CARBON CAPTURE PROCESS FROM CEMENT INDUSTRY

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ABSTRACT: *Carbon dioxide is viewed as a significant source of greenhouse gases GHG, accountable for global warming in which man-made CO_2 contributes about 63.5%. The cement business contributes around 5% of worldwide anthropogenic carbon dioxide emissions. The main objective of this study is to design, build, and operate a framework to catch CO_2 from a slipstream of vent gas from a business coal-fired cement kiln, convert that CO_2 to items having business esteem, show the economic feasibility of the CO_2 capture and conversion process, and accordingly advance the innovation to a point of status for business scale production. The advancement will also validate market prospects by offers of synthetic compounds into existing markets, and recognize chances to improve innovation execution and lessen costs at business scale. The Mineralization process is a unique process designed to remove carbon dioxide from a gaseous waste stream, convert the carbon dioxide to a mineralized carbon product, and produce valuable chemical byproducts. In this research, carbon dioxide from waste flue gas reacts with sodium hydroxide to form sodium bicarbonate. The project is designed as a multi-bubble column system. This scheme is designed to remove over 90 % of the CO_2 from the flue gas stream. The resulting products produced by the process are of high purity, and readily marketable.*

Keywords: Carbon dioxide; cement; mineralization; sodium hydroxide; energy balance; material balance.

1. INTRODUCTION

In the present air, the radiative driving from human activities is substantially more significant for present and future environmental change than the assessed radiative constraining from changes in natural processes." CO_2 has expanded from non-renewable energy source use in building warming and cooling, transportation and the production of cement and various products. Deforestation discharges CO_2 and decreases its take-up by plants. CO_2 is additionally discharged in natural procedures like the deterioration of plant matter. The compound CO_2 is presently the subject of overall consideration, just like a significant segment in the untoward grouping of ozone harming substances in the world's climate [1, 2].

Before the beginning of the mechanical transformation, CO_2 concentration was recorded at 280 portions for each million or ppm however recent measures it at 391.80 ppm (atmospheric CO_2 for December 2018), a huge increment due to man's modern activities and express dismissal for the earth. The upper safety limit for atmospheric CO_2 is 350 portions for each million (ppm). Environmental CO_2 levels have remained higher than 350 ppm since mid-1988. "A present theory hypothesizes that CO_2 controls the temperature of the earth". As stated by the theory, CO_2 controls the temperature as the CO_2 particles in the air absorb infrared radiation. The CO_2 and different gases in the air are transparent to the visible radiation that conveys the sun's vitality to the earth. Although, the earth, in turn, reradiates a significant part of the vitality in the imperceptible infrared region of the spectrum. This radiation is generally extraordinary at wavelengths near the main absorption band (13 - 17 microns) of the CO_2 range. At the point when the CO_2 concentration is adequately high, even its weaker absorption bands become successful, and more infrared radiation is absorbed. As the CO_2 cover averts its emission into space, the caught radiation heats the climate [3, 4].

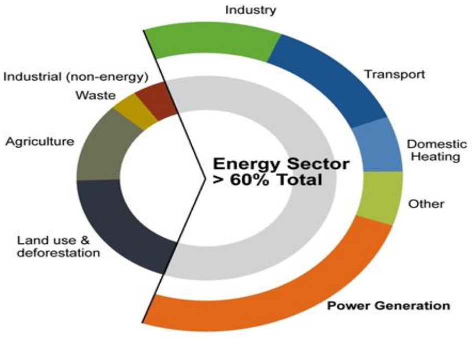

Figure 1. Total man-made emission

Cement is a significant development ingredient delivered in essentially worldwide. CO_2 is a byproduct of a chemical change process utilized in the creation of a clinker, a part of cement, wherein limestone ($CaCO_3$) is changed over to lime (CaO). CO_2 is released during the production of cement by fossil fuel burning. A cement manufacturing plant comprises of the three procedures. Raw blend making incorporates crushing, granulating, grinding and drying. Clinker production (pyro-processing) incorporates the sintering of raw mix to frame clinker on a furnace. The complete grinding process includes a succession of mixing and crushing tasks that changes the clinker to complete Portland cement. CO_2 in cement, manufacture of clinker, $CaCO_3$ is changed over to CaO and CO_2. Makeup roughly 7.0 % of worldwide carbon dioxide discharges, 1 Ton cement produces 900 to 1000 kg CO_2 [5, 6].

Table (1): The electrical energy produced by fossil fuel combustion

G8 Nation	Fossil Fuel Combustion	Total	Percentage
	Billion Tons	Billion Kwh	
United States	2,758.65	3,891.72	70.9%
Japan	640.17	982.76	65.1%
Russia	569.72	869.07	65.6%
Germany	354.78	561.57	63.2%
Italy	223.16	268.18	83.2%
Canada	154.55	569.41	27.1%
United Kingdom	278.21	373.26	74.5%
France	52.23	535.45	9.8%

The level of vitality cost in Portland concrete manufacturing is 20% - 30%. The 90% or a greater amount of fuel is expended for clinker burning. About 40% of electric power is exhausted for complete grinding, and somewhat less than 30% each is devoured by the raw material procedure and the clinker burning method. The complete grinding procedure generally expends electric power for the factory and the clinker consuming method mostly for the fan. The raw material grinding process expends a huge volume of intensity for the factory and fan. Practically all CO_2 emissions (about 96.5%) originate from petroleum products use.

Table (2): Carbon dioxide production from different sources

Process	Sources	Emissions
Fossil Fuels		(Mton CO_2/Year)
Power (Coal, Oil, Gas and Others)	4,942	10,539
Cement Production	1,175	932
Refineries	638	798
Iron and Steel Manufacturing	269	646
Petrochemical Industry	470	379
Oil and Gas Processing	N/A	50
Other Sources	90	33
BioMass Bio-Ethanol and Bio-Energy	303	91
Total	7,887	13,466

The three kinds of non-renewable energy sources that are utilized the most are coal, gaseous petrol, and oil. At the point when non-renewable energy sources are combusted, the carbon put away in them is transmitted for the most part as CO_2. Figure 2 shows the all-out man-made emanation. Every single industrialized country (except for France and Canada) get the larger part (between 60-80%) of their power from the burning of petroleum products. Table 1 is an outline for all G8 countries. In industrial production, like manufacturing, mining, construction and agriculture. Manufacturing is the biggest of the four and can be separated into five groups: paper, food, chemicals, petroleum refineries and metal/mineral products. These groups represent most by far of the vitality use and CO_2 outflows by the part. CO_2 production from various sources appears in Table 2 [7, 8].

Carbon capture and storage (CCS)

CO_2 capture and capacity or CO_2 sequestration (CCS), is the term used to portray a lot of advancements planned for catching CO_2 discharged from modern and vitality related sources before it enters the climate, compressing it, and infusing it profound underground insecure geographical developments, and ensuring it remains stored there inconclusively. Carbon catch has been being used for a considerable length of time. The gas and oil enterprises have utilized carbon catch for quite a long time as an approach to enhance gas and oil recuperation. Now, most research centers around carbon catch at non-renewable energy sources fueled vitality plants, the source of most of the man-made CO_2 discharges. Considerable power plants rely on coal to make vitality, and the consumption of coal emanates CO_2 into the climate. A few scientists visualize a future where all new power plants utilize carbon catch. Coal is right now the predominant fuel in the power division, representing around 40% of power created universally, with hydro control representing 17.5%, petroleum gas for 17.3%, atomic for 16.8%, oil for 9%, and non-hydro renewables for 1.6%. Coal is anticipated to remain the prevailing fuel for the control age in 2020 (maybe diminishing to 36%). Given dirty energy sources are essential to discover approaches to catch and cut CO_2 discharges from these sources fundamentally [9, 10].

Three distinct sorts of advancements that exist for carbon capture appears in Figure 3. Post-combustion catch is ordinarily applied to non-renewable energy source consuming force plants. The CO_2 is expelled from the pipe gas after the combustion of the petroleum derivative. Current strategies incorporate physical division for CO_2 concentrations over 10% and chemical separation for lower focuses. Pre- combustion is generally utilized in manure, synthetic, vaporous fuel, and power generation. For this situation, the petroleum derivative is somewhat oxidized and afterwards moved into CO_2 and more H_2. The H_2 can be utilized as fuel, and later CO_2 can be caught from a moderate exhaust fumes stream. The fuel is singed in O_2 rather than air and the outcomes vent gas comprises essentially CO_2 and water fume. Power plant processes dependent on oxyfuel combustion are some of the time introduced as "zero-emission" cycles on the grounds that the pipe gas stream itself is put away. The capacity of CO_2 can be either land stockpiling, sea stockpiling or mineral stockpiling. CO_2 can be geographically put away in profound springs or exhausted oil or gas fields. Now and again, CO_2 might be utilized in improved oil recuperation (a moderately experienced innovation), upgraded gas recuperation and improved coal bed methane recuperation. In sea stockpiling, CO_2 will be funneled into a water segment at a profundity of 1000m or more and breaks down in this way, or stored onto the ocean bottom as a CO_2 'lake' at a profundity more noteworthy than 3000 m. In mineral stockpiling, CO_2 will be responded with copiously available metal oxides that produce stable carbonates. This procedure usually happens over numerous years and is liable for a significant part of the surface limestone. CO_2 can be re-utilized as a feedstock for the generation of oil-rich green growth in sun oriented films to deliver plastics, transport fuel or creature nourishments, or as a feedstock in the industry for example production of carbonated refreshment [11, 12].

