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Simulation and Techno-economic Analysis of a CO_2 Capture Scheme for Combustion Processes

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Abstract. Literature search has revealed that most post-combustion CO_2 capture processes available to power plants are hinged on sorption technologies (chemical absorption and desorption) which are cost and energy-intensive. The colossal cost of solvents used for the absorption of CO_2, the high energy requirements for the absorption & solvent recovery processes, plant design & control related complications, and environmental concerns about the solvents are some of the challenges associated with this option of post-combustion CO_2 capture approach which is widely being adopted to mitigate CO_2 emissions from power plants. In this study, the capture of CO_2 from flue gas by exploring the concept of change of state of matter, which is essentially a physical approach to CO_2capture, is presented as an alternative to the typical methods in the industry.

Keywords: Fuel combustion, anthropogenic emissions, climatic change, physical properties,dry ice, cost effectiveness

1. Introduction

As energy is continuously being consumed to support industrialization, economic development, and modern life, the resulting large volume of CO_2 emission is leading to climatic changes, and it is of great concern to environmentalists and policymakers[1]–[4]. Mitigation of global climatic change through the reduction of global CO_2 emissions has, therefore, become a shared interest of international communities[5], [6]. With the increased awareness and interest in lowering CO_2 emissions, several technologies have and are still emerging for abating CO_2 emissions[7]–[12]. Each of the technologies can be assessed based on emission reduction potential, development potential for different CO_2 emission sources and the economic implications[13]–[16]. The current technologies for CO_2 emission reduction may be broadly classified into three. The first,

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being technologies that enhance the efficiency of energy conversion, thereby reducing energy consumption and the attendant CO_2 emission. This category of technologies includes advanced

power generation technologies such as integrated gasification combined cycle (IGCC) and natural gas combined cycle (NGCC)[14], [15]. Technologies that develop low – carbon fossil fuels or adopt new energy and renewable energy sources are also being exploited and gaining much relevance. This category of technology includes geothermal power systems, solar power systems, nuclear energy, and hybrid energy systems[17]–[26]. The third category of technology for abating CO_2 emission involves separating, capturing and storing CO_2 emanating from the consumption of fossil fuels[27]–[29]. All of these technologies have received broad attention from academia and policymakers [30], [31].

Notably, adopting some technologies that fall within the first two broad categories of CO_2 emission reduction technologies in energy conversion systems require a significant overhaul of existing technologies. Since substantial overhaul of existing energy conversion systems or complete switch to newer and more efficient technologies is capital intensive and renewable sources of energy may not be able to meet the increasing demand for energy in short to medium terms, it is, therefore, pertinent to concert effort on developing and utilizing abatement technologies that precondition fuels to limit CO_2 emission from combustion (pre-combustion technologies), condition flue gas to meet environmental standards (post-combustion technologies) and thereby complementing improved efficiency efforts in the combustion of the fuels. These will help in the considerable reduction of CO_2 emission while necessary financial muscle and technical know-how are being acquired for the adoption of newer technologiesandduring the transition period from the use of fossil energy sources and renewable ones.

Currently, most Post-combustion CO_2Capture (PCC) processes in the power industry are based on sorption principles and the most widely applied is the absorption of CO_2 in amine solutions which is saddled with some drawbacks[27], [32]–[34]. Substantial capture of CO_2 from the flue gas from a power plant using amine absorption technology requires large amounts of energy, mostly in the form of heat. Sizing a full-scale carbon capture plant poses unique challenges to constructability, operability, and maintenance, particularly for PCC processes and plants using chemical absorption and desorption as process principles[35]. For such processes, it has been documented that technological upscaling risks can be minimized through process design supported by laboratory experiments [36], [37].In addition to that, Gardarsdottir et al. [38], reported that implementation of control strategies on mono-ethanol amine (MEA) based post-combustion CO_2 capture system (which is a variant of amine absorption technology) can result in better capture system performance with respect to the capture efficiency and that the CO_2 capture cycle responded to load changes within a reasonable time in PCC fitted power plant. These notwithstanding, the need to develop other alternative methods for CO_2 capture should not be undermined. Research efforts have to be tailored towards coming up with alternative CO_2 capture technologies that are cost-effective, less energy-intensive and result in value-added products from the flue gas. There are power cycles that make use of energy in waste heat and low-grade heat sources. The potentials of CO_2 as working fluid in such power cycles appear to be great[39]–[41]. It is possible to use dry ice for cooling and in the future powerless

refrigeration.Dry icecan be used for quenching fires including wild fires and can be used in all applications wherever CO_2 gas is used.It is also imperative to point out that dry ice which may be termed as multipurpose in nature can be cheaply obtained in abundance, by taking the advantage of the relative abundance of

