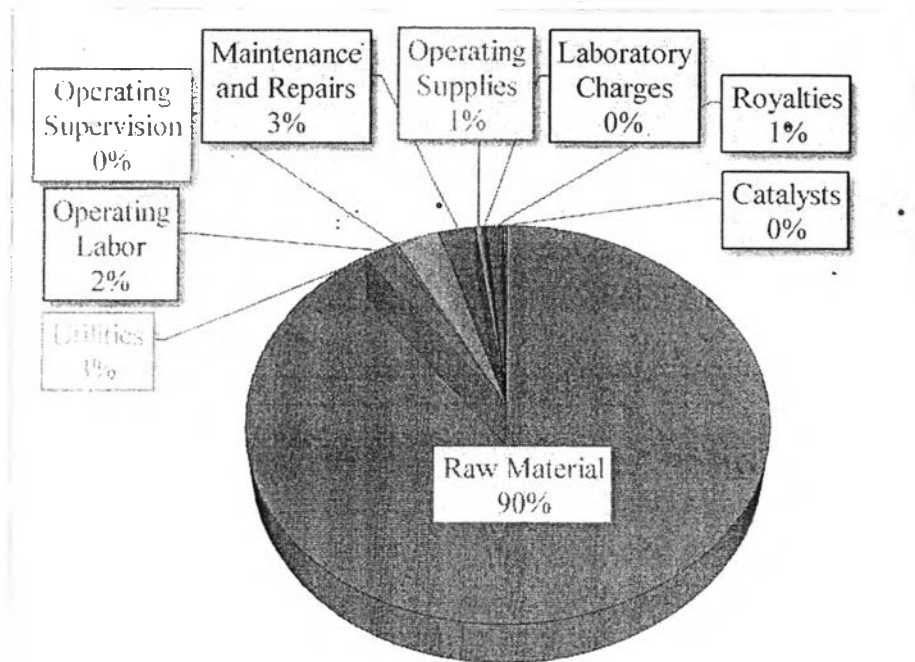


Figure 4.7 Breakdown of the variable cost.

Therefore, next interesting evaluation for the production cost is to show each of raw material and utility cost to see which components have the most effects. These results are presented in Figure 4.8 and Figure 4.9, respectively.

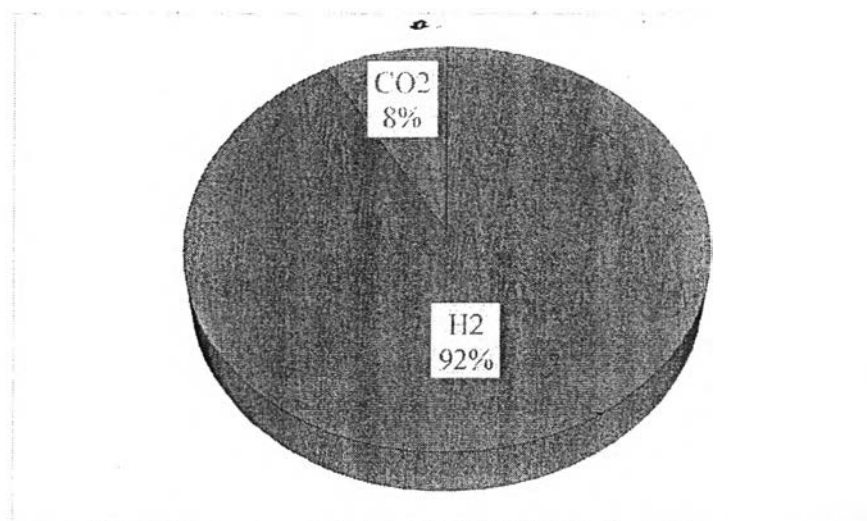


Figure 4.8 Breakdown of raw materials cost.

From Figure 4.8, hydrogen accounts for more than 90 percent of the cost for raw materials. Although the process requires less quantity of it, the cost for hydrogen production is very expensive.