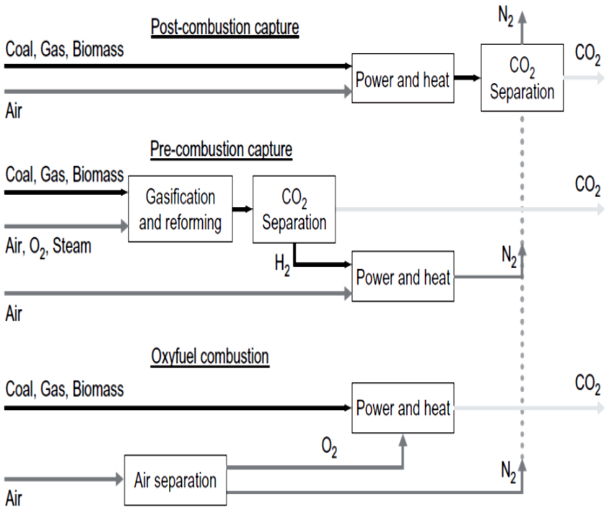

Figure 3. Technologies of capture

The catch, transport and storage procedure would build the vitality prerequisite of a plant with CCS by about 25% for a coal-terminated plant and about 15% for a gas-terminated plant. It likewise includes extra working expenses and included ventures or capital expenses. Some new advances are probably going to be more costly than develop CCS advances. CO_2 capture and storage (CCS) is the term used to depict a lot of innovations planned for catching carbon dioxide discharged from mechanical, and vitality related sources like power plants, treatment facilities, and other modern procedures like bond produce before it enters the environment, packing it, and infusing it profound underground in secure topographical developments, and guaranteeing it remains put away there inconclusively. There are three truly distinct stages in carbon catch and capacity catch and division, transportation, injection, and storage. Cement plants are large industrial sources of CO_2 emanations with high CO_2 concentration in their vent gases of around 14-33% contrasted with 12-14% CO_2 for coal-terminated power plants and 4% for gas terminated plants and hence signify a good opportunity for executing CCS [13, 14].

Pakistan has now a per capita utilization of 131 kg of cement, at 135 kg per capita for India however considerably underneath the World Average 270 kg and the regional average of over 400 kg for peers in Asia and over 600 kg in the Middle East. Concrete interest continued festered during 90's inferable from the absence of advancement activities. In 1997, per capita utilization was 73 kg in Pakistan and India. By 2018-19, utilization in India rose to 115 kg/capita. A comparison for some countries is appeared in Table 3 [15, 16].

Table (3): Comparison of few countries cement consumption

Country	Kg/Capita
Bangladesh	50
Pakistan	117
India	115
USA	375
Iran	470
Malaysia	530
EU	560
China	625
UAE	1095

According to a theory, CO_2 manages the temperature of the earth because infrared radiation is more concentrated at the wavelength close is exceptionally close of the primary assimilation band of the CO_2 spectrum a greater amount to absorb and warms up the atmosphere. The temperature has already risen 0.8 ^oC, and it will increase to about 2-5 ^oC by the end of this century. Atomic and renewable generation will have a significant impact in the vitality blend but since of the idea of this generation, it will take a very long time to get completely on the web. In any event, when a new form is online, it won't relieve the discharges of existing non-renewable energy source copying plants that are anticipated to be online for a long time to come. According to IPCC (International Panel for Climate Change), we have to reduce CO_2 emissions up to 50% and avoid the peak limit of 550ppm by 2050 to avoid the worst. In the meantime, carbon capture and storage is the only solution left to put a dent in the CO_2 levels [17, 18].

2. METHODOLOGY

The plant consists of four areas flue gas preparation and processing (Area 100 m^2), caustic preparation (Area 200 m^2), CO_2 absorption columns (Area 300 m^2) and solid separation (Area 400 m^2). Flue gas from the cement kiln is the carbon dioxide source for sodium bicarbonate. The purpose of the flue gas preparation area is to acquire the required amount of flue gas from the kiln and bring it to the required pressure and temperature for use in downstream processing. Cooling the flue gas to the required 35°C is accomplished through a series of shell and tube heat exchangers. Flue gas preparation area consists of the heat exchangers.

The heat exchanger (E-101) flue gas at the temperature of 300°C and pressure 1 atm to go in the shell of the heat exchanger and leaves at 140°C. Flue gas is cooled by process water which goes into the tubes at 35°C and leaving at 70°C. The heat exchanger is two shell-four passes. The heat exchanger (E-102) flue gas at the temperature of 140°C and pressure 1 atm to go in the shell of the heat exchanger and exit at 40°C. Flue gas is cooled by process water which moves in the tubes at 35°C and leaves at 70°C. Blower (B-103) centrifugal blower then vents off the remaining 11% clean flue gas from the bi-carbonation tower. A blower is used to build up pressure to overcome pressure drop and liquid level in the columns.

Mixer (MX-201) to (MX-204) four batch mixers are installed in which two mixers each of 19000 gal work at the same time. The mixers are operated such that continuous feed is provided to the process. The NaOH required for the process is 98.5 m^3. The level maintained in the mixer is 75%. The time required to make the solution is 26 seconds. Mixers also work as storage tanks for the process. The sodium hydroxide 50 wt. % enters the mixer at 35°C and is mixed with water to get a 4 wt. % solution of NaOH is displayed in Figure 4.

The caustic feed pump provides a motive force to move the caustic soda solution of the mixers and into the bubble columns. The pump is sized for 1750 gpm. The pump is variable speed to allow the caustic feed rate to be adjusted when the plant is running at less than full capacity. While cast iron or steel centrifugal pumps may be utilized for caustic soda solution, such pumps undergo high maintenance and small service life, for ideal service, alloy 20 is preferred. The pump is located at ground level will lift the caustic soda solution from the mixer. The pump will be self-priming to account for any gas in the system. To assist in priming the pump and in clean-out, de-ionized water can be used to flood the pipeline at the inlet of the pump. Seamless carbon steel, butt-welded schedule 40 pipes is suggested, the flanged pipe has also been used. All piping are fitted above ground.

Once the flue gas has been cooled to 35°C in the flue gas preparation area, the total flue gas is passed through a pair of CO_2 bubble columns operating in parallel. The columns provide a large contact area between flue gas bubbling upward through the column and 4% NaOH solution maintained in the column. Here the CO_2 in the flue gas reacts in the following reaction to form sodium carbonate [19, 20]:

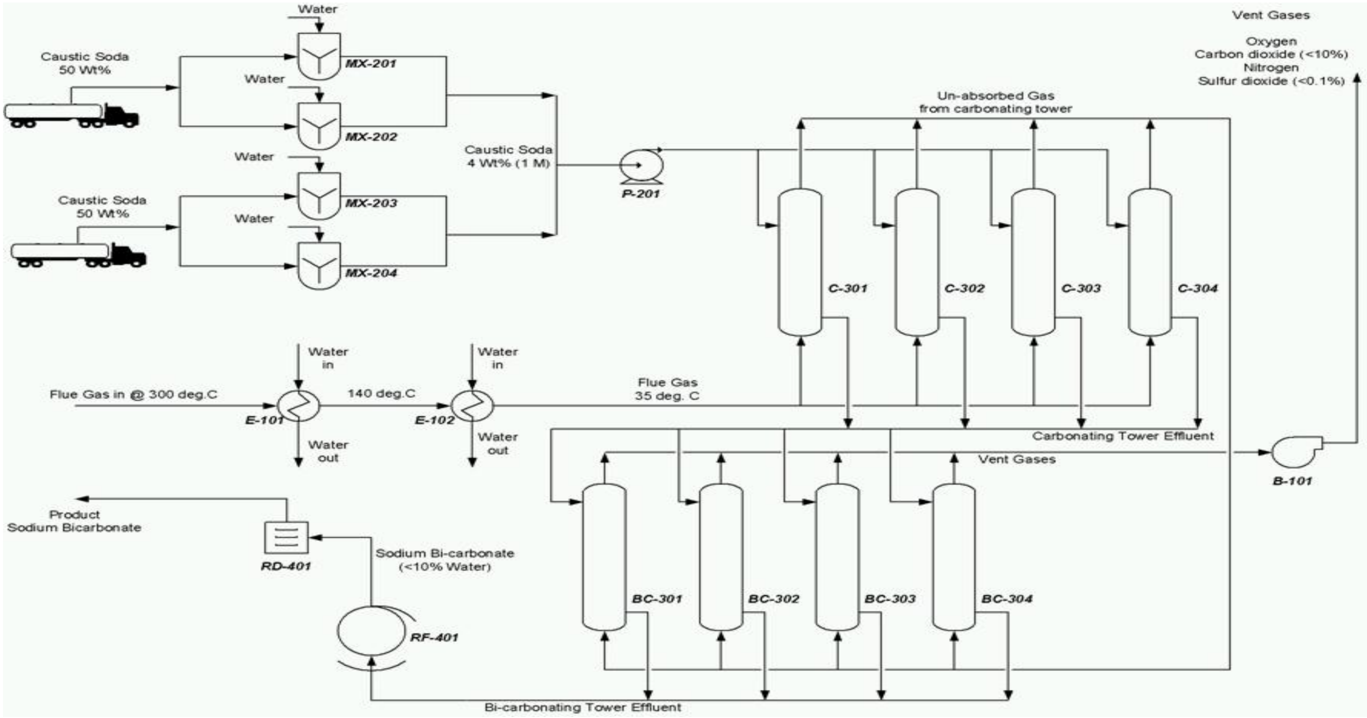

Figure 4. Carbon capture process flow

$CO_{2(g)}+2NaOH_{(l)}\rightarrow Na_2CO_3+H_2O$ (1)

The sodium carbonate solution is then fed to the next bubble columns where it is further reacted with flue gas to yield sodium bicarbonate ($NaHCO_3$).