CO_2 in combustion flue gases. Hence,the focus of this research is to simulateand evaluate a scheme whichcaptures carbon dioxide in solid form by exploiting the physical properties of flue gas. This is to ensure that while reducing emission of CO_2 into the environment fromcombustion units (such as from the boiler in power plants), the CO_2 can alsobe made available in a form that can be used in relevant power cycles and other important area of applications; thereby making abatement of CO_2emission economically rewarding and attractive to power plant owners and other operators of combustion processes.

2. Methodology

In this section, the materials used, data source and process description for this study are presented.

2.1. Materials

The primary research tool for this work was HYSYS 8.8,which was used for process simulation and EXCEL used for data evaluation.

2.2. Data

In this study, data that typifies the boiler section of a power plant in Nigeria is considered. 858240.5kg/hof flue gas from the combustion of 50190 kg/h of natural gas (composition is as given in Table 1) at an air-fuel ratio of 16.1 was used as feed into the CO_2 capture scheme simulated using HYSYS. The fuel combustion (also simulated using HYSYS) is to generate the required heat to raise the enthalpy of feed water and reheat steam for the thermal power plant which is designed to produce approximately 220 MW of electricity at full load. The flue gas which exits the boiler (where combustion of fuel and heat exchange between the combustion gas andthe feed water along with the reheat steam takes place), having raised the enthalpy of 653902kg/h of feed water (from 907.7449 kJ/kgto 3342.762 kJ/kg) and 585452.3 kg/h of cold reheat steam (from 3009.88 kJ/kgto3442.81 kJ/kg) leaves at a temperature and pressure of 475.26 ℃ and 241 kPa respectively. The composition of the flue gas isas given in Table 2. The Peng Robinson fluid package in HYSYS 8.8was used to obtain necessary and reasonably accurate thermodynamic data on the combustion and CO_2 capture processes.

Table 1: Natural gas composition

Component	CH_4	C_2H_6	C_3H_8	CO_2	N_2
Mole fraction	0.894	0.086	0.004	0.006	0.010

Table 2: Flue gas composition

Component	CH_4	C_2H_6	C_3H_8	CO_2	N_2	H_2O	CO	NO_2	NO	N_2O	O_2
Mass composition [%]	2E-14	1.04E-29	7.49E-37	13.83445	72.24742	12.30878	1.255882	3.33E-05	0.144865	1.26E-05	0.208561

2.3. Process Description

Before conditioning the flue gas to obtain dry ice in the proposed CO_2 capture scheme, the flue gas is allowed to cool naturally.The presence of water vapourin the flue gas necessitates its natural cooling to a suitable temperature for water to condense out. After the flue gas cooling and water removal by condensation, the phase change property of CO_2 was harnessed for the production of dry ice from the 'dried' flue gas as follows: The 'dried' flue gas was taken through multi-staged compression and cooling[42]. The first three stages of cooling were achieved using cooling utility while the final stage cooling was performed using a portion of the cold recycled exhaust of anisentropic expander. The separation of CO_2rich stream from the compressed and cooled flue gas was done having cooled the stream from the final stage of compression, leaving behind theCO_2 deprived flue gas stream. Beforethe isentropic expansion of the CO_2 deprived flue gas stream to atmospheric pressure, itwas cooled using a fraction of therecycled exhaust ofthe expander. The free expansion of the CO_2 richstream leaving the phase separator was carried out to atmospheric pressure.To obtain a super cooled dry ice, a fraction of the isentropic expander exhaust was used to cool the CO_2 rich stream which was freely expandedbefore its collection atthe bottom of a flash chamber and storage in cryogenic vessels[42].The process flow diagram of the CO_2 capture process, which was generated from the simulation using HYSYS 8.8 is presented in Figure 1.