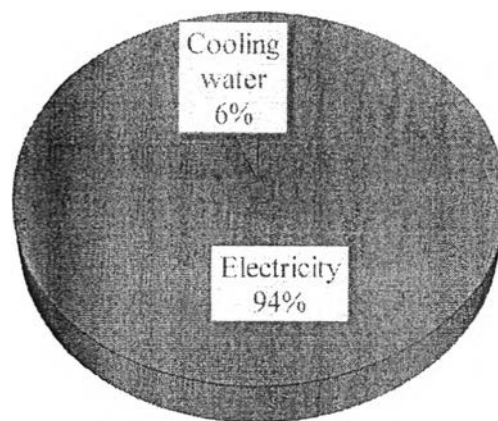


Figure 4.9 Breakdown of utilities cost.

As seen in Figure 4.9, electricity has the highest influence on the production cost because it is mainly consumed in the compression part.

4.1.3.3 *Economic Sensitivity Analysis of Base Case Design*

The economic sensitivity analysis is made to the raw materials, product price, labor cost, capital cost, equipment cost and utilities cost.

As seen from Figure 4.10, the highest influence on Net Present Value (NPV) is the cost of raw materials, which makes raw materials become the primary factor affecting the profit. Furthermore, the product price has the high impact on the NPV.

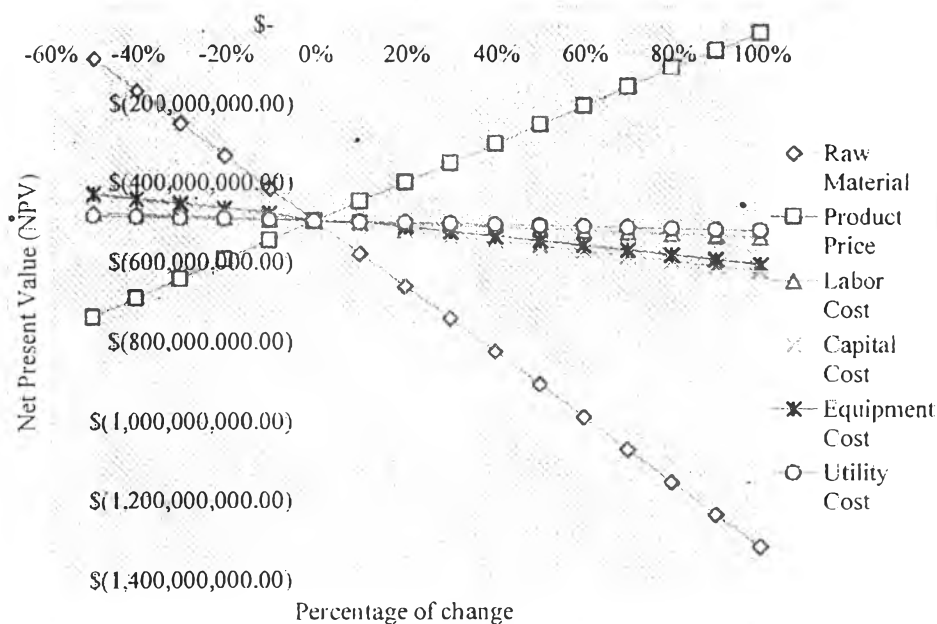


Figure 4.10 Sensitivity analysis compare to NPV.

4.1.3.4 Profitability of Base Case Design

Profitability is the measure of the amount of profit that can be obtained from a given situation. It is as common denominator for all business activities. The determination and analysis of profits obtainable from the investment of capital and the choice of the best investment among various alternatives are major goals of the investment analysis.

For this work, the life time of the project is assumed to be 20 years. The plant will be built in US. The MARR (Minimum Acceptable Rate of Return) is fixed to be 15 percent. The depreciation for the plant is estimated to be 20 years by MACRS method. The income tax rate that has to be paid to the government is assumed to be 25 percent. According to the increase of methanol price in the future, the inflation is set. The inflation rate of construction, product and total product cost are assumed to be 2 percent, 10 percent and 10 percent, respectively. The inflation rate of the product is set by the real increasing price data in the previous year (Zicha, 2014) and the rests are set by using the product price as reference. The details of all assumption are

presented in Table B.1.9. The summary of investment analysis for the base case design is shown in Table 4.8.

