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Design of Major Components of Green Urea Production Plant

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Abstract

Green urea refers to urea produced using renewable energy sources such as municipal solid wastes (MSW) to create a more sustainable and eco-friendly approach to waste management and urea production. As the economy, population, and living standards proliferate swiftly, municipal solid waste has become a significant challenge, particularly in developing countries such as Nepal. This project aims to size the major components of a green urea plant that uses decomposable and combustible waste as feedstock. 15 CSTR digesters with a volume of 53323.2 m^3 produced 41473.2 m3 of biogas to convert 740.6 Tons Per Day (TPD) of decomposable waste and 220.06 TPD of combustible waste into 574.4 TPD of green urea. A circulating fluidized bed gasifier, a Proton Exchange Membrane (PEM) electrolyzer, and an air separation unit using the cryogenic process were also selected. The study also involved picking a tubular steam methane reformer as the primary and an auto-thermal steam methane reformer as the secondary steam methane reformer. The shift reaction was carried out using a catalytic water gas shift reactor, and a three-bed type ammonia reactor was chosen for the Haber Bosch process. Finally, the Snamprogetti process was selected for urea synthesis.

In addition, the study involves the preparation of a 2D system layout to understand the process flow and a 3D solid works drawing to visualize the plant design. The financial analysis of the green urea plant includes the calculation of CAPEX, and OPEX and it is found that total capital investment is 431.45 M$ and operating expenses are 102.2M$/year. Gasification, steam methane reformer, Air separation unit, ammonia synthesis, urea synthesis, electrolyzer, water gas shift reactor, and anaerobic digestion plant consist 7.15%, 13.55%, 3.32%, 16.13%, 26.12%, 13.37%, 13.43% and 6.93% of total capital investment respectively and 14.62%, 1.61%, 7.78%, 19.73%, 5.71%, 32.6%, 12.9% and 14.62% of total operating costs per year.

***Keywords**:* Green Urea, Municipal Solid Waste, CSTR, Methane Reformer, Air separating Unit, Cryogenic

1. Introduction

Urea, with the chemical formula (CH_4N_2O), serves as a fundamental compound crucial for various applications, spanning agriculture, industry, and biochemistry. A colorless, odorless, and highly soluble organic compound is commonly used as a nitrogen fertilizer in agriculture. Urea contains approximately 46% nitrogen by weight, making it an efficient source of nitrogen for plant growth. When applied to soil, urea is broken down by soil microorganisms into ammonium and eventually into nitrate, which is then taken up by plants.

Nepal is an agricultural country. About 68% of Nepal's population depends upon agriculture to sustain their livelihoods (Central Bureau of Statistics, 2020) . However, due to insufficient fertilizer availability, meeting the demand of the population becomes challenging. The annual demand for urea fertilizer is about seven lakh metric tons (700,000 MT), still due to the lack of a proper fertilizer production plant in Nepal, the government has to depend on imports for the supply (Investment Board Nepal, 2017)**.** The government has not been able to supply timely and on a requisite quantity as indicated by Figure. Due to inadequate and untimely supply

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of required fertilizer, the agricultural output has been suppressed and the country is forced on importing a lot of food products.

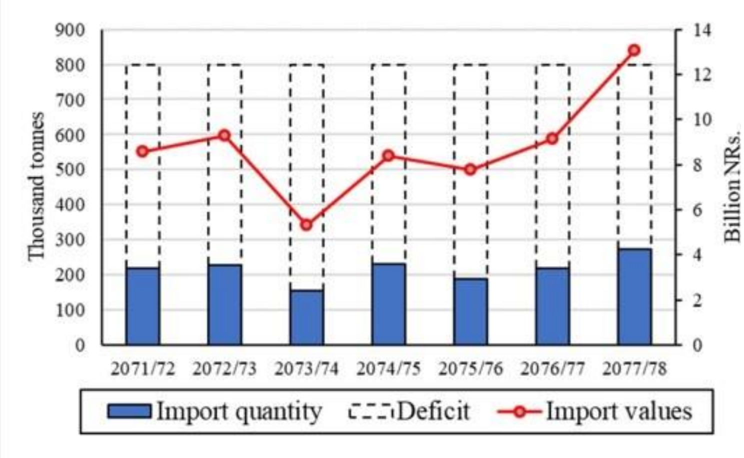

Figure 1. Urea deficit trend in Nepal (MOLAD, 2020)

1.1 Objectives

Our project's main objective is to design major components of green urea plants in Nepal. To achieve this objective, the following specific objectives are outlined.