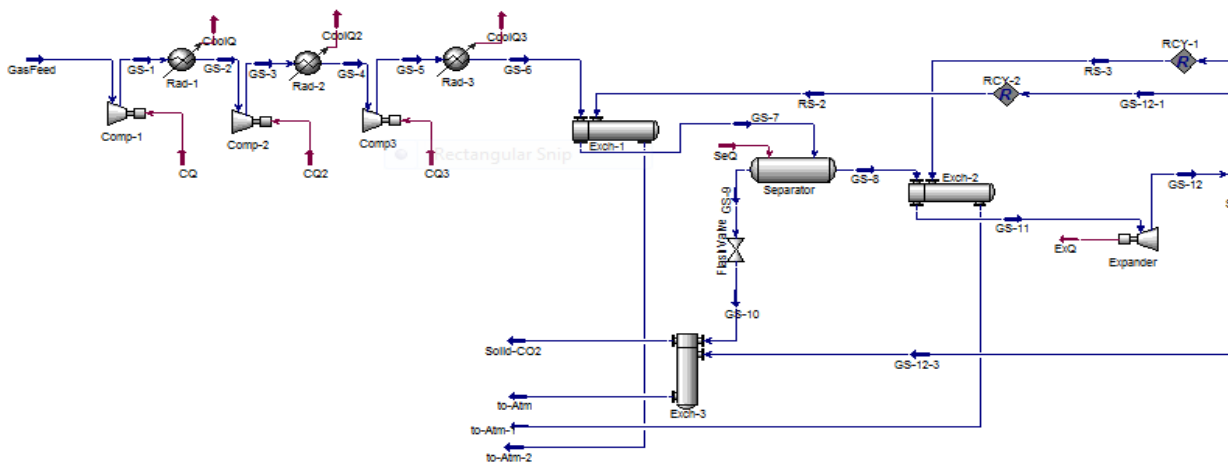

Figure 1: Process flow diagram of the proposed CO_2capture process obtained using HYSY

2.4. Economic Evaluation

To investigate the economic viability of the CO_2 capture scheme, the purchase costs of components of the plant were obtained from manufacturers. Where necessary, the equipment cost data obtained was updated using a cost index as given in equation (1).The total investment associated with setting up the CO_2 capture plant for dry ice production (which includes costs of assembly, piping, instrumentation &control, electrical installation and civil construction) and the projected maintenance and operation costs for the plant was estimated using the method adopted by[43].

$$Present\ cost = original\ cost\left(\frac{index\ value\ at\ present}{index\ value\ at\ time\ original\ cost\ was\ obtained}\right) \quad (1)$$

The total cost of investment was then amortized by adopting a useful life period of 15 years and an annual interest rate of 12% as given is equation (2)

$$A = Inv\left[\frac{j(1+j)^N}{(1+j)^N - 1}\right] \quad (2)$$

where A, annuity [US$/year]; Inv, investment value; j, interest rate; N, number of useful years.

With the assumption of 8600 operating hours annually, the hourly cost of owning, maintaining and operating each component of the plant was evaluated using equation (3)

$$Z\left({}^{US\$}/_{h}\right) = (amortized\ capital\ cost + operating\ \&maintenance\ \ cost)\left[{}^{US\$}/_{yr}\right]\frac{1yr}{8600hr} \quad (3)$$

The carbon tax associated with an unattended release of CO_2 into the environment through the combustion process is computed at the rate of 8$ per tonne CO2 equivalent[44]. Revenue accruable from the sale of dry ice is put at 1.1013$/kg[45]. The costing of energy used in achieving the production of the dry ice is done at the rate of 0.117004 $/kWh[46][47].

3. Results

Figures 2-5 show the effect of cooling temperature on the condensation of the flue gas constituent before the CO_2 capture process