Table 4.8 Profitability of the base case

Profitability	
not include time value of money	
Rate of Return	-346%
Pay Back Period	-0.25
Net Return	\$ (205,055,939.05)
include time value of money	
Annual End of Year cash flows and discounting	
Net Present Worth	\$ (868,328,459.00)
DCFR	0.15
Continuous cash flows and discounting	
Net Present Worth	\$ (931,832,621.75)
DCFR	0.14

According to the result, all parameters are of negative values, which means this project is clearly not worth for investment. Moreover, cumulative cash flow shows that the plant will have no profit after constructed as seen in Figure 4.11. Therefore, the process must be improved to achieve both environmental and economic targets.

4.2 Sensitivity Analysis

Zhang *et al.* (2006) and Fornero *et al.* (2011) indicated that inlet reactor temperature, pressure and feed composition (H_2/CO_2) are key factors in the improvement of methanol synthesis. They therefore are varied respectively and observed how the net CO_2 emission and production cost are changing. The production cost in this section is calculated by the following equation.

$$\begin{aligned}
 & \textit{Production cost} \\
 & = CO_2 \textit{ capture cost} + H_2 \textit{ production cost} \quad (4.2) \\
 & + \textit{Utilities cost}
 \end{aligned}$$

Although this is the simple form of the real production cost, it is still valuable to evaluate the process in this step. The reason is the production cost mainly depends on the cost of raw materials as discussed above. In addition, basically if the process structure does not change significantly between different scenarios, the variation of production cost leading to the change of capital cost will result in the same trend. Therefore, the above equation will be used as a tool to quickly evaluate the economic aspect in next sections.

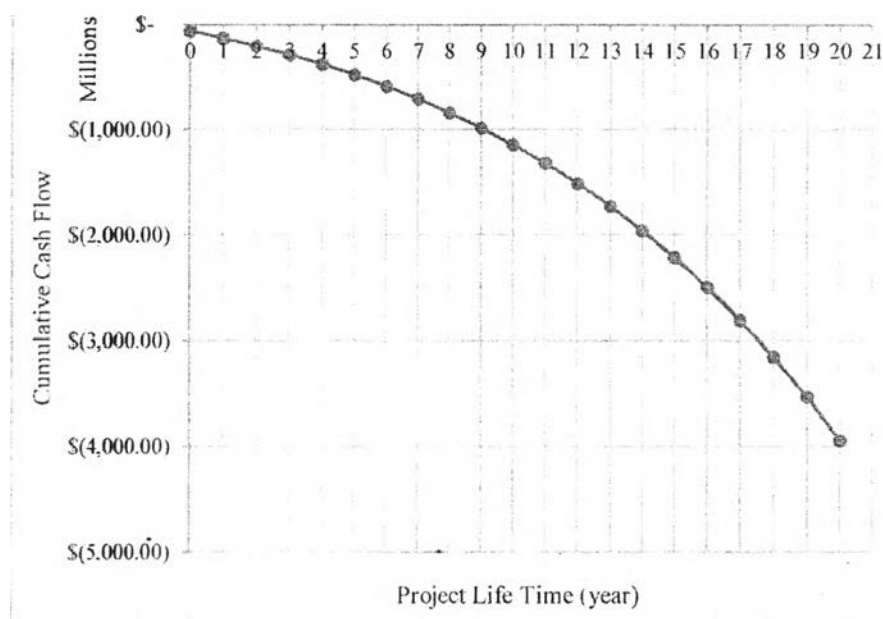


Figure 4.11 Cumulative cash flow for 20 year project of the base case.

To compare among difference scenarios, the methanol productivity is fixed at around 20000 kg/hr with its purity higher than 99.5 percent.