- To model the system layout and model of the process involved
- To perform sizing and selection of significant components
- To estimate the cost of green urea plant

1.2 Problem Statement

The pressing challenge faced by Nepal in producing urea due to the lack of long-term natural gas supplies necessitates a shift towards alternative and sustainable methods. Importing natural gas for urea manufacturing not only perpetuates the nation's dependency on imports but also raises concerns regarding fertilizer security, the capital-intensive nature of cross-border pipelines, and import imbalances (Investment Board Nepal, 2017). Furthermore, constructing a urea plant using fossil fuels contradicts the global initiative to reduce carbon emissions. In response to these challenges, the Nepali government is actively exploring ways to produce urea locally without relying on fossil fuels. The potential establishment of green urea plants in Nepal becomes a crucial consideration. Leveraging the significant issue of municipal solid waste, particularly in the Kathmandu Valley, which produces approximately 1045 TPD of MSW, offers a sustainable solution (Koo, Adhikari and Mali, 2022)**.**

A comprehensive feasibility study for the design of major components for green urea plants in Nepal is currently non-existent. Currently, there is a dearth of research focused on modeling the system layout and processes involved in green urea production within the Nepalese context. Additionally, there has been no analysis conducted to perform sizing and selection procedures for the crucial components required for efficient plant operation. Furthermore, there is a notable absence of efforts aimed at estimating the costs associated with establishing and operating green urea plants in Nepal.

2 Research Methodology

2.1 Selection of best type among different types of process

The selection of the best type of process for each major component within the green urea production plant involves a comprehensive evaluation process informed by a thorough literature review. Initially, a comprehensive understanding of diverse processes including gasification, electrolysis, air separation, steam methane reforming, shift reaction, ammonia synthesis, and urea synthesis is achieved through an extensive literature review. This serves as the cornerstone for assessing the myriad technological options available in the field. Subsequently, a meticulous evaluation of the advantages and disadvantages associated with each process is conducted, with a nuanced analysis of factors such as efficiency, environmental impact, and economic considerations. By weighing the merits and drawbacks of different technologies, the most suitable processes for each major component are identified. Technical feasibility is calculated through assessing factors such as process efficiency, scalability, technology readiness level, and compatibility with existing infrastructure, ensuring that the selected processes align with the overall goals and values of green urea production while being operationally viable.