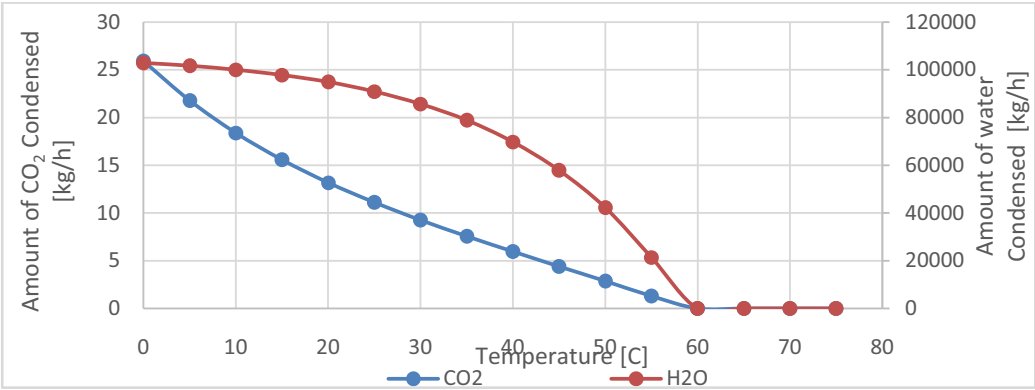

Figure 2: Effect of temperature on the removal of H_2O and CO_2 by condensation

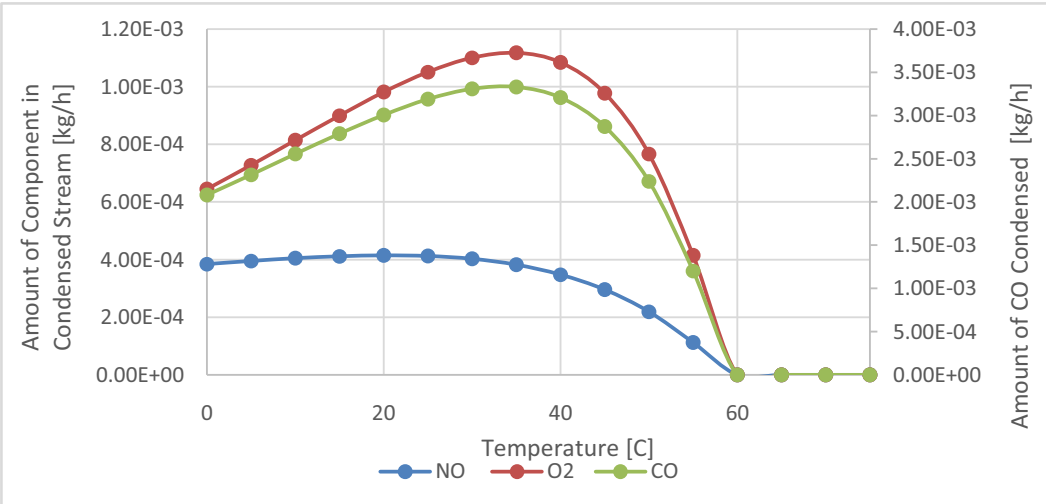

Figure 3: Effect of temperature on removal of NO, O_2, and CO by condensation

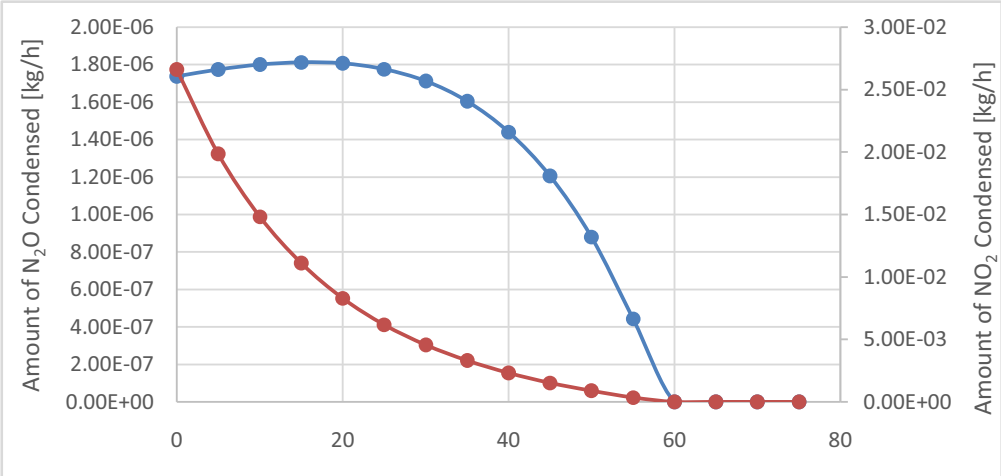

Figure 4: Effect of temperature on the removal of N_2O and NO_2 by condensation

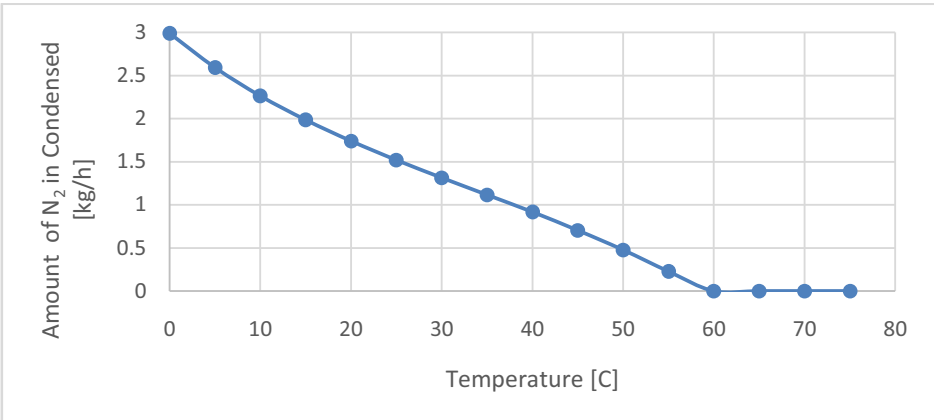

Figure 5: Effect of temperature on the removal of N_2 by condensation

Tables 3 and 4 give the conditions and compositions of streams in the simulated CO_2 capture process. The energy requirement for achieving the production of dry ice is presented in Table

5.Table 6 show the breakdown ofthe installation, Operational, and maintenance of CO_2 capture plant. The summary of all costs associated with dry ice production and the projected revenue from its sale is presented in Table 7

Table 3: Thermodynamic data of streams in the simulated CO_2 capture from natural gas flue

Stream	T [°C]	P [kPa]	Molar Flow [kgmole/h]	Mass Flow [kg/h]	Entropy Change [kJ/kg.°C]	Enthalpy Change [kJ/kg]	Specific Physical Energy [kJ/kg]
GasFeed	25.00	101.32	25314.35	752588.91	0.00000	0.00	0.00
GS-1	114.28	212.78	25314.35	752588.91	0.06075	91.44	73.33
GS-2	25.00	212.78	25314.35	752588.91	-0.20852	-0.38	61.79