4.2.1 Inlet Methanol Reactor Temperature

As seen in Table 4.9 and 4.10, when changing the inlet methanol reactor temperature, the process shows the lowest net CO₂ emission and production cost at inlet methanol reactor temperature of 200°C. Because the methanol synthesis from CO₂ and hydrogen is exothermal reaction, it is disadvantageous to methanol

production when the reaction temperature increases. As a result, the direct CO₂ emission increases because of more unreacted CO₂ in purge gases. In addition, at higher temperatures, higher unreacted gases require more energy at the recycle compressor, which leads to an increase of indirect CO₂ emission. At lower inlet temperatures, the reaction is kinetically limited, which makes simulation not converge. On the other hand, quite noticeable rise of the production cost is observed due to higher requirement of electric demand when elevated temperatures are employed. In addition, less methanol productivity at high temperatures is the significant factor, leading to an increase of the production cost. In this step, 200°C is chosen as the new temperature of the process.

Table 4.9 Relationship between the inlet methanol reactor temperature and net CO₂ emission

Temperature (°C)	Direct CO ₂ Emission (kgCO ₂ /tMeOH)	Indirect CO ₂ Emission (kgCO ₂ /tMeOH)	Net CO ₂ Emission (kgCO ₂ /tMeOH)	Methanol Productivity (kg/hr)	Methanol Purity
200	-1388.7	2843.8	1455.1	20230	0.9960
205	-1389.2	2847.2	1458.0	20211	0.9958
210	-1389.7	2850.7	1461.1	20191	0.9956
215	-1390.2	2854.5	1464.3	20170	0.9954
220	-1390.9	2858.5	1467.6	20147	0.9952

Table 4.10 Relationship between the inlet methanol reactor temperature and production cost

Temperature (°C)	Energy Consumption (electricity) (\$/tMeOH)	Water Consumption (\$/tMeOH)	CO ₂ Capture cost (\$/tMeOH)	H ₂ Production cost (\$/tMeOH)	Production Cost (\$/tMeOH)
200	19.0	1.2	51.0	573.5	644.8
205	19.1	1.2	51.1	574.1	645.5
210	19.2	1.2	51.1	574.6	646.2
215	19.3	1.2	51.2	575.3	646.9
220	19.4	1.2	51.2	575.9	647.7

4.2.2 Inlet Methanol Reactor Pressure

As shown in Table 4.11 and 4.12, reducing inlet methanol reactor pressure leads to a decrease in terms of the net CO₂ emission as well as the production cost. Obviously, since the methanol production causes a decline in the number of moles, the reaction is appropriate to elevated pressures. However, the higher the inlet pressure is applied, the more electric power at the inlet and recycle compressors is required, leading to more indirect CO₂ emission. Therefore, less net CO₂ emission is seen at lower pressures. Additionally, lessening inlet reactor pressure leads to the quite noticeable reduction of production cost due to the lower electric power cost. The new pressure is selected at 55 bar to guarantee methanol productivity and purity.

Table 4.11 Relationship between the inlet methanol reactor pressure and net CO₂ emission

Pressure (bar)	Direct CO ₂ Emission (kgCO ₂ /tMeOH)	Indirect CO ₂ Emission (kgCO ₂ /tMeOH)	Net CO ₂ Emission (kgCO ₂ /tMeOH)	Methanol Productivity (kg/hr)	Methanol Purity
55	-1387.9	2819.6	1431.7	20237	0.9957
60	-1388.0	2824.3	1436.3	20241	0.9958
65	-1388.2	2830.0	1441.8	20241	0.9959
70	-1388.4	2836.3	1447.8	20237	0.9960
75.7	-1388.7	2843.8	1455.1	20230	0.9960

Table 4.12 Relationship between the inlet methanol reactor pressure and production cost

Pressure (bar)	Energy Consumption (electricity) (\$/tMeOH)	Water Consumption (\$/tMeOH)	CO ₂ Capture cost (\$/tMeOH)	H ₂ Production cost (\$/tMeOH)	Production Cost (\$/tMeOH)
55	16.9	1.2	51.0	573.3	642.4
60	17.4	1.2	51.0	573.2	642.8
65	17.9	1.2	51.0	573.2	643.3
70	18.4	1.2	51.0	573.3	644.0
75.7	19.0	1.2	51.0	573.5	644.8

4.2.3 Feed Ratio

In this case, the process cannot converge except the H₂/CO₂ ratio of 3. In fact, Zhang *et al.* (2006) found that if the H₂/CO₂ ratio increases more than 4, the selectivity of methanol will remarkably decrease caused by the formation of methane. He concluded that the H₂/CO₂ ratio of 3 is the maximum selectivity for methanol. Researches from Fornero *et al.* (2011) also demonstrated that this ratio is advantageous to the methanol synthesis.