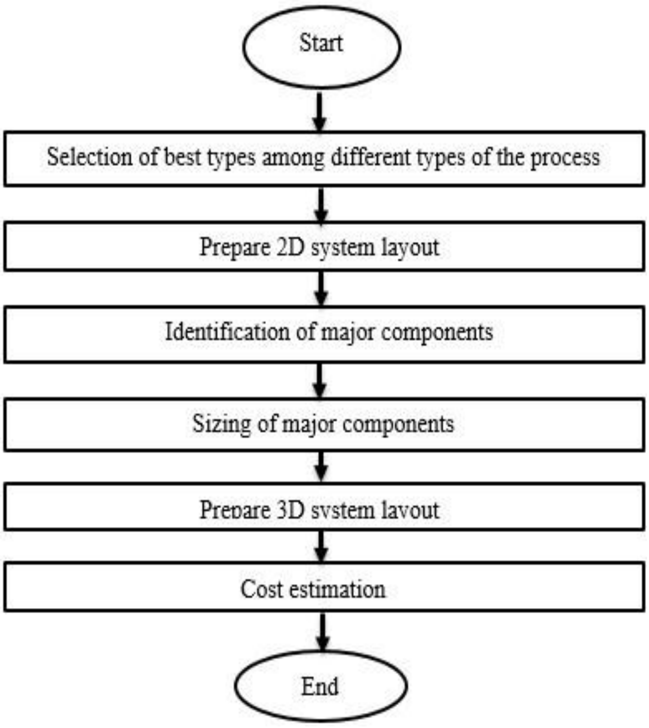

Figure 2. Methodology

2.2 Prepare 2D system Layout

The second phase of the methodology is centered around developing a 2D system layout for the major components of the green urea production plant. The primary objective is constructing a two-dimensional representation that facilitates a comprehensive understanding of the overall process flow. This involves the utilization of process flow diagrams and equipment specifications to craft a schematic layout of the plant. The process further includes the identification of interconnections between major components and establishing a coherent sequence of operations. The 2D system layout serves as a foundational visual tool, enabling stakeholders to grasp the intricacies of the green urea production process and aiding in subsequent design and planning phases.

2.3 Identification of Major Components

In the third step of the methodology, the focus is on the identifying of Major Components crucial for the green urea production plant. The objective is to compile a comprehensive list of critical components integral to the overall process, with consideration given to the selected technologies. It involves categorizing the components into distinct sections such as gasification, electrolysis, air separation, steam methane reforming, shift reaction, ammonia synthesis, urea synthesis, and anaerobic digestion. By systematically categorizing and listing these components, the methodology aims to provide a clear inventory of essential elements that constitute the core infrastructure of the green urea production plant.

2.4 Sizing of Major Components

In the fourth phase of the methodology, the goal is to determine the suitable size and capacity of each major component for achieving the desired green urea production targets. It involves employing process simulation tools such as ANSYS to analyze feedstock requirements and output for each component. The sizing is then optimized by considering factors such as efficiency, production capacity, and economic considerations. Additionally, the methodology includes the crucial step of validating sizing calculations through sensitivity analyses and simulation studies, ensuring robust and reliable results in the determination of component dimensions.

2.5 Prepare 3D system layout

In the fifth step, the methodology aims to create a detailed three-dimensional (3D) representation of the green urea production plant to enhance visualization. It involves utilizing advanced design software like SolidWorks to generate a 3D model based on the 2D layout. The process includes incorporating detailed representations of equipment, pipelines, and structural elements to provide a comprehensive view of the plant's physical configuration. Importantly, the design ensures spatial coordination and accessibility, addressing maintenance and operational requirements effectively.

2.6 Cost Estimation

In the final step of the methodology, the objective is to estimate both the capital and operating costs associated with each principal component and the overall green urea production plant. This involves collecting data on equipment costs, construction expenses, and operational expenditures. Various cost estimation methodologies, including factored estimates, parametric models, or vendor quotations, are employed. The methodology concludes by summarizing the cost breakdown for each major component and conducting an evaluation of the financial feasibility of the entire plant.

3 Literature Review

3.1 JICA Report **(*Japan International Cooperation Agency, 1984)***

The feasibility study conducted by the Japan International Cooperation Agency (JICA) in 1984, recommended using air fractionalization, electrolysis, and cement factory flue gases for establishing a urea plant in Nepal, a recommendation that has not been practically realized to date. According to the JICA report (Japan International Cooperation Agency (JICA), 1984) , the total financing required for the establishment of the urea plant is 145 M$ with the utilization of flue gas from the cement factory. It uses a nitrogen plant capacity of 132 TPD, an Ammonia plant of 160 TPD, a Urea plant capacity of 28.4 TPD, a hydrogen plant capacity of 28.4 TPD, a CO_2 plant of 207 TPD, and 86 MW electricity for the production of 275 TPD of urea.