GS-3	73.72	324.24	25314.35	752588.91	-0.17215	49.20	100.53
GS-4	25.00	324.24	25314.35	752588.91	-0.32735	-0.76	96.84
GS-5	81.61	526.89	25314.35	752588.91	-0.28591	56.87	142.11
GS-6	25.00	526.89	25314.35	752588.91	-0.46499	-1.46	137.18
GS-7	-45.00	526.89	25314.35	752588.91	-0.73773	-72.81	147.14
RS-2	-174.67	101.30	18340.18	515111.82	-1.15877	-207.79	137.69
to-Atm-2	-75.31	101.32	18340.18	515111.82	-0.42310	-103.54	22.61
GS-8	-135.00	526.89	22672.14	636781.37	-1.29899	-168.97	218.33
GS-9	-135.00	526.89	2642.21	115807.54	-2.98771	-560.84	329.94
GS-10	-135.06	101.32	2642.21	115807.54	-2.98399	-560.84	328.83
GS-11	-135.00	526.89	22672.14	636781.37	-1.29903	-168.97	218.33
RS-3	-174.67	101.30	2165.98	60834.78	-1.15877	-207.79	137.69
to-Atm-1	-172.97	202.65	2165.98	60834.78	-1.35623	-207.74	196.62
GS-12	-174.67	101.30	22672.14	636781.37	-1.15877	-207.79	137.69
GS-12-1	-174.67	101.30	18340.18	515111.82	-1.15877	-207.79	137.69
GS-12-2	-174.67	101.30	2165.98	60834.78	-1.15877	-207.79	137.69
GS-12-3	-174.67	101.30	2165.98	60834.78	-1.15877	-207.79	137.69
Solid-CO2	-138.60	101.32	2642.21	115807.54	-3.02597	-566.57	335.63
to-Atm	-164.24	101.32	2165.98	60834.78	-1.05372	-196.90	117.26

Table 4: Mass composition of streams in the simulated CO_2capture from natural gas flue