4.2.4 Optimal Design Factors and Performance Results

After sensitivity analysis, new operating conditions of the process are presented in Table 4.13. Table 4.14 shows and compares the results the optimized case with those from the base case. Obviously, the optimized case has good competitiveness to the base case in terms of environment as well as economy. Regarding net CO₂ emission, there is a slight reduction of 2.2 percent from 1464 to 1432 kgCO₂/tMeOH. The process witnesses a decrease of 0.71 MM\$/year in an amount of production cost of the optimized case as compared to the base case while the most dramatic decrease is seen in the capital cost of the optimized case with near 1.2 MM\$.

Table 4.13 Comparison of operating conditions between the optimized and base case

Specification	Base Case	Optimized Case
H ₂ /CO ₂	3	3
Inlet reactor temperature (°C)	215	200
Inlet reactor pressure (bar)	75.7	55

Table 4.14 Comparison of environmental and economic aspects between the optimized and base case

Specification	Base Case	Optimized Case
Net CO ₂ emission (kgCO ₂ /tMeOH)	1464.3	1431.7
Capital cost (MM\$)	55.46	54.26
Production cost (MM\$/year)	128.29	127.58

Although the optimized case achieves better results than the base case in both CO₂ emission and economic aspects, the process is still far from our targets which require negative net CO₂ emission as well as the obtainable profits. Figure 4.12 shows that the new case is also impossible to attract the interest from investors.

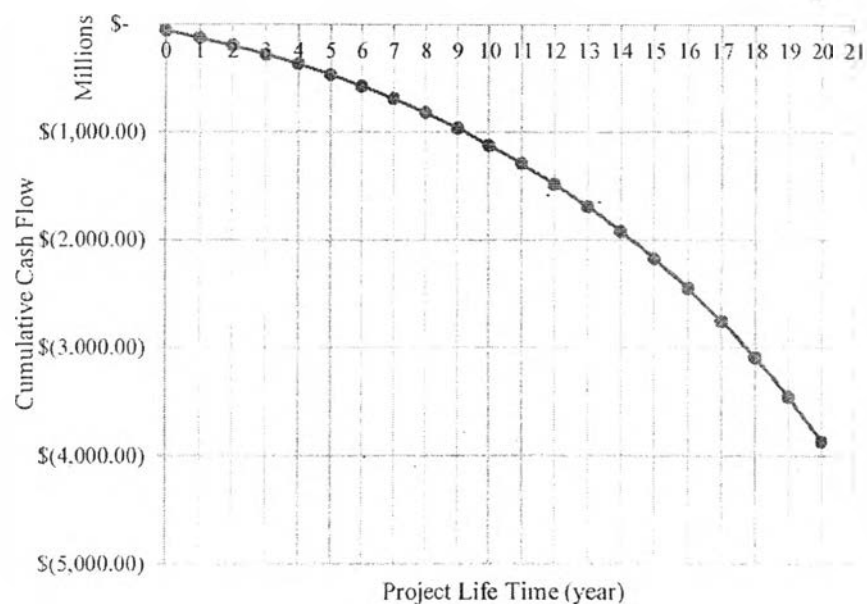


Figure 4.12 Cumulative cash flow for 20 year project of the optimized case.

4.3 Alternative Design Ideas

From the results of sustainability analysis and economic evaluation, alternative idea is to employ renewable sources for producing hydrogen to reduce net CO₂ emission and production cost.