3.2 **IBN Report (Investment Board of Nepal, 2014)**

Detailed Feasibility study for setting up of Urea Fertilizer Plant in Nepal conducted by Office of Investment Board Government of Nepal (IBN) (Investment Board Nepal, 2014) states A 7 lakh MTPA urea fertilizer plant would mean setting up an ammonia plant of 1220 MTPD and a urea plant of 2125 MTPD along with other OSBL and offsite facilities. The OSBL plant configuration encompasses an internal steam and power generation plant of a capacity of 16 MW to meet the electricity demand for the ammonia and urea plant. The raw water requirement for the fertilizer plant has been estimated at 747 m3/hr. Other OSBL facilities include urea storage facility, effluent treatment plant, cooling water facility, etc. The project cost for the urea fertilizer plant is estimated at US$ 665 million. The estimated project cost for the same plant configuration using coal gasification and water electrolysis is US$ 995 million and US$ 1305 million, respectively. In addition to high capital costs, plants based on coal and water electrolysis are also characterized by several technological challenges and lesser cost efficiencies compared to natural gas. The cost of production for 7 lakh MTPA of urea is estimated to be US$ 187.6 million at 100% operation and per unit cost of production of urea is US$ 268 for natural gas as feedstock. The sales price of urea is estimated at US$ 285 per MT which is less than US$ 303 per MT, the price at which the urea is currently being imported in Nepal.

3.3 Techno-Economic Assessment of Green Urea Production in Nepal Utilizing Municipal Solid Waste and Hydroelectricity

A thesis named "Techno-Economic Assessment of Green Urea Production in Nepal Utilizing Municipal Solid Waste and Hydroelectricity" by Saroj Karki .

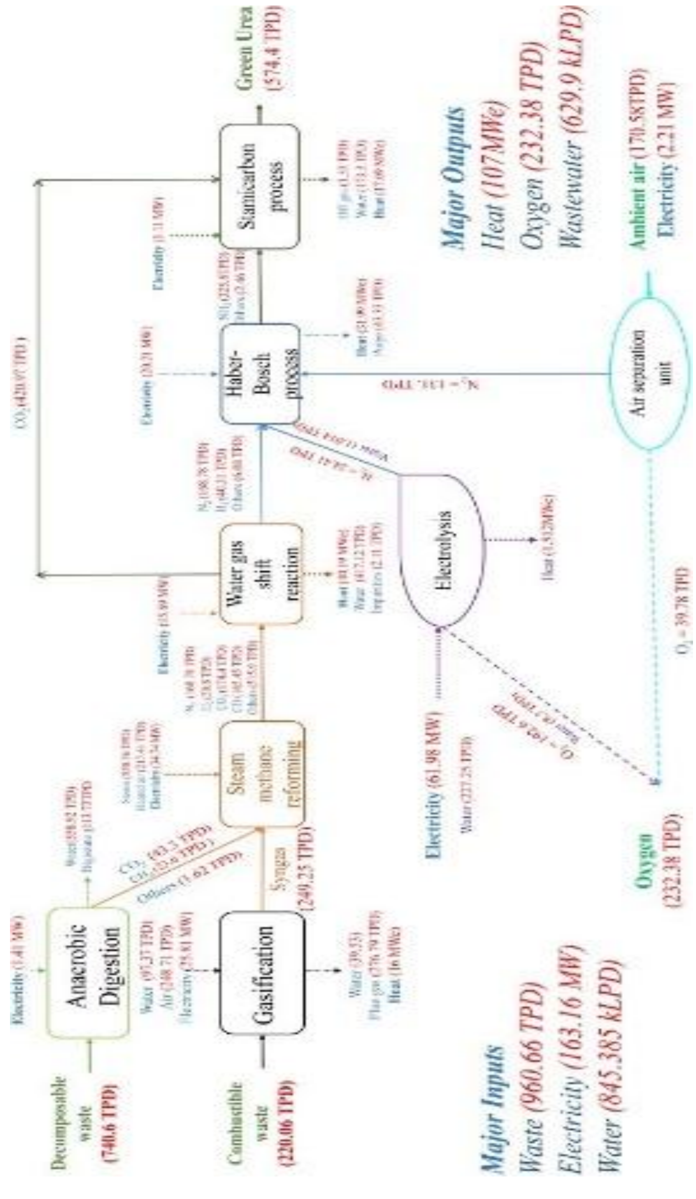

Figure 3. Schematic diagram showing entire inputs and outputs of municipal solid waste-based green urea synthesis process executed in Aspen Plus simulation software (Karki et al., 2022)