Stream	CO_2	O_2	N_2	CO	NO_2	NO	N_2O
GasFeed	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-1	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-2	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-3	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-4	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-5	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-6	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
GS-7	0.1578	0.0024	0.8239	0.0143	3.72E-07	0.00165	1.44E-07
RS-2	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
to-Atm-2	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
GS-8	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
GS-9	0.9926	0.0000	0.0063	0.0002	2.42E-06	0.00088	8.95E-07
GS-10	0.9926	0.0000	0.0063	0.0002	2.42E-06	0.00088	8.95E-07
GS-11	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
RS-3	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
to-Atm-1	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
GS-12	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
GS-12-1	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
GS-12-2	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
GS-12-3	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09
Solid-CO2	0.9926	0.0000	0.0063	0.0002	2.42E-06	0.00088	8.95E-07
to-Atm	0.0059	0.0028	0.9726	0.0169	2.10E-14	0.00179	7.40E-09

Table 5 : Energy requirement for dry ice production

Equipment	Energy requirement/Production [MW]
First Stage Compressor Work	19.116
First Stage Cooling Utility	19.196
Second Stage Compressor Work	10.365
Second Stage Cooling Utility	10.445
Third Stage Compressor Work	12.048
Third Stage Cooling Utility	12.193
Expander Work	6.867
Net Requirement	**76.495 MW**

Table 6: Cost of installation, operation, and maintenance of CO_2 capture plant

Equipment	Purchase Equipment Cost ($)	Total Equipment Cost ($)	Levelized Purchase Equipment Cost ($/year)	Levelized Purchase Equipment Cost ($/h)	Levelized Operation and Maintenance Cost ($/h)	Total Cost ($/h)
Desiccant dryer	240514.99	399254.8834	50904.9976	6.31221971	0.1893666	6.50159
Compressor 1	47795840	79341094.4	10115987.14	1255.08525	37.6525576	1292.737811
Compressor 2	28115200	46671232	5950580.671	738.285443	22.14856329	760.4340063
Compressor 3	16602900	27560814	3514002.953	435.980515	13.07941546	449.0599307
Cooler 1	2365700	3927062	500700.2864	62.1216236	1.863648709	63.98527234
Cooler 2	2617800	4345548	554057.2388	68.7415929	2.062247787	70.80384069
Cooler 3	1389200	2306072	294024.1104	36.4794182	1.094382545	37.57380071
Heat Exchanger 1	147500	244850	31218.36761	3.8732466	0.116197398	3.989444
Heat Exchanger 2	7700	12782	1629.704614	0.2021966	0.006065898	0.2082625
Heat Exchanger 3	8600	14276	1820.189569	0.22582997	0.006774899	0.23260487
Separator	165300	274398	34985.73672	4.34066212	0.130219864	4.470881987
Flow Splitter Valve	797	1323.02	168.6850101	0.02092866	0.00062786	0.021556521
Expander	22359050	37116023	4732291.812	587.132979	17.61398937	604.7469685
Flash Chamber	80000	132800	16932	2.099568	0.062987	2.16256
Flash Valve	732	1215.12	154.9277633	0.01922181	0.000576654	0.019798461
Pelletizer	40000	66400	8466	1.049784	0.0314935	1.08128
		TOTAL				**3298.03**

Table 7 : Cost implication of dry ice production from the power plant flue gas

Amount of CO_2 Captured [kg/h]	Hourly Cost of Owning the Plant ($/h)	Cost of Electricity Used ($/h)	Carbon Tax Averted ($/h)	Probable Revenue ($/h)
114950.6	3298.03	8950.24	919.60	126595.1

4. Discussion

As shown in Table 1, the natural gas used for this study is composed of 89.4 mole % methane, 8.6 mole % ethane, 0.4 mole % propane, 0.6 mole % carbon dioxide and 0.1 mole % nitrogen. In terms of mass composition, these correspond to 81.265 mass % methane, 14.653 mass % ethane,