In near future, steam reforming still plays an important role to produce hydrogen. However, when the supply of fossil fuel reduces, alternative resources have to be developed and utilized to produce hydrogen for both environmental and economic aspects. Rapidly increasing the cost of fossil fuels as well as reducing in the cost of alternatives due to technological improvement and adequate plant sizes will promote the usage of alternative energy sources.

Taking hydrogen from water is the available method to convert alternative energy into hydrogen. When either electricity and electrolysis or heat and a thermochemical process are used, hydrogen and oxygen can be broken apart from water by alternative energy sources (Bartels *et al.*, 2010). In addition, using biomass as the source to produce hydrogen from thermochemical and biological techniques is the focus in recent researches.

No carbon is released during the production and usage of the hydrogen fuel if no carbon compounds are emitted by employing alternative energy sources. This is impossible to produce hydrogen from fossil fuels because large amounts of carbon are being discharged during both production and energy consumption.

In this research, five alternatives to produce hydrogen are studied, including wind/electric, nuclear/steam electrolysis, solar thermal, biomass and hydroelectric. While wind/electric, nuclear/steam electrolysis, solar thermal and hydroelectric are the indirect ways to produce hydrogen through electricity generation before employed in a water breaking process, biomass resources are able to generate hydrogen directly by pyrolysis and gasification processes.

4.3.1 Performance Results

As shown in Figure 4.13, all alternatives that produce hydrogen from renewable energies have the negative net CO₂ emission. It is evident that renewable energy-based hydrogen production shows a competitive CO₂ reduction. Three technologies including wind/electric, nuclear/steam electrolysis and hydroelectric have lowest CO₂ production with around -734.5, -575.5 and -555.6 kgCO₂/tMeOH, respectively. Hydrogen production from steam reforming is still the problem in terms of sustainability although the methanol synthesis has a negative net CO₂ emission.

Figure 4.14 shows the comparison of capital and production cost of each alternative. Because hydrogen is bought outside the plant, this factor only affects the production cost. All scenarios, therefore, have the same the capital cost. Obviously, the production cost from alternative energies is higher than that from natural gas except biomass and hydroelectric. The production cost with steam reforming is around 127.58 MM\$/year. Meanwhile, the lowest cost comes from hydroelectric with approximately

70.92 MMS\$/year, that is nearly a half of steam reforming price whereas the cost from biomass is 35 percent lower than that from natural gas.

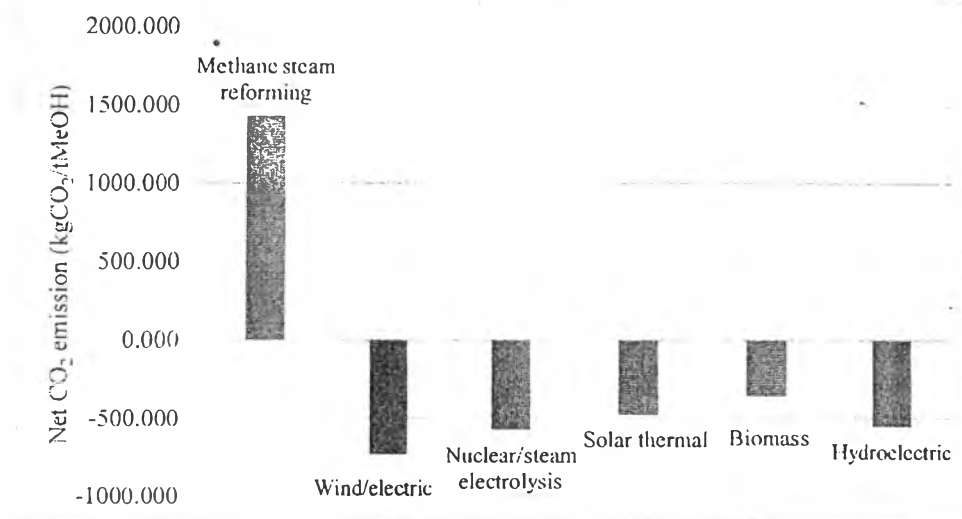


Figure 4.13 Net CO₂ emission for the total process of each scenario for hydrogen production.