$Na_2CO_3+H_2O_{(l)}+CO_{2(g)}\rightarrow 2NaHCO_3$ (2)

Major equipment in the CO_2 absorber column area consists of the carbonating and bi-carbonating column. Carbonating column (C-301), (C-302), (C-303) and (C-104) the flue gas bubbled from the spargers at the bottom of the bubble column. In this column, 45% of the CO_2 is converted to Na_2CO_3 and Water. As the flow rate of CO_2 is very high, so four carbonation towers are employed to distribute the flow of CO_2. Low temperature and high-pressure conditions are maintained to favour the absorption of CO_2. The Carbonating towers are constructed at a height so that the product flows to the bi-carbonator by gravity. Bi-carbonating column (BC-301), (BC-302), (BC-303), (BC-304) the remaining unconverted CO_2 enters the Bi-Carbonation Columns which are in series to the Carbonation Columns. Low temperature and high-pressure conditions are maintained to favour the absorption of CO_2. 55% of the remaining CO_2 is absorbed by the sodium carbonate to precipitate sodium bicarbonate as the solubility product of sodium bicarbonate is less than sodium carbonate. The remaining 11% CO_2 is vented off by the blower.

The slurry recuperated from the bubble section reactors is sent to a rotational drum channel (RF-401). The fluid to be filtered is sent to the tub underneath the drum. The drum turns through the fluid, and the vacuum sucks liquid and solids onto the drum pre-coat surface, the fluid part is "sucked" by the vacuum from the filter media inside the drum, and the filtrate siphoned away. The solids hold fast to outside of the drum, which at that point passes a blade, removing the solids and a little part of the filter media to expose a new media surface that will pass in the fluid as the drum pivots. The blade progresses consequently as the surface is evacuated. The fluid is reused to the blender.

Rotary dryer (RD-401) the product now enters the dryer, and as the dryer pivots, the material is raised by a movement of inside fins covering the internal dryer wall. When the material gets sufficiently high to move back off the blades, it drops down to the base of the dryer and goes through the hot gas stream. This gas stream can be moreover advancing to the release end from the feed end, or to the feed end from the release end. The last item containing is then sent to the silo for storage.

Material Balance

For the process of CO_2 absorption by the absorbent NaOH, we are required to evaluate material flow rates throughout the process. Material Balance, overall and individual, will provide us with the known values of cement production per year, coal composition and required values of flue gas composition, flue gas flow rate, absorbent (NaOH) quantity, products formed composition.

For cement production, CO_2 is produced by two reactions, which are calcination reaction and combustion of fuel used. The production of cement per year, which is for our process 0.1 MTon/year. Coal consumption and calcination CO_2 produced by the process, which is 0.143 Mton coal consumed for 1 Mton cement product. 1.32 Mton CO_2 produced for 1 Mton cement product. For our production of 0.1 Mton cement per year.

$$Coal\ Consumption = \left(\frac{0.143\ MTon\ Coal}{1\ MTon\ Cement}\right)\left(\frac{0.1\ MTon\ Cement}{year}\right)$$

$$Coal\ Consumption\ = 1.43\times10^{-2}\ Mton/year$$

And

$$\text{Calcination CO}_2\text{ Produced} = \left(\frac{1.32\text{ MTon CO}_2}{1\text{ MTon Cement}}\right)\left(\frac{0.1\text{ MTon Cement}}{\text{year}}\right)$$

$$Calcination\ CO_2\ Produced = 1.32\times10^{-2}\ Mton/year$$

A sub-bituminous coal composition C 70%, H_2O 5%, C (ash) 10%, O_2 7%, H_2 4%, S 3% and N_2 1% used in cement production. Now calculating individual flue components flow rates:

The flow rate of CO_2:

$$\dot{m}_{CO_2} = Calcination\ CO_2 + Combustion\ CO_2$$

$$\dot{m}_{CO_2} = \frac{4.18\ kg\ CO_2}{sec} + \left(\frac{44\ kg\ CO_2}{kmol}\right)\left(\frac{kmol}{12\ kg\ C}\right)\left(\frac{0.32\ kg\ C}{sec}\right)$$

$$\dot{m}_{CO_2} = 5.35\ kg/sec$$

The flow rate of H_2O:

$$\dot{m}_{H_2O} = Combustion\ H_2O$$

$$\dot{m}_{H_2O} = \left(\frac{0.02\ kg\ H_2}{sec}\right)\left(\frac{18\ kg\ H_2O}{kmol}\right)\left(\frac{kmol}{2\ kg\ H_2}\right)$$

$$\dot{m}_{H_2O} = 0.16\ kg/sec$$

Similarly,

The flow rate of SO_2:

$$\dot{m}_{SO_2} = 0.03\ kg/sec$$

The flow rate of O_2; we are using 12% excess air:

$$\dot{m}_{O_2} = O_2\ required\ for\ combustion\ C,S\ and\ H_2 + O_2\ in\ Coal + Excess\ O_2$$

$$\dot{m}_{O_2} = 0.17\ kg/sec$$

The flow rate of N_2:

$$\dot{m}_{N_2} = (O_2\ required\ for\ combustion\ C,S\ and\ H_2\ [kg\ O_2/sec])\left(\frac{0.77\ kg\ N_2 in\ Air}{0.23\ kg\ O_2 in\ Air}\right)(1.12)$$

$$\dot{m}_{N_2} = 4.32\ kg/sec$$

The results for the flue gas composition are given in Table 4.

Table (4): Flue gas flow rates and composition

Component	Molecular wt.	Mass flow rate kg/sec	Molar flow rate kmol/sec	Mole %
CO_2	44	5.35	0.12	41.82
H_2O	18	0.16	0.01	3.12
SO_2	64.06	0.03	0.00	0.15
O_2	31.99	0.17	0.01	1.87
N_2	28.01	4.32	0.15	53.05
Total		10.04	0.29	100

Absorbent Quantity; Material Balance on Carbonation Towers

Figure 5 shows the carbonation tower material flow. In the carbonation tower the following reaction:

$2\ NaOH + CO_2 \rightarrow Na_2CO_3 + H_2O$

Mass of NaOH =4.38 kg/sec =0.11 kmol/sec

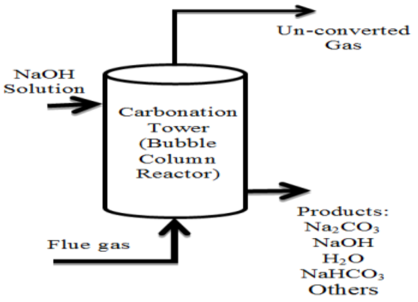

Figure 5. The carbonation tower material flow

Mass of H_2O = 0.99kg/sec =0.05 kmol/sec

Remaining CO_2 from the first reaction

Mass of CO_2 = (1-Conversion) (Flow of CO_2)

Mass of Na_2CO_3 =5.80 kg/sec =0.05 kmol/sec

Mass of CO_2 = 2.94 kg/sec =0.07 kmol/sec

The solution of NaOH (absorbent) is using 1 molar solution of NaOH. We have calculated the water required to produce a 1 molar solution. We are using the stoichiometric requirement of NaOH as our base of calculation.

$$Water\ Requirement = \left(\frac{m^3}{1\ kmol\ NaOH}\right)\left(\frac{994\ kg\ Water}{1\ m^3}\right)\left(\frac{kmol}{18\ kg\ Water}\right)\left(0.11\frac{kmol\ NaOH}{sec}\right)$$

Water Requirement=6.08 kmol/sec =109.4 kg/sec

Products Formed; Material Balance on Bi-Carbonation Towers

In the bi-carbonation tower the following reaction:

$Na_2CO_3 + H_2O + CO_2 \rightarrow 2NaHCO_3$

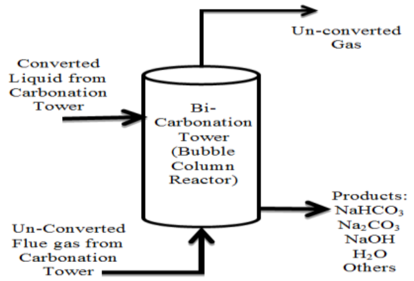

Figure 6. The bi-carbonation tower material flow

Figure 6 shows the bi-carbonation tower material flow.

Mass of $NaHCO_3$ = 8.99 kg/sec = 0.11 kmol/sec

Mass of Na_2CO_3 =5.67kg/sec= 0.05kmol/sec

Mass of H_2O =0.96 kg/sec =0.05 kmol/sec

Mass of CO_2 Left

Mass of CO_2 left =0.59 kg/sec =0.01 kmol/sec

Percent CO_2 left =11.00%

Effluent Contains

$\mathrm{Na_2CO_3} = 0.13\ \mathrm{kg/sec} = 1.22 \times 10^{-3}\ \mathrm{kmol/sec}$

$\mathrm{Water} = 109.6\ \mathrm{kg/sec} = 6.09\mathrm{kmol/sec} = 0.11 m^3/\mathrm{sec}$

$\mathrm{NaHCO_3} = 8.99\ \mathrm{kg/sec} = 0.11\ \mathrm{kmol/sec}$

$\mathrm{Impurities} = 0.06\ \mathrm{kg/sec}$

$$\mathrm{Total\ mass\ of\ slurry} = 118.81\frac{\mathrm{kg}}{\mathrm{sec}} = 6.20\frac{\mathrm{kmol}}{\mathrm{sec}}$$

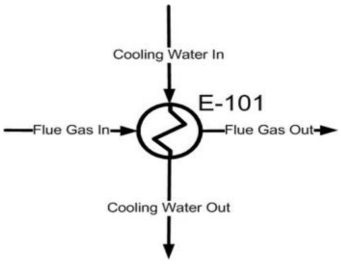

Figure 7. Flow of material from exchanger E-101

Energy Balance

The purpose of applying energy balance on our process is to determine the heat duties of process equipment and unknown temperatures of the streams. For energy balance, the known values are the temperature of flue gas (300 ^oC), coolant water inlet temperatures, the flow rate of flue gas, absorbent and other reagent flow rate, heats of formation and specific heats (from Aspen^{TM} properties). The required values from our energy balance are heat duties for heat exchangers, the heat produced by reaction and coolant water flow rate. Exchanger E-101 is utilized to cool the flue gas. Flue gas inlet temperature is 300 ^oC. Flow of material from exchanger E-101 is displayed in Figure 7.