0.999 mass % propane, 1.496 mass % carbon dioxide and 1.587 mass % nitrogenrespectively. It is highly desirable to achieve complete combustion of fuels to harness maximum heating value from the

fuel and also to reduce the contribution of organic contents of the fuel and carbon monoxide (associated with incomplete combustion)to greenhouse gas effects. Moreover, the equivalent contributions of these compounds are usually higher than that of CO_2. For instance, relative to CO_2, the global warming potential of CO, CH_4, C_2H_6, C_3H_8 is 1.9, 28.5, 1.46, 0.19, respectively, where that of CO_2 is taken to be 1 [48], [49]. However, in the bid to achieve complete combustion, it is expedient that the Air to Fuel Ratio (AFR) of combustion be regulated alongside other combustion parameters to avoid damping of the combustion gas and excessive formation of compounds such as NO and NO_2 which contribute to air smog, acidification and eutrophication. When AFR is excessively high, the temperature and the heating value of the resulting combustion gas reduce drastically and the combustion gas will not be able to provide the necessary heat required for raising the enthalpy of the feed water and cold reheat steam. In such cases where the combustion gas has been dampened through excessive air, more fuel will have to be burnt to obtain the necessary heat and this result in unsustainable depletion of the finite fuel sources, increase emissions to the environment and the cost of production.

When the combustion of natural gas was carried out at air to fuel ratio of 16.1, the resulting $858240.5\ kg/h$ combustion gas was at a temperature of 1983.15 ℃ and has a heating value of $411.44\ kJ/kg$ which was sufficient to produce the steams required for driving the turbine for electricity generation. As shown in Table 2, the flue gas leaving the boiler has a very negligible amount of methane, ethane and propane. By mass fraction, nitrogen dominates the flue gas, next to it is carbon dioxide which makes up about 13.8% of the flue gas and this is followed by water vapour which is about 12.3% of the flue gas. If the flue gas is indiscriminately discharged, carbon dioxide, which forms about 12.3% of the flue gas is equivalent to the emission of about 118732.9 kg/h of CO_2 into the environment. Such a release will contribute to aggravation of the adverse which anthropogenic greenhouse gases have on the environment and where environmental regulations are strigent, this act comes with an heavy penalty. The water vapour in the flue gas is about 105638.9 kg/h.

Using physical separation technique; the objective of this study is to obtain CO_2 (in this case, dry ice) from the flue gas that can be utilized for diverse purposes. It is, therefore, necessary to condition the flue gas to enhance the separation of CO_2 from other constituents of the flue gas. The flue gas which leaves for the stack at 475.26 ℃ was to be cooled to a temperature which facilitates the selective removal of its components by condensation. The study on the selection of an appropriate cooling temperature as presented in Figures 2-5 was instrumental in choosing the inlet temperature of the flue gas into the CO_2 capture scheme. Figure 2 reveals that at a temperature of 60 ℃ and above, very little can be achieved in terms of the removal of water vapour from the flue gas. At temperatures between $30 - 60$℃, notable amounts of water have condensed out of the flue gas. Plateau effect in the removal of water by condensation begins to set in at about 25℃, whereas the condensation of CO_2, which is not desirable in this case, continues to increase as temperature drop continues. By allowing the flue gas to cool down to

around 25℃, condensation of about 86.13%of the water vapour initially present in the flue gas have been achieved. The removal of water by condensation is to stem the possibility of CO_2 hydrate formation (during the process of CO_2 capture) at region of temperature and pressure suitable for the hydrate formation [50], [51]. This

is particularly crucial because hydrate formation can result in handling problems and corrosion in vessels [52].

As shown in figure 3, significant condensation of NO, O_2, and CO did not occur until the temperature of the flue gas dropped below 60 ℃and the decreased in the amount of these gases removed by condensation set in at 35 ℃. From figure 4, the removal of N_2O by condensation peaked at 20 ℃, whereas the removal of NO_2 by condensation continued to experience an increase well below 20 ℃. As depicted in figure 5, the removal of N_2 by condensation continued to increase at temperatures below 60 ℃ and when the temperature dropped to 0 ℃, only about 2.99 kg/h out of 620056.6 kg/h initially present was condensed out; with the nature of the graph, it can be predicted that not much will be achieved in terms of removal of N_2 by condensation even if the cooling temperature is set well below 0 ℃. Since a significant removal of water by condensation was achieved at 25 ℃ and other gases are minutely present in the flue gas, it was adjudged reasonable to peg the condensation temperature at 25 ℃. The use of desiccant dryer for removing the remaining water in the flue gas at a temperature of 25℃ may be considered to limit the loss of CO_2 by condensation caused by cooling the flue gas beyond 25℃ if it can be economically justified. After the cooling/condensation stage, the feed flue gas into the CO_2 capture scheme was at the rate of 752588.91 kg/h, as shown in Table 3. The condensation and water removal, which took place before the process of CO_2 capture led to a change in the composition of flue gas fed into the CO_2 capture scheme. Hence the feed flue gas into the CO_2 capture scheme composed about 15.78 mass % CO_2, 82.39 mass % N_2, 1.43 mass % CO, 0.24 mass % O_2; while other components were minutely present has shown in Table 4.