3.4. The 2D and 3D layout of Green Urea plant

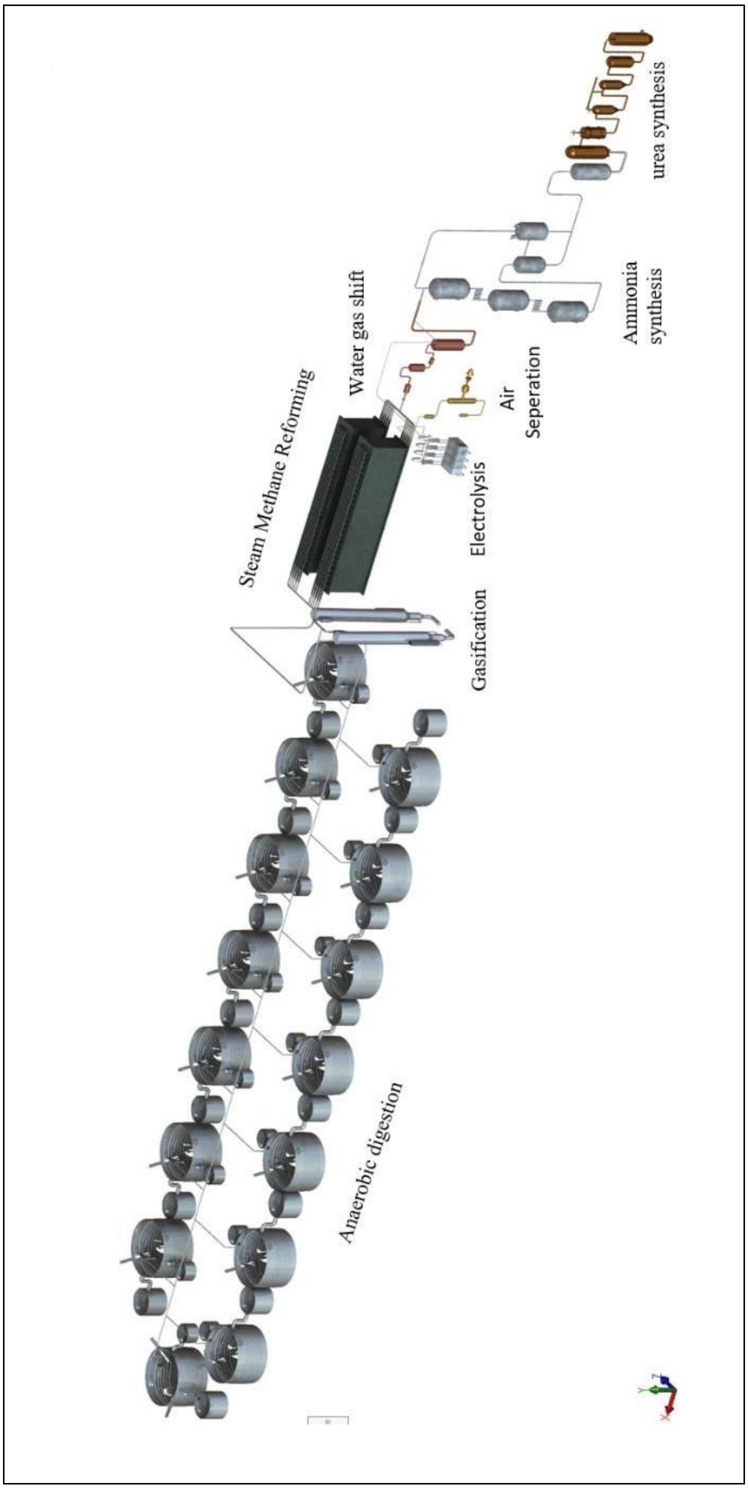

Figure 4. 3D diagram of green urea plant

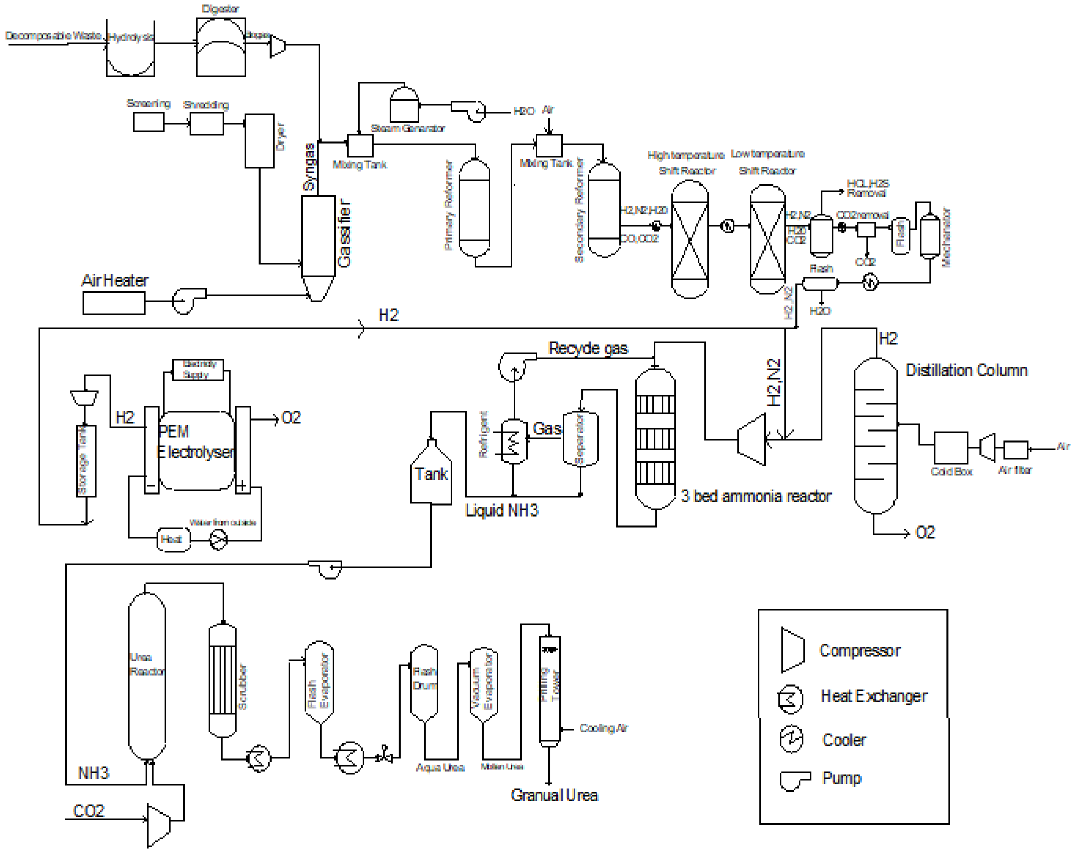

Figure 5. 2D layout of green urea plant

4. Process description

4.1 Anaerobic Digestion

Anaerobic digestion is the natural process that breaks down organic matter (OM) without oxygen to release a gas known as biogas, leaving an organic residue called digestate (Global Methane Initiative, 2016) . The digester of a CSTR plant is generally cylindrical and made of concrete or steel. The dimensions of the digester tank, feeding tank, Residue tank, and Lagoon Pit are calculated as (AEPC, Design construction and manual of double membrane continuous stirred tank reactor(AEPC, 2019).

Total daily feed (Qi) = Wastes + water

Total volume of substrate =Qi × HRT (Equation 1)

The final volume of Digester (Vd) = 20 − 30% more than the volume of substrate

$$\text{Volume}=\frac{\pi\times D\times D\times H}{4} \qquad \text{(Equation 2)}$$

$$\text{No of digester} = \frac{Volume\ of\ Digester \times 4}{\pi \times D \times D \times H} \qquad \text{(Equation 3)}$$

Amount of Gas = Biogas Yield for MSW wastes × OLR × Substrate volume

The volume of feeding tank = half the volume of daily feed

Internal volume of feeding tank = volume of substrates that should be stored in the tank by 1.20 to consider 20% as headspace.