Heat duty calculations on E-101 is shown in Table 5.

Table (5): Calculations for energy balance on E-101

	Flow rate	$C_{p,T1}$	$C_{p,T2}$	$C_p.T_1$	$C_p.T_2$	H	Q	Q
	$\dot{m}$	a	b	$c = a \cdot T_1$	$d = b \cdot T_2$	$H = c - d$	$m \cdot H$	
	$\frac{kg}{sec}$	$\frac{kcal}{kg \cdot {}^0C}$	$\frac{kcal}{kg \cdot {}^0C}$	$\frac{kcal}{kg}$	$\frac{kcal}{kg}$	$\frac{kcal}{kg}$	$\frac{kcal}{sec}$	$\frac{kJ}{sec}$
CO_2	5.35	0.23	0.23	68.70	32.10	36.60	195.8	819.68
H_2O	0.16	0.46	0.46	136.80	64.23	104.70	560.2	2344.9
SO_2	0.03	0.18	0.16	54.27	23.04	22.17	118.6	496.50
O_2	0.17	0.23	0.23	68.10	31.61	36.00	192.6	806.25
N_2	4.44	0.25	0.25	75.00	35.01	42.90	229.5	960.78
Total	10.15	1.34	1.33	402.87	186.00	242.36	1296	**5428**

Table (6): The calculation for energy balance on E-102

	Flow rate	$C_{p,T1}$	$C_{p,T2}$	$C_p.T_1$	$C_p.T_2$	H	Q	Q
	$\dot{m}$	a	b	$c = a \cdot T_1$	$d = b \cdot T_2$	$H = c - d$	$m \cdot H$	
	$\frac{kg}{sec}$	$\frac{kcal}{kg \cdot {}^0C}$	$\frac{kcal}{kg \cdot {}^0C}$	$\frac{kcal}{kg}$	$\frac{kcal}{kg}$	$\frac{kcal}{kg}$	$\frac{kcal}{sec}$	$\frac{kJ}{sec}$
CO_2	5.35	0.23	0.21	32.10	6.22	25.88	138.4	579.66
H_2O	0.16	0.46	0.44	64.23	13.31	58.0	310.4	1299.3
SO_2	0.03	0.16	0.15	23.04	4.57	16.82	90.02	376.79
O_2	0.17	0.23	0.22	31.61	6.60	25.39	135.9	568.69
N_2	4.44	0.25	0.25	35.01	7.47	28.79	154.1	644.88
Total	10.15	1.33	1.27	186.0	38.17	154.90	828.8	**3496.3**

where; $C_{p,T}$ = Specific Heat Capacity at $T_1$and T_2, H = Enthalpy, Q = Heat

The mass flow rate of coolant water; since,

$$\dot{Q} = \dot{m} \cdot C_p \cdot \Delta T \tag{3}$$

$$\dot{m} = \frac{\dot{Q}}{C_p \cdot \Delta T}$$

$$\dot{m} = \frac{5428\ [kJ/sec]}{(4.528\ [kJ/kg\ {}^0C])(70^0C - 25^0C)}$$

$$\dot{m} = 26.64 \frac{kg}{sec} = 95903 \frac{kg}{hr}$$

Exchanger E-102 also cools the flue gas. Flue gas inlet temperature is 140 ^oC. Figure 8 shows the flow of material from exchanger E-102.

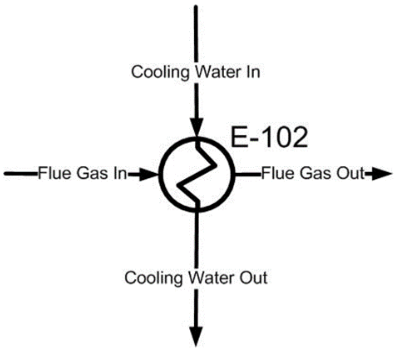

Figure 8. Flow of material from exchanger E-102

Heat duty calculations on E-102 is shown in Table 6.

The mass flow rate of coolant water;

$$\dot{m} = \frac{3469.3\ [kJ/sec]}{(4.528\ [kJ/kg\ {}^0C])(70^0C - 25^0C)}$$

$$\dot{m} = 17.03 \frac{kg}{sec} = 61295 \frac{kg}{hr}$$

Energy balance on carbonation towers; in our carbonation tower the main reaction is a carbonation reaction with the aqueous solution of NaOH, which is;

$2\ NaOH + CO_2 \rightarrow Na_2CO_3 + H_2O$

$$Heat\ of\ Reaction = Products\ Heat\ of\ formation - Reactants\ Heat\ of\ formation$$

$$\Delta H_R = \sum H_{f,products} - \sum H_{f,Reactants} \tag{4}$$

'H_f' for individual species that are taking part in the reaction taken from past studies are given in Table 7.

Table (7): The heat of formations and flow rate for material

Species	Heat of formation	Flow rate
Reactants	kJ/mol	mol/sec
NaOH	-469.62	109.45
CO_2	-393.68	66.88
Products		
Na_2CO_3	-1151.64	54.72
H_2O	-285.96	54.72

Putting the values will give
$Heat\ of\ reaction = \Delta H_R = (-1151.64 - 285.96) — (469.62 - 393.68) = -104.68\ kJ/mol$

The negative heat of reaction shows that it is an exothermic reaction.
Now calculating heat produced by exothermic reaction; Q_R
$Q_R = (F_{Na_2CO_3})(\Delta H_R)$
$Q_R = \left(\frac{54.72\ mol}{sec}\right)\left(\frac{104.68\ kJ}{mol}\right) = -5728.65\ kJ/sec$
In our bi-carbonation tower, the main reaction is a bi-carbonation reaction of CO_2 with the aqueous solution of Na_2CO_3, which is
$Na_2CO_3 + H_2O + CO_2 \rightarrow 2\ NaHCO_3$
Since H_f for individual species that are taking part in the reaction is presented in Table 8.

Table (8): The heat of formations and flow rate for material

Species	Heat of formation	Flow rate
Reactants	kJ/mol	mol/sec
Na_2CO_3	-1152	53.5
CO_2	-94.05	53.5
H_2O	-286	53.5
Products		
$NaHCO_3$	-930	107.01

$Heat\ of\ reaction = \Delta H_R = (-930) - (-1152 - 94.05 - 286) = -28.04\ kJ/mol$
Now calculating heat created by exothermic reaction; Q_R

$$Q_R = (F_{Na_2CO_3})(\Delta H_R) \tag{5}$$

$$Q_R = \left(\frac{107.01\ mol}{sec}\right)\left(\frac{-28.04\ kJ}{mol}\right) = -\ 3001\ kJ/sec$$

3. RESULTS

Designing of Equipment

Exchanger E-101 is a shell and tube type heat exchanger utilized to cool the flue gas from 300 to 140 ^oC. The general purpose of designing an exchanger is to regulate the heat transfer area. Since heat duty has been calculated from energy which was:
$\dot{Q} = 5428.13\ kJ/sec$
From literature, heat capacities of water and flue gas are:
$C_{p,Water} = 4.53\ kJ/kg°C$
$C_{p,Flue\ Gas} = 1.05\ kJ/kg°C$
Fluid temperatures of flue gas, as a hot fluid has the following inlet and outlet temperatures:
$\text{Hot Fluid Temperature at the inlet}, T_1 = 300°C$
$\text{Hot Fluid Temperature at the outlet}, T_2 = 140°C$
Water, as a cold fluid has the following inlet and outlet temperatures:
$\text{Cold Fluid Temperature at the inlet}, t_1 = 25°C$
$\text{Cold Fluid Temperature at the inlet}, t_2 = 70°C$
The log means temperature difference:

$$\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{ln\frac{(T_1 - t_2)}{(T_2 - t_1)}} \tag{6}$$

$$\Delta T_{lm} = \frac{(300 - 70) - (140 - 35)}{ln\frac{(300 - 70)}{(140 - 35)}} = 160\ °C$$

For correction factor F_t, we have to determine the two dimension-less temperature ratios R and S, which are determined as follows:

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} \tag{7}$$

$R = 4.5$
and

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} \tag{8}$$

$S = 0.13$
We have selected 1-shell two 2-passes exchanger for this operation. Temperature correction factor F_t for this type is; $F_t = 0.96$. The next step is to assume the total heat transfer coefficient:
$U_{Assumed} = 250\ W/m^2°C$
The provisional area is evaluated by the equation as:
$\dot{Q} = UA\Delta T_{lm}F_t$
(9)
Putting the values in the above equation will give
$A = 139.24\ m^2$
First, choosing tube dimensions, shown in Table 9.