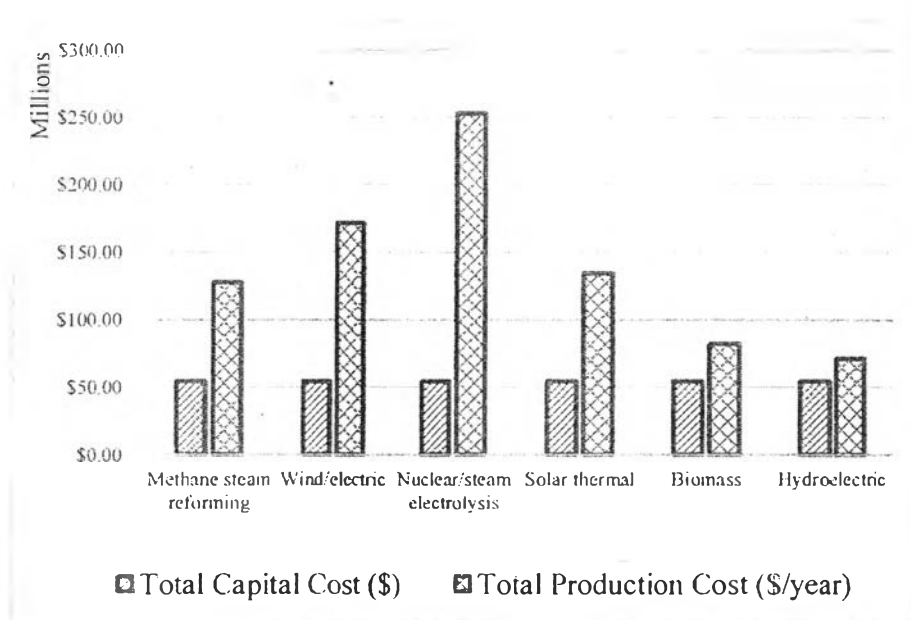


Figure 4.14 Comparison of the total capital cost and total production cost of each scenario for hydrogen production.

Undeniably, the production cost plays the key role in determining which design has the potential to invest. This can be clearly seen through evaluation of net present value (NPV) and internal rate of return (IRR) as in Figures 4.15 and 4.16.

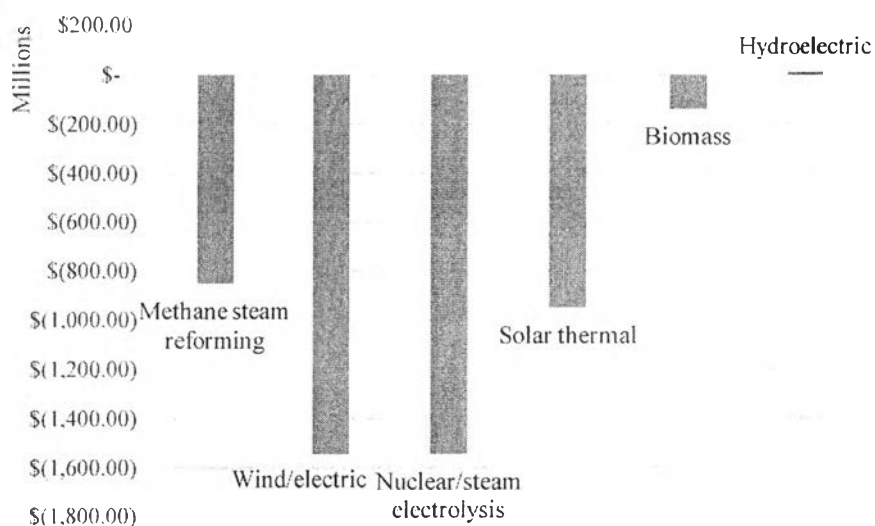


Figure 4.14: Comparison of NPV of each scenario for hydrogen production for 20 years life time.