The volume of the Residue Tank = 4 × volume of the feeding tank. (Equation 4)

The volume of liquid slurry pit = 25 % of daily slurry volume × 2 days for retention time (Equation 5)

In the digester and mixing tank, inclined shaft agitators with a propeller (Marine type) are selected to achieve better performance, offering several advantages.

Incline shaft agitators

- The motor rating of such agitators ranges from 7.5kW to 22kW which is sufficientfor handling significant amount of substrate. (AEPC, 2019)
- This type of agitators can be easily mounted and dismounted for repair and maintenance without the need to tamper with the roof membrane and therefore not affecting the gas generation process.
- This agitator design is quite popular with almost 13% of biogas plants in Germany using such system.

Propeller (Marine Type)

- Marine type propellers are designed for high efficiency mixing which ensures uniform mixing of the substrate and efficient transfer of heat and mass in the reactor.
- Marine type propellers are designed to provide high mixing efficiency while consuming low energy. This leads to lower operating costs and reduced carbon footprint.

Screw press separators are also used for the separation of slurry and digester solids. Due to varying climatic conditions, maintaining the mesophilic temperature within the digester requires the installation of heating systems (AEPC, 2019) .

4.2 Gasification

Gasification is the process that converts carbonaceous materials into gasses, including the most significant fractions of nitrogen, carbon monoxide, hydrogen, and carbon dioxide. The gasification process is carried out at a temperature range of 500 – 1500^{o}C.To gasify municipal solid wastes, circulating fluidized bed gasifiers are chosen as the optimal choice among gasifier types (Wen-Ching Yang (Ed.), 2003) .

Terminal velocity is calculated (Harald Drück *et al.*, 2020) .

$$\text{Terminal velocity} = \left[\frac{g {d_p}^{1.6}(\rho_p - \rho_{g)}}{13.9\ {\rho_g}^{0.4} {\mu_g}^{0.6}}\right]^{0.71} \qquad \text{(Equation 6)}$$

Where g = acceleration due to gravity

d_p = diameter of particles

ρ_p = density of particles

ρ_g = density of fluidizing gas

μ_g = Viscosity of gas

Stoichiometric air (mth) is determined by using the equation.(Harald Drück *et al.*, 2020) .

Stoichiometric air = 0.1153C + 0.3434(H - O/8) + 0.0434S (Equation 7)

Where C, H, O and S are the percentages of Carbon, Hydrogen, Oxygen and Sulphur respectively.

$$\text{cross-sectional area of riser} = \frac{m_a}{\rho_g \times U_g} \qquad \text{(Equation 8)}$$

ρ_g = density of air at 850^0C = 0.3145 kg/s

U_g = fluidization velocity = 5 m/s

4.3 Air separation unit

An air separation unit (ASU) is a system that separates air into its constituent parts, primarily oxygen, nitrogen, and sometimes argon, through a process known as cryogenic distillation (Al-Haj Ibrahim and Sabagh, 2014) .

$$K = 3600\ (-0.17T^2 + 0.27T - 0.047) \qquad \text{(Equation 9)}$$

$$u_v = \frac{K}{3600}\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{0.5} \qquad \text{(Equation 10)}$$

$$D = \left(\frac{4V_w}{3600\pi\rho_v u_v}\right)^{0.5} \qquad \text{(Equation 11)}$$

Where,

k = A factor that depends on plate spacing

T= plate spacing (m)

U_v = Maximum allowable vapor velocity(m/sec)

ρ_l = Liquid density(kg/m^3)

ρ_v = Vapour density(kg/m^3)

D = diameter (m)

V_w = vapor rate (kg/hr)

4.4 Electrolysis

PEM electrolysis is considered a splendid method for high-purity hydrogen production in future industrial applications, due to its high current density, greater energy efficiency, smaller gas crossover, wider operating temperatures (20–80°C), smaller mass-volume characteristic, more importantly, the specialty of adaptive to renewable energy volatility (Zhang *et al.*, 2012) .