Table (9): Tube configuration for exchanger E-101

Parameter	inch	mm	m
Outside dia (OD)	0.75	19.05	0.01905
Length (L)	5	4830	4.83
BWG	14		
ID	0.58	14.83	0.014834
Thickness (t)	0.08	2.1	0.002108
Material	Cupro-Nickel		

Area of the single tube

$$A_{1\ tube} = (\pi)(ID)(L) \tag{10}$$

$A_{1\ tube} = 0.2993\ m^2$
The number of tubes is expressed by the following formula:

$$Number\ of\ Tubes = n = \frac{A}{\pi \cdot OD \cdot L} \tag{11}$$

$Number\ of\ Tubes = n = 474\ tubes$
Bundle and shell diameter; for two tube passes, $K_1 = 0.25$ and $n_1 = 2.21$
This gives the bundle diameter:

$$Bundle\ dia = OD\left(\frac{n}{K_1}\right)^{1/n_1} \tag{12}$$

$Bundle\ dia. = 0.59\ m$
Splitting floating head type, bundle diametric clearance = 57 mm
$Shell\ dia.$
$= Bundle\ Dia. + Bundle\ diametric\ Clearance$
$Shell\ dia. = 0.65\ m$
Tube-side heat transfer co-efficient; properties for flue gas are taken at mean temperature =52.50 ^oC from Aspen^{TM} properties.
$\text{Viscosity} = \mu = 5.29 \times 10^{-3} \text{N} \cdot \text{sec/m}^2$
$\text{Thermal Conductivity} = \text{k} = 0.64\ \text{W/m°C}$
$\text{Density} = \rho = 986.9\ \text{kg/m}^3$
Calculating,

$$Tube\ Cross\ sectional\ area = \frac{\pi}{4}(ID)^2 \tag{13}$$

$Tube\ Cross\ sectional\ area = 1.73 \times 10^{-4} m^2$
also,
$Tube\ per\ pass = \frac{n}{2} = 238$
also,

$$Total\ Flow\ Area = Tube\ per\ pass \times Tube\ Cross\ sectional\ area$$

$Total\ Flow\ Area = 4.16 \times 10^{-2}\ m^2$

and,

$$Tube\ side\ velocity = u_t = \frac{(Volumetric\ flow\ rate)}{(Total\ Flow\ Area)} \qquad (14)$$

$$Tube\ side\ velocity = u_t = \frac{(34.25\ [kg/sec])/\rho}{4.16 \times 10^{-2}\ [m^2]} 0.84\ m/sec$$

The Velocity is Satisfactory, between 1 to 2 m/sec, but maybe little low. Reynold number shows when the pressure drops [13, 14];

$$Re = \frac{\rho u_t ID}{\mu} \qquad (15)$$

$Re = 23366 = 2.34 \times 10^4$

Prandtl Number

$$Pr = \frac{(C_{p,Flue\ Gas})(\mu)}{k} \qquad (16)$$

$Re = 23366 = 2.34 \times 10^4$

Length diameter ratio:

$\frac{L}{OD} = 262$

J_h Factor:

$J_h = 2.3 \times 10^{-3}$

Inside heat transfer co-efficient; using the following correlation

$$h_i = \frac{4200(1.35 + (0.02T\ [°C]))(U_t^{0.8}[m/s])}{(ID\ [mm])^{0.2}} = 5133.7\ W/m^2°C$$

This h_i is too low if U_o is to be 250 $W/m^{2o}C$, so increase the number of tubes passes to 4. Now, assuming 2-shell 4-tube passes exchanger. Following are the changed values for this scheme:

$F_t = 0.9907$

$A = 143\ m^2$

$A_{1\ tube} = 0.2993\ m^2$

$Number\ of\ Tubes = n = 474\ tubes$

$Tube\ side\ velocity = u_t = 1.68\ m/sec$

Now the velocity is suitable between 1 to 2 *m/sec*.

$K_1 = 0.175$ and $n_1 = 2.285$

$Bundle\ Dia. = 0.6\ m$

Splitting floating head type, bundle diametric clearance = 62 mm

$Shell\ dia. = 0.65\ m$

$Re = 46494 = 4.65 \times 10^4$

$Pr = 3.74 \times 10^{-3}$

$\frac{L}{OD} = 262$

$J_h = 3.4 \times 10^{-4}$

Now h_i becomes

$h_i = 8902\ W/m^2°C$

This h_i is too low if U_o is to be 250 $W/m^{2o}C$. As tube side velocity is low, increase the tubes passes to 4. Shell side heat transfer co-efficient; values of water properties, at mean temperature 220°C, are given as under:

$Viscosity = \mu = 2.48 \times 10^{-5} N \cdot sec/m^2$

$Thermal\ Conductivity = k = 3.56 \times 10^{-2}\ W/m°C$

$Density = \rho = 0.8517\ kg/m^3$

$Heat\ Capacity = C_p = 1.045\ kJ/kg°C$

Baffle spacing: Choosing baffle spacing for 15 baffles

$$Baffle\ space = \frac{Shell\ Dia.}{15} = 4.4 \times 10^{-2}\ m$$

Tube pitch: is taken as 1.25 of outer dia. of tubes

$Tube\ pitch = 1.25 \times OD = 2.3 \times 10^{-2}\ m$

Shell side flow area and mass velocity:

$$Flow\ Area_{Shell} = (Baffle\ space)(\text{Shell dia.})\frac{(Tube\ pitch - OD)}{Tube\ pitch} \qquad (17)$$

$Flow\ Area_{Shell} = 5.983 \times 10^{-3}\ m^2$

and

$Mass\ Velocity = 1578.23\ kg/sec.m^2$

Equivalent Dia.:

$$d_e = \frac{1.10}{OD}\{(Tube\ pitch)^2 - 0.917(OD)^2\}$$

$d_e = 1.35 \times 10^{-2} m$

Reynold number, Prandtl number, and J_h factor: are calculated as were calculated in the tube side.

$Re = 9.25 \times 10^5$

$Pr = 7.28 \times 10^{-4}$

$J_h = 1.30 \times 10^{-3}$

Outside heat transfer co-efficient

$$h_o = \frac{k \cdot J_h \cdot Re \cdot Pr^{0.33}}{d_e} \qquad (18)$$

$h_o = 1 \times 10^7\ W/m^2°C$

Over-all heat transfer co-efficient; the thermal conductivity of the material used: is given as under;

$k_{Cupro-Nickel} = 50\ W/m^2°C$

Fouling Factor: for water and flue gas are taken as their typical values from the literature;

$h_{if} = h_{f,flue\ gas} = 5000\ W/m^2°C$

$h_{of} = h_{f,Water} = 6000\ W/m^2°C$

Overall heat transfer coefficient: based on the outside surface area

$$\frac{1}{U} = \frac{1}{h_o} + \frac{1}{h_{of}} + \frac{OD ln\left(\frac{OD}{ID}\right)}{2k_w} + \left(\frac{OD}{ID}\right)\frac{1}{h_{if}} + \frac{1}{h_i} \qquad (19)$$

$\frac{1}{U} = 4.04 \times 10^{-3} = 247\ W/m^2°C$

This value is approximately the same as the assumed value of U earlier, which shows our provisional area of 140 m^2 is correct for the design procedure. Pressure drop tube side pressure drop can be calculated by the following correlation:

$$\Delta P = N_p \times \left(8j_f\left(\frac{L}{ID}\right)\left(\frac{\mu}{\mu_w}\right)^{-m} + 2.5\right)\frac{\rho u_t^2}{2} \qquad (20)$$

where; $\Delta P =$ Tube-side pressure drop (N/m^2), $N_p =$ Number of tube passes, $u_t =$ Tube-side velocity (m/s), $L =$ Length of one tube, $j_f =$ Dimensionless friction factor, Values of j_f for heat exchanger tubes can be obtained volume 6 found value is 3.4E-3

ρ = Density of tube fluid

m = 0.25 for laminar flow, Re < 2100,

m = 0.14 for turbulent flow, Re >2100.

Now, putting all the values in this equation will give

$$\Delta P = 65009.8\frac{N}{m^2} = 65\ kPa = 0.65\ bar$$

Shell side pressure drop can be calculated by the following correlation.

$$\Delta P = \frac{8j_f d_s L \rho u_s^2}{2d_e\beta} \qquad (21)$$

where; $\Delta P =$ Tube-side pressure drop (N/m^2), $u_s =$ Shell-side velocity (m/s) = mass velocity/density, $L =$

Length of one tube, j_f = Dimensionless friction factor, found value is *1.8E-3*, ρ = Density of shell-side fluid, d_s = Shell diameter (m), d_e = Equivalent dia. m

$$\Delta P = 134917\,\frac{N}{m^2} = 134.9\ kPa = 1.35\ bar$$

Designing of pump P-101; used for the pumping of the aqueous solution of caustic soda into the reactor. For this pump design, we have the following known data:

Mass flow rate $= \dot{\mathrm{m}} = 113.82$ kg/sec

Temperature $= \mathrm{T} = 30.00^{0}\mathrm{C}$

Density $= \rho = 1039.00\ \mathrm{kg/m^3}$

Flow rate volumetric $= \dot{\mathrm{v}} = 1.10 \times 10^{-1} m^3/sec$

Specific Gravity $= \mathrm{S_g} = 1.04$

Viscosity $= \mu = 8.20\mathrm{E}-04\ \mathrm{Pa.s}$

Pipe dia. For an assumed velocity; assume typical velocity, $u = 2.25\ m/sec$; for this velocity,

$Area\ of\ pipe = A = 4.87 \times 10^{-2}\ m^2$

$Dia\ of\ pipe = d = 2.49E-01\ m\ =\ 9.80\ inch$

For this area of the pipe, pipe specification is selected and is tabulated in Table 10.

Table (10): Pipe specifications for pump P-101

Specifications	Unit
Nominal Dia.	10.00 inch
Schedule No.	40.00
Outer Dia.	10.75 inch
Inner Dia.	10.02 inch
Wall Thickness	0.37 inch

Actual design velocity is calculated by the following formula:

$$u_{actual} = \dot{v} \cdot A$$

$$u_{actual} = 2.15\ m/s$$

Maximum allowable velocity:

$$u_{allowable} = 1.2^2 u_{actual}$$

$$u_{allowable} = 3.10 m/s$$

Frictional loss/Pressure drop in pipe Reynold number is calculated first

$$Re = 694340.31$$

Fluid mass velocity:

$$G = 2237.10\ kg/m^2sec$$

Pressure Drop in line: is calculated by the following correlation.

$$\Delta P_{line} = \frac{4.13 \times 10^{10} m^{1.84} \mu^{0.16}}{\rho \cdot ID^{4.84}}$$

where viscosity is in mN.sec/m^2 and inner diameter (ID) is in mm, and all others are in the SI system.

$$\Delta P_{line} = 3.68 \times 10^{-1} Pa$$

Relative roughness is the ratio of absolute roughness and pipe inside diameter.

where absolute roughness for commercial steel pipe is taken from literature.

$$Absolute\ roughness = 0.046\ mm$$

This gives the value for relative roughness

$$\varepsilon = \frac{Absolute\ roughness}{ID} = 1.81 \times 10^{-04}$$

Friction factor:

$$f = 0.0018$$

Miscellaneous losses have to install 3, 90^{o} elbows and two valves in 100m pipe. Their equivalent lengths are taken from literature.