As summarized in Table 3, the first stage compression of the flue gas caused its temperature and pressure to increase from 25℃ and 101.32 $kPa$to 114.28 ℃ and 212.78 kPa respectively and this required a work of 19.116 MW by the compressor as presented in Table 5. Before commencing the second stage compression, the temperature of the flue gas was cooled back to 25 ℃ using a cooling utility with an energy requirement of 19.196 MW. The temperature and pressure changes in the second stage compression was from 25 to 73.72 ℃ and 212.78 to 324.24 kPa. For the third stage compression, the temperature increased from 25 to 81.61 ℃, while pressure changed from 324.24 to 526.89 kPa. The work requirement for the second and third stages of compression & cooling are also presented in Table 5. Based on the phases present in the compressed flue gas which was cooled using the first heat exchanger, the separator partitioned the stream into CO_2 rich stream and CO_2 deprived stream with a flowrate of 115807.54 and 636781.37 kg/h respectively. Stream analysis using the CO_2 freeze-out option in HYSYS 8.8 indicated the presence of solid in the CO_2 rich stream leaving the separator. The separation process released about 32.71 MW cold energy which may be utilized in meeting some of the cooling requirements in the cooling stages. The expander used for the expansion of the CO_2 deprived stream (before being used as the cold stream for the heat exchangers) produced

work of approximately 6.867 MW. This work can be used to meet some of the work requirements in the compression stages. From this CO_2 capture scheme, the supercooled dry ice which was obtained after further cooling by a fraction of the exhaust from the expander was found to be about 99.26 % pure. The CO_2 deprived effluent, which leaves the capture system at the rate of $636781.4\ kg/h$ is approximately 97.26% N_2, and CO_2

makes up only about 0.59% of the stream (Table 4). From Table 5, it is seen that the net energy requirement for this process was 76.495 MW if the expander were to supply some of the work required by the compressors. The cost of owning the equipment for producing the dry ice is around 3298.03 dollars per hour, and the utility cost was estimated to be about 8950.24 dollars per hour (Table 6). In Nigeria where the carbon tax is modestly put at 8 dollars per tonneCO2 equivalent, a carbon tax of 919.60 dollars per hour can be averted by converting the CO_2 that could have been emitted as waste into the environment to valuable dry ice which is 99.26% pure at the rate of 115807.54 kg/h. The carbon tax avoided can be way higher than 919.60 dollars for countries where there are stringent penalties on greenhouse gas emissions. As summarized in Table 7, if the sale of dry ice is put at 1.1013 dollars per kilogram, a gross revenue of approximately 126595.1 dollars per hour may be expected from the dry ice production.

5. Conclusion and Recommendations

The production of dry ice with high purity level was targeted in this study which necessitated the separation of other flue gas components from CO_2. The condensation temperature has to be carefully selected toremove most of the water vapour present in the flue gas by condensation, and minimize the loss of CO_2 by condensation. Based on the condensation behaviour of the component gases making up the flue gas, a temperature of 25 ℃ was identified as suitable for the flue gas cooling stage in this study. In this study, an attempt to optimally utilize the energy within the capture system is made by partly recycling the expanded CO_2 deprived stream and using it as the cooling stream for the heat exchangers in the CO_2 system. Meeting part of the work required for the compression stage through the power generated by the expander is also put forward. In addition to this, a way to utilize the cold energy released during the separation of CO_2 rich stream from other flue gas components can be considered. The analysis carried out based on the current operating conditions in the CO_2 capture system studied indicates that the energy requirements for the production the dry ice is on the high side. In a bid to make the process more attractive, further studies can be carried out on the effect of operating conditions (temperature and pressure) on the formation of dry ice, its purity level and energy required for the multi-component system of gases.

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