According to the results from Figures 4.15 and 4.16, the positive NPV or IRR means those designs were preferred for investment. As discussed above, the production cost has the principal effect on investment decisions. Clearly, every alternative with the higher production cost than the base case has the negative NPV and IRR such as wind/electric, nuclear/steam electrolysis and solar thermal. Interestingly, although having the lower production cost, biomass has the negative NPV and IRR. The reason is that the production cost is still higher than the profitability from selling product. Only hydroelectric obtains both high positive values for NPV and IRR because it has the reduced production cost compared to that from the base case and selling methanol brings more money than operating the process. From the result, hydroelectric has the NPV of 14.69 MM\$ and IRR of 31 percent. As pointed out in Figure 4.17, the breakeven point of hydroelectric is 6.5 years, which shows that

the project will bring profit after long operation period. However, it is still seem as the potential process to invest in terms of economy.

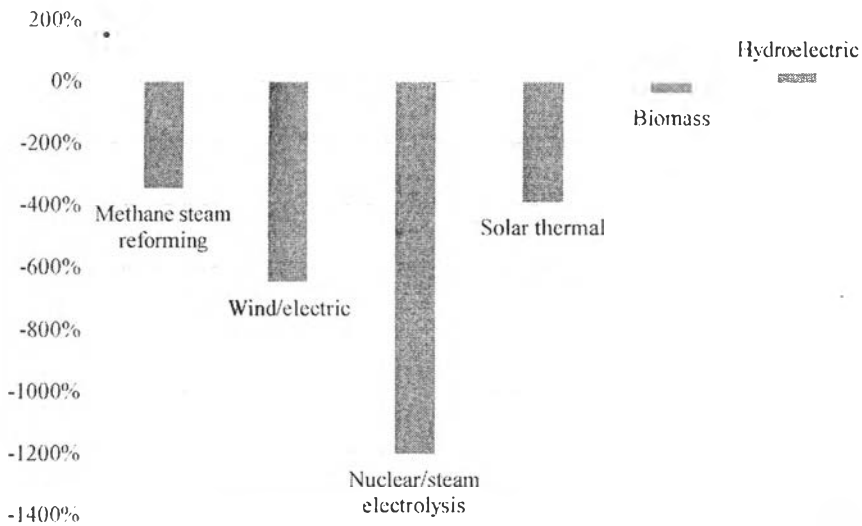


Figure 4.16 Comparison of IRR of each scenario for hydrogen production for 20 years life time.

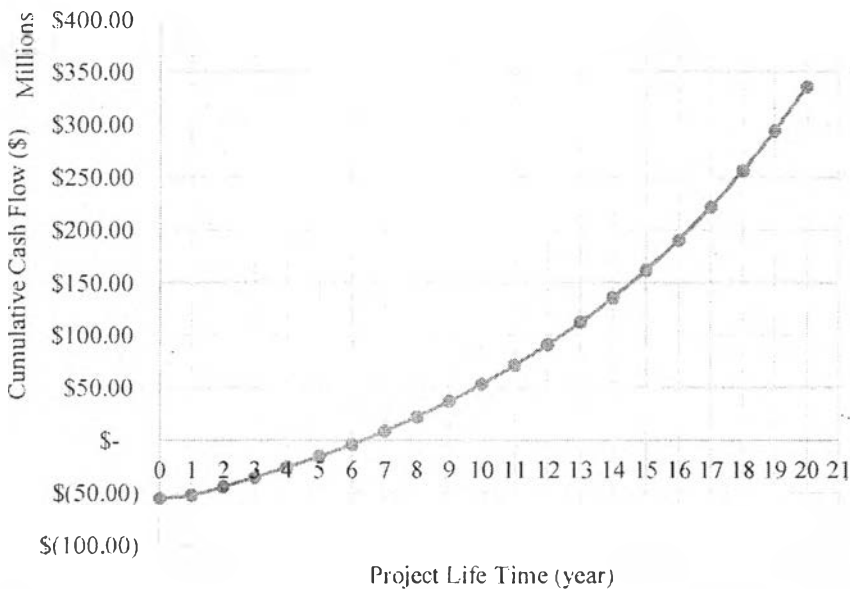


Figure 4.17 Cumulative cash flow for 20 year project of hydroelectric.