Equivalent Length for 90 Elbows = 17.56 m

Equivalent Length for Valves = 1.27 m

Pipe length = 100 m

Will give the total equivalent length for design:

Total length = L = 118.83 m

Total pressure drop: is calculated by the following correlation:

$$\Delta P_f = \frac{8fL\rho u^2}{2(ID)} \qquad (22)$$

$$\Delta P_f = 16192.89\ Pa = 16.2\ kPa$$

The energy balance calculation will provide us with the details of the work requirements. The elevation is assumed for design purposes.

$$\Delta z = z_1 - z_2 = -30\ m$$

Inlet pressure

$$P_1 = 101.33\ kPa$$

Outlet pressure

$$P_2 = 202.65\ kPa$$

Pressure difference

$$\Delta P = -101.33\ kPa$$

Work is calculated by applying the energy balance equation as:

$$g\Delta z + \frac{\Delta P}{\rho} - \frac{\Delta P_f}{\rho} - W = 0$$

$$W = -294.38\ J/kg$$

Head required for the pump can be given as;

$$H = \frac{\Delta P_f}{\rho g} - \frac{\Delta P}{\rho g} - \Delta z \qquad (23)$$

$$H = 30.01\ m$$

Power requirement:

$$Efficiency\ (\eta) = 70\%$$

$$Power = \frac{|Work\ Required|}{\eta} \qquad (24)$$

$$Power = 48\ kWatt$$

NPSH$_{\mathrm{avail.}}$ for the pump is calculated by the following equation:

$$NPSH_{avail.} = \frac{P}{\rho} + H - \frac{P_f}{\rho} - \frac{P_v}{\rho} \qquad (25)$$

where,

Pv = vapour pressure of the liquid at the pump suction

$Pv = 5.40\ kPa$

Pf = the pressure loss in the suction piping

$Pf = 16.19\ kPa$

H = height of liquid above the pump suction = 0

P = pressure above the liquid in the feed vessel

$P = 101.33\ kPa$

$NPSH_{avail.}$ = net positive suction head available at the pump suction

Putting all the values in the above equation will give

$$NPSH_{avail.} = 7.82m$$

NaOH Feed Mixer MX-101 Designing

Mixer MX-101 is used to mix water in the feed NaOH to acquire 4 wt. % NaOH solution for the reaction of NaOH with the flue gas. The temperature is 35^{o}C. The design of the mixer is based on the reactor requirement, which was calculated in material balance. The required solution weight is given as follows:

$$Total\ Solution\ Weight\ required = Water\ required\ + NaOH\ required$$

$$Total\ Solution\ Weight\ required = 113.82\ kg/sec$$

or in terms of volume

$$Solution\ Volume = 0.11\ m^3/sec$$

Assuming the volume of a single mixer, V = 19000 gals.

Tank diameter is calculated by:

$D_t = 6.6513V^{0.33}$

$D_t = 176.90\ in = 4.49\ m$

Tank depth is assumed to be the same as tank dia.

$Tank\ Depth = 176.90in = 4.49\ m$

Agitator diameter is based on typical design can be calculated as:

$$Da = \left(\frac{1}{3}\right)D_t = 1.50\ m$$

Agitator position and blade width are calculated Eq. (26) and Eq. (27):

$$E = \left(\frac{1}{3}\right)D_t \qquad (26)$$

$E = 1.50\ m$

$$W = \left(\frac{1}{5}\right)Da \qquad (27)$$

$W = 0.30\ m$

Turbine length is given as:

$$L = \left(\frac{1}{4}\right)Da \qquad (28)$$

$L = 0.37m$

Fluid Height: In the mixer, the liquid fill is supposed to be 75%

$H = 0.75(D_t)$

$H = 3.37m$

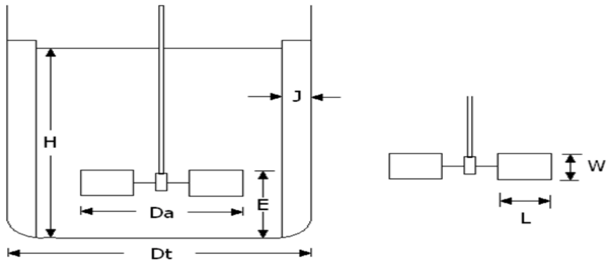

Figure 9. Mixing tank with agitator dimension.

Figure 9 shows the mixing tank with agitator dimension. For our mixing purpose, a disk turbine with six flat blades is used. Impeller rotation speed is assumed to be 90 rpm. Fluid properties viscosity and density of the solution is given below:

$\text{Viscosity} = \mu = 0.009\ \text{Pa.sec}$

$\text{Density} = \rho = 1043.46\ \text{kg}/m^3$

Reynold Number is calculated as under:

$$Re = \frac{(Da^2)(speed\ [rps])(\rho)}{\mu}$$

$Re = 3.90 \times 10^5$

Since $Re > 10^4$, In baffled tanks, for $Re > 10^4$ the power number is independent of Re and viscosity is not a factor

$Np = Kt$

For a baffled tank with 4 baffles and width equal to 10% of tank diameter. For Six Blade disk $Kt = 5.75$

$Re = 3.90 \times 10^5$

$Mixing\ Time\ Factor\ (nt_T) = 39$

$$Mixing\ Time\ (t_T) = \frac{Mixing\ Time\ Factor\ (nt_T)}{Speed\ [rps]} \qquad (29)$$

$Mixing\ Time\ (t_T) = 26\ sec$

Tip velocity

$$Tip\ Speed\ (u) = \pi \times Da \times Speed\ [rps] \qquad (30)$$

$Tip\ Speed\ (u) = 7.06\ m/sec$

Power required

$$P = (Kt)(speed\ [rps])^3(Da)^5(\rho) \qquad (31)$$

$P = (5.75)(1.5)^3(1.5)^5(1043.46)$

$P = 152629.26\ W = 152.63\ kW$

Table (11): Slurry specification

	Weight	Mass Fraction	Density	Solids	Solid Fractions	Solid Density
	kg/sec		kg/m^3	kg/sec		kg/m^3
Na_2CO_3	0.13	1E-03	435.70	0.13	0.01	6.12
Water	109.6	0.92	994.10			
$NaHCO_3$	8.99	0.08	345.30	8.99	0.98	338.22
Impurities	0.06	5E-04	990.50	0.06	0.01	6.41
Total	118.8	1		9.18	1	350.74

Rotary Vacuum Filter is to be used of separation of reactor effluent or to separate salts from water. The reactor effluent contains a huge amount of water with some desired salts and some un-desired. Slurry specifications, mass flow rate, mass fraction, density, solid weight, solid fractions, and solid density, are shown in Table 11. To find that there are 7.73% solids in the slurry.

For Normal Concentration (1-10%), settling rate would be "slow", we have ranges of

$\text{Filterate Rate} = 0.02 - 5\ \text{m}^3/\text{h.m}^2$

$\text{Cake Formation Rate} = 0.02 - 0.12\ \text{mm/sec}$

$\text{Leaf Test Rate} = 25 - 250\ \text{kg/h.m}^2$

$\text{Speed} = 0.1 - 0.25\ \text{rev/min}$

Weight of Dry Cake per m^3 of filtrate is calculated by the following relation:

$$W = \frac{Total\ solid\ weight}{\left(\frac{Liquid\ Weight}{\rho_{water}}\right)}$$

$W = 83.22\ kg/m^3$

Weight of Dry Cake per unit time: would be equal to total solid weight and is represented by $V_R N_R W$.

$V_R N_R W = 33041.58\ kg/hr$

The volume of filtrate per unit time is calculated by dividing W by $V_R N_R W$.

$$V_R N_R = 397.02\frac{m^3}{hr} = 0.11m^3/sec$$

Speed is assumed to be:

$N_R = 0.20\ rev/min = 12\ rev/hr$

Cycle time since 1 rev takes 5 min,

$t_C = 5\ min \quad = 300\ sec$

The volume of Filterate per Revolution: is calculated as under:

$$V_R = \frac{\left(397.02\ \frac{m^3}{hr}\right)}{\left(12\frac{rev}{hr}\right)}$$

$V_R = 33.08\, m^3/rev$

Filtering time t_f: is calculated by multiplying 'fraction of drum area submerged in slurry' (ψ_f) by cycle time (t_c). Take $\psi_f = 0.35$

$t_f = 1.75\, min = 105\, hr$

Filterate rate is selected from data to be 4 m^3/h-m^2,

$$\frac{V_R N_R}{A_D} = 4\, m^3/hm^2$$

$A_D = 99.25\, m^2$

Volumetric rate of deposition of solid (bulk):

$V_R N_R(Solid\ Fraction) = 0.01$

Bulk volume of cake deposited per revolution:

$(V_R N_R)(Solid\ Fraction)(t_C) = 2.56\, m^3$

Cake thickness:

$$\frac{(V_R N_R)(Solid\ Fraction)(t_C)}{A_D} = 0.03\, m = 25.75\, mm$$

The density of cake is the solid density

$\rho_c = 350.74\, kg/m^3$

Mass of Cake Deposited per revolution

$(V_R N_R)(Solid\ Fraction)(t_C)(\rho_c) = 896.44\, kg/m^2$

Power requirement for vacuum pump data;

Pressure (P) $= 1.00$ atm $= 101.24$ kPa

Vaccum Pressure (P_{Vaccum}) $= 0.27$ atm $= 27.54$ kPa

Temperature (T) $= 35$ ℃

Fraction of drum area submerged in the slurry (ψ_f) $= 0.35$

Fraction of drum area available for suction (ψ_a) $= 0.10$

$\alpha/\beta = 0.60$

Since the resistance of filter, the medium is negligible

$\mu_{Water} = 0.74$ cP

μ_{Air} (at 4 Psia) $= 0.02$ cP

$\mu_{Water}/\mu_{Air} = 38.34$

$P - P_{VAccum} = P_1 = 73.70$ kPa

$P = P_2 = 101.30$kPa

The volume of Dry Air per unit Time ($\upsilon_{Dry\ Air}$):

$$\upsilon_{Dry\ Air} = \frac{V_R N_R W \psi_a \left(\frac{\mu_{Water}}{\mu_{Air}}\right)\left(\frac{\alpha}{\beta}\right)}{2\psi_f W} \quad (32)$$

$$\upsilon_{Dry\ Air} = 0.36 \frac{m^3}{sec}$$

Volumetric flow rate at vacuum pump inlet (υ_{in}):

$$\upsilon_{in} = \upsilon_{Dry\ Air}\left(\frac{P}{P - P_{VAccum}}\right) \quad (33)$$

$$\upsilon_{in} = 0.50\, \frac{m^3}{sec}$$

Power (P):

$$P = \frac{k}{k-1} P_1 \upsilon_{in}\left(\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1\right) \quad (34)$$

where; $k = (C_p/C_v) = 1.4$, will give power for 100% efficiency and $P_{100\%} = 12.22\, kW$

The efficiency of the isentropic Compressor is taken to be 60% as its typical value. Multiplying $P_{100\%}$ with 0.6 to get;

$Power = P = 20.37\, kW = 27.32 hp$

Investment Cost Required For Project

Equipment numbers are mentioned according to the PFD

$Chemical\ Engineering\ Cost\ \ Index\ of\ 2019 = 593.8$

$Chemical\ Engineering\ Cost\ \ Index\ of\ 2014 = 525.4$

$$Cost\ in\ the\ year\ 2019 = Cost\ in\ the\ year\ 2014 \times \frac{C.E\ Cost\ index\ in\ year\ 2019}{C.E\ Cost\ index\ in\ the\ year\ 2014}$$

$$Cost\ in\ the\ year\ 2019 = Cost\ in\ the\ year\ 2014 \times 1.13$$

Heat Exchanger E-101; Exchanger type shell and tube

Area =142.72 m^2

Material Carbon Steel

Internal Pressure= 28.78 kPa

Purchased cost in 2014= \$ 108,300.00

$Purchase\ cost\ in\ 2019 = \$\ 122{,}399.20$

Heat Exchanger E-102 (approx.); Exchanger type Shell and tube

Area=100 m^2

Material Carbon Steel

Internal Pressure =28.78 kPa

Purchased cost in 2014= \$ 99,600.00

$Purchase\ cost\ in\ 2019 = \$\ 112{,}566.58$

Pump P-101; Pump Type Centrifugal

Discharge Pipe Diameter 10 inch

Material Alloy 20

Seal Type Mechanical Seal

Purchased cost in 2014= \$ 28,400.00

$Purchase\ cost\ in\ 2019 = \$\ 32{,}097.30$

Pump P-102 (approx.); Pump Type Centrifugal

Discharge Pipe Diameter 10 inch

Material Carbon Steel

Seal Type Mechanical Seal

Purchased cost in 2014= \$ 13,400.00

$Purchase\ cost\ in\ 2019 = \$\ 15{,}144.50$

Blower B-101 (approx.); Bower Capacity 6351.75 ft^3/min

Blower Type Axial Large 1 atm, 0.5 atm vacuum

Material Carbon Steel

Purchased cost in 2014 = \$ 44,100.00

$Purchase\ cost\ in\ 2019 = \$\ 9{,}841.23$

Rotary Drum Filter RF-101; Filter Area 100 m^2

Material Carbon Steel

Pressure Atmospheric

Purchased cost in 2014= \$ 472,600.00

$Purchase\ cost\ in\ 2019 = \ \$\ 534{,}126.15$

Mixer MX-101, MX-102, MX-103, MX-104

$Chemical\ Engineering\ Cost\ \ Index\ of\ 2014 = 390.4$

$$\frac{\text{Chemical Engineering Cost index in the year 2019}}{\text{Chemical Engineering Cost index in the year 2014}} = 1.52$$

Agitator type	Disk Turbine with six flat blades
Mixer Volume	72 m^3
Material	Carbon Steel

Purchased cost in 2014= \$ 72,391.40

$Purchase\ cost\ in\ 2019 for\ 1\ Mixer = \$\ 110{,}107.62$

We have 4 Mixer in Parallel with the same capacity

$Purchase\ cost\ in\ 2019 for\ 4\ Mixers = \$\ 440{,}430.46$

Bubble Column Reactor (approx.); Carbonation Reactor C-101, C-102, C-103 & C-104

Reactor type Fermenter

Reactor Volume=26 m^3, Material Carbon Steel

Pressure= 1 to 5 atm

Purchased cost in 2014= \$ 79,800.00

$Purchase\ cost\ in\ 2019 = \ \$\ 90{,}188.88$

We have 4 Carbonation Reactor in parallel with the same capacity

$Purchase\ cost\ for\ 4\ reactors\ in\ 2019 = \$\ 360{,}755.54$

Bubble Column Reactor (approx.); Bi-Carbonation Reactor BC-105, BC-106, BC-107 & BC-108

Reactor type Fermenter

Reactor Volume=26 m^3

Material Carbon Steel

Pressure= 1 to 5 atm

Purchased cost in 2014= $ 79,800.00

$$Purchase\ cost\ in\ 2019 = \$\ 90{,}188.88$$

There are 4 Carbonation Reactor in parallel with the same capacity

$$Purchase\ cost\ for\ 4\ reactors\ in\ 2019 = \$\ 360{,}755.54$$

Rotary Dryer RD-101 (approx.); Dryer type Indirect Gas Fired

Surface Area 46.7 m^2

Material Carbon Steel

Purchased cost in 2014= $ 187,000.00

$$Purchase\ cost\ in\ 2019 = \$\ 211{,}344.88$$

Total Plant Equipment Cost (PCE); PCE is the sum of all the above equipment costs.

$$Plant\ Equipment\ Cost\ 2019\ (PCE) = \$\ 2{,}239{,}461.37$$

Fixed Capital Cost

$$Total\ Physical\ Plant\ Cost\ (PPC) = PCE\ (1 + f1 + f2 + f3 + f4 + f5 + f6 + f7 + f8 + f9)$$

Process Type Fluids-Solids

where; f1= 0.45 (Equipment erection), f2= 0.45 (Piping), f3= 0.15 (Instrumentation), f4= Electrical(0.1), f5= 0.1 (Building Process), f6= 0.45 (Utilities), f7= 0.2 (Storages), f8= 0.05 (Site development) and f9=0.2 (Ancillary buildings)

$$Total\ Physical\ Plant\ Cost\ (PPC) = \$\ 7{,}054{,}303.33$$

$$Fixed\ Capital = PPC * (1 + f10 + f11 + f12)$$

Process Type Fluids-Solids

where; f10 = 0.25 (Design and engineering), f11 = 0.05 (Contractor's fee) and f12 = 0.1 (Contingency)

$$Fixed\ Capital = \$\ 9{,}876{,}024.66$$

Working capital cost; is taken as 10-20% of the fixed capital cost. We are taking 15%.

$$Working\ Capital = 0.15 * fixed\ capital$$

$$Working\ Capital = \$\ 1{,}481{,}403.70$$

Land cost

$$Land\ Cost = 2(PPC)$$

$$Land\ Cost\ required\ for\ this\ Project = \$\ 14{,}108{,}606.66$$

Total investment for the project

$$Total\ Investment\ Required\ For\ Project = Fixed\ Capital + Working\ Capital + Land\ Cost$$

$$Total\ Investment\ Required = \$\ 25{,}466{,}035.02$$

Annual operating cost,

Plant Attainment=0.9, Annual Operating Time=7884 hrs. /year

Variable costs; Raw material cost

NaOH (50 w/w %) required=31536 kg/hr.

NaOH (50 w/w %) Price=$ 215/ton

$$Annual\ Cost\ for\ NaOH = \$\ 53{,}455{,}412.16$$

Utilities Cost

Main Water required=3424032 gal/day, Electricity Requirement=2089260 kW/year, Main Water Price= $ 0.0017, Electricity Price= $ 0.15070, Annual Cost for Electricity=$ 314,851.48, Annual Cost for Water= $ 956,075.34

$$Annual\ Cost\ for\ Utilities = \$\ 1{,}270{,}926.82$$

Miscellaneous Materials

$$Annual\ Miscellaneous\ Cost = 10\%\ of\ Total\ Maintenance\ Cost$$

$$Annual\ Cost\ for\ Miscellaneous = \$\ 98{,}760.25$$

Shipping and packaging; annual cost for shipping and packaging are neglected usually. Variable cost is calculated by summing raw material cost, utility cost, miscellaneous cost and shipping and packaging cost [21, 22].

$$Variable\ Cost = \$\ 54{,}825{,}099.22$$

Fixed costs;

$$Fixed\ Cost = Maintenance + Operating\ labour + Plant\ overheads + Insurance + Local\ Taxes + Licence\ Fee + Capital\ charges + Laboratory + Supervision$$

Maintenance cost; 10% of Fixed Capital Cost

$$Annual\ Maintenance\ Cost = \$\ 987{,}602.47$$

Operating labour cost

Labor Quantity (Assume) =50, 15 labors per shift, 3 shift a day & 5 labors extra Shift =8 hr/day and Labor Cost for per hour and per worker = $ 1/hr

$$Annual\ Operating\ Cost = \$\ 146{,}000.00$$

Laboratory cost; 20 - 23 % of operating labour cost

$$Laboratory\ Cost = \$\ 32{,}120$$

Supervision cost; 20 % of operating labour cost

$$Supervision\ Cost = \$\ 29{,}200$$

Plant overhead cost; 50 % of operating labour cost

$$Annual\ Plant\ Overhead\ Cost = \$\ 73{,}000$$

Capital charges; 10% of fixed capital cost

$$Annual\ Capital\ Charges = \$\ 987{,}602.47$$

Local taxes; 2% of fixed capital cost

$$Local\ Taxes\ for\ Annum = \$\ 197{,}520.49$$

Insurance cost; 1% of fixed capital cost

$$Insurance\ Cost\ for\ Annum = \$\ 98{,}760.25$$

License fee and royalty payment; 1 % of fixed capital cost

$$Annual\ Licence\ Fee = \$\ 98{,}760.25$$

So, the fixed cost is

$$Fixed\ Cost = \$\ 2{,}650{,}565.92$$

Annual operating cost; annual operating cost is the sum of fixed cost and variable cost [23, 24].

$$Annual\ Operating\ Cost = \$\ 57{,}475{,}665.14$$

Annual production cost

$$Production\ Cost\ per\ kilogram = \frac{Annual\ Operating\ Cost}{Annual\ Production\ Rate} \qquad (35)$$

$$Production\ cost\ per\ kilogram = \$\ 0.23/kg$$

Revenue; Soda Bi-Carbonate as our product, Soda Bi-Carbonate (from material balance) =255156063.6 kg/year

Soda Bi-Carbonate Market Price= $ 325/ton

$$Annual\ Revenue = \$\ 82{,}925{,}720.66$$

Total Profit

$$Total\ Profit = Total\ Annual\ Revenue - Annual\ Operating\ Cost$$

$$Annual\ Profit = \$\ 25{,}450{,}055.51$$

$$Profit\ per\ day = \$\ 77{,}473.53$$

4. CONCLUSIONS

Instead of covering it underground, researchers are creating forms that utilization CO$_2$ emissions as raw material for chemicals and are also converting to saleable products. In short CO$_2$ from industrial exhausts can be changed over into valuable items and synthetic intermediates. Supported by government financing, plans in the industry are taking a gander at different alternatives to make economically reasonable procedures. Driving CO$_2$ to respond is tricky, yet new procedures are being formed to change over it into polymer feedstock, biofuels, and carbonate salts. The CO$_2$ gas is the principal greenhouse gas that researchers prescribe humankind should control discharges of to stay away from hazardous environmental change. However, with the world depending on consuming petroleum products for power, no sufficient controls on CO_2 production have yet been enacted. Carbon catch and capacity is an incredible conceivable solution. Thus, just as the specialized troubles of catching the gas, a noteworthy inquiry stays over how to manage the CO_2 once caught. Presently, scientists are creating forms that utilization exhaust gases as beginning materials for helpful materials and organizations are urging CO_2 to drop it's far off un-reactivity, and join the movement of modern assembling. The research effectively captures more than 90% CO_2, additionally expels sulfur dioxides, nitrogen dioxide, mercury, and other overwhelming metals from vent gas streams so that it might supplant existing scrubber technology. The strong carbonates that the procedure at first makes also give an alternative to carbon sequestration that stays away from pipelines, underground infusion, and worries about CO_2 spilling again into the environment, however by itself don't make a benefit. Although, mineralization can transform the carbonates into high purity sodium bicarbonate. The economics of the process was assessed and the process is profitable with a payback period of just 2 years. The best part about the process is it can be retrofitted and scaled up to the requirements. This process can also be applied to other sectors as well such as power plants and other GHG emitting industries.

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REFERENCES

[1] García-Gusano, D., Garraín, D., Herrera, I., Cabal, H. and Lechón, Y., "Life Cycle Assessment of applying CO_2 post-combustion capture to the Spanish cement production", *Journal of Cleaner Production*, **104** 328-338 (2015).

[2] Hills, T., Leeson, D., Florin, N. and Fennell, P., "Carbon Capture in the Cement Industry: Technologies, Progress, and Retrofitting", *Environ Sci Technol*, **50** (1): 368-377 (2016).

[3] Hägg, M.B., Lindbråthen, A., He, X., Nodeland, S.G. and Cantero, T., "Pilot Demonstration-reporting on CO_2 Capture from a Cement Plant Using Hollow Fiber Process", *Energy Procedia*, **114** 6150-6165 (2017).

[4] Cormos, A.-M. and Cormos, C.-C., "Reducing the carbon footprint of cement industry by post-combustion CO_2 capture: Techno-economic and environmental assessment of a CCS project in Romania", *Chemical Engineering Research and Design*, **123** 230-239 (2017).

[5] Atsonios, K., Grammelis, P., Antiohos, S.K., Nikolopoulos, N. and Kakaras, E., "Integration of calcium looping technology in existing cement plant for CO_2 capture: Process modeling and technical considerations", *Fuel*, **153** 210-223 (2015).

[6] Zhou, W., Jiang, D., Chen, D., Griffy-Brown, C., Jin, Y. and Zhu, B., "Capturing CO_2 from cement plants: A priority for reducing CO_2 emissions in China", *Energy*, **106** 464-474 (2016).

[7] Leeson, D., Mac Dowell, N., Shah, N., Petit, C. and Fennell, P.S., "A Techno-economic analysis and systematic review of carbon capture and storage (CCS) applied to the iron and steel, cement, oil refining and pulp and paper industries, as well as other high purity sources", *International Journal of Greenhouse Gas Control*, **61** 71-84 (2017).

[8] Hornberger, M., Spörl, R. and Scheffknecht, G., "Calcium Looping for CO_2 Capture in Cement Plants – Pilot Scale Test", *Energy Procedia*, **114** 6171-6174 (2017).

[9] De Lena, E., Spinelli, M. and Romano, M.C., "CO_2 capture in cement plants by “Tail-End” Calcium Looping process", *Energy Procedia*, **148** 186-193 (2018).

[10] Ishak, S.A. and Hashim, H., "Low carbon measures for cement plant – a review", *Journal of Cleaner Production*, **103** 260-274 (2015).

[11] Hills, T., Florin, N. and Fennell, P.S., "Decarbonising the cement sector: A bottom-up model for optimising carbon capture application in the UK", *Journal of Cleaner Production*, **139** 1351-1361 (2016).

[12] Diego, M.E., Arias, B. and Abanades, J.C., "Analysis of a double calcium loop process configuration for CO_2 capture in cement plants", *Journal of Cleaner Production*, **117** 110-121 (2016).

[13] Jakobsen, J., Roussanaly, S. and Anantharaman, R., "A techno-economic case study of CO_2 capture, transport and storage chain from a cement plant in Norway", *Journal of Cleaner Production*, **144** 523-539 (2017).

[14] Salas, D.A., Ramirez, A.D., Rodríguez, C.R., Petroche, D.M., Boero, A.J. and Duque-Rivera, J., "Environmental impacts, life cycle assessment and potential improvement measures for cement production: a literature review", *Journal of Cleaner Production*, **113** 114-122 (2016).

[15] Kajaste, R. and Hurme, M., "Cement industry greenhouse gas emissions – management options and abatement cost", *Journal of Cleaner Production*, **112** 4041-4052 (2016).

[16] Schneider, M., "Process technology for efficient and sustainable cement production", *Cement and Concrete Research*, **78** 14-23 (2015).

[17] Jordal, K., Voldsund, M., Størset, S., Fleiger, K., Ruppert, J., Spörl, R., Hornberger, M. and Cinti, G., "CEMCAP – Making CO_2 Capture Retrofittable to Cement Plants", *Energy Procedia*, **114** 6175-6180 (2017).

[18] Carrasco-Maldonado, F., Spörl, R., Fleiger, K., Hoenig, V., Maier, J. and Scheffknecht, G., "Oxy-fuel combustion technology for cement production – State of the art research and technology development", *International Journal of Greenhouse Gas Control*, **45** 189-199 (2016).

[19] Koytsoumpa, E.I., Bergins, C. and Kakaras, E., "The CO_2 economy: Review of CO_2 capture and reuse technologies", *The Journal of Supercritical Fluids*, **132** 3-16 (2018).

[20] Hills, T.P., Sceats, M., Rennie, D. and Fennell, P., "LEILAC: Low Cost CO_2 Capture for the Cement and Lime Industries", *Energy Procedia*, **114** 6166-6170 (2017).

[21] Gardarsdottir, S., De Lena, E., Romano, M., Roussanaly, S., Voldsund, M., Pérez-Calvo, J.-F., Berstad, D., Fu, C., Anantharaman, R., Sutter, D., Gazzani, M., Mazzotti, M. and Cinti, G., "Comparison of Technologies for CO_2 Capture from Cement Production—Part 2: Cost Analysis", *Energies*, **12** (3): 542 (2019).

[22] Pan, S.-Y., Lorente Lafuente, A.M. and Chiang, P.-C., "Engineering, environmental and economic performance evaluation of high-gravity carbonation process for carbon capture and utilization", *Applied Energy*, **170** 269-277 (2016).

[23] Roussanaly, S. and Anantharaman, R., "Cost-optimal CO_2 capture ratio for membrane-based capture from different CO_2 sources", *Chemical Engineering Journal*, **327** 618-628 (2017).

[24] Tsunatu, D., Mohammed-Dabo, I. and Waziri, S., "Technical Evaluation of Selexol-Based CO_2 Capture Process for a Cement Plant", *British Journal of Environment and Climate Change*, **5** (1): 52-63 (2015).