The reversible voltage of a PEM water electrolysis process can be determined by the Nernst equation, i.e.,

$$V_O = 1.229 - 8.5 \times 10^{-4}(T - T_O) + 4.3085 \times 10^{-5} \times T \times ln\left(\frac{P_{H_2}\sqrt{P_{O_2}}}{P_{H_2O}}\right) \qquad \text{(Equation 12)}$$

Where,

P_{H_2} & P_{O_2} $are\ the\ partial\ pressures\ of\ hydrogen$

$and\ oxygen, respectively.$

$$V_{act} = \left(\frac{\alpha_A+\alpha_C}{\alpha_A\times\alpha_C}\right)\times\frac{RT}{2F}\times ln(\frac{j}{j_0}) \quad \text{(Equation 13)}$$

Where,

$$j_0 = 1.08\times10^{-17}\ exp(0.086T) is\ the\ exchange\ current\ density$$

j = Current density

R = Universal gas constant.

F = Faraday's constant

$$\alpha_A = Charge\ transfer\ coeff. For\ Anode.$$

$$\alpha_C = Charge\ transfer\ coeff. For\ Cathode.$$

4.5 Steam Methane Reformer

The steam methane reforming process is the process of breaking hydrocarbons into hydrogen and carbon monoxide from the catalytic conversion in the presence of steam. The steam reacts with methane at very high temperatures and pressure.
Reactor size depends upon the number of tubes used at the reformer. The formula related to the calculation of reformer size is taken from (Eyalarasan *et al.*, 2013) .

$$Space\ velocity = \frac{Volumetric\ flow\ rate\ of\ feed}{Volume\ of\ packing} \quad \text{(Equation 14)}$$

$$Void\ fraction\ (\epsilon) = \frac{V_R - V_P}{V_R} \quad \text{(Equation 15)}$$

Where V_R = Volume of reactor

V_p = Volume of packing

4.6 Water gas shift reaction

The water gas shift reaction (also known as the WGS reaction) is a chemical reaction that involves the conversion of carbon monoxide (CO) and water (H_2O) into carbon dioxide (CO_2) and hydrogen gas (H_2). Different catalysts, such as iron oxide (Fe_2O_3), chromium oxide (Cr_2O_3), copper, zinc, are used. In a fluidized-bed reactor WGS reactor, the catalyst particles are fluidized by the reactants, which allows for better heat and mass transfer. The fluidized-bed reactor can operate at lower temperatures and is less prone to catalyst deactivation than to the fixed-bed reactor.

The chemical equation for the water gas shift reaction is

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

Sizing of piping for boiling feed water inlet and supersaturated steam outlet. (Devkota, 2019)

$$Volumetric\ flow\ rate\ = \frac{Mass\ flow\ rate(\frac{Kg}{hr})}{Density(\frac{Kg}{m3})} \quad \text{(Equation 16)}$$

$$The\ volume\ of\ the\ catalyst\ bed = \frac{Volumetric\ flow\ rate}{Space\ velocity} \quad \text{(Equation 17)}$$

$$No.of\ tubes\ (n)\ =\ \frac{\pi \times dtube \times dtube \times Ltube}{4} \times \rho cat \quad \text{(Equation 18)}$$

4.7 Ammonia Synthesis Plant

The ammonia reactor comprises three catalyst beds and two heat exchangers as inter-stage coolers. To maintain the catalyst beds at the optimal temperature for achieving maximum conversion, the feed gas is supplied and distributed as a quench in the space between the catalyst beds. In the ammonia converter, the top bed is arranged to have the lowest quantity of catalyst to limit the temperature rise before the first inter-stage cooler.

To calculate the diameter of this shell, the below equation is used (Mohammed, 2021) .

$$D_s = d_{tube} x \sqrt{(\frac{n_{tubes}}{\phi})} \quad \text{(Equation 19)}$$

Where,

D_s = Shell Diameter.

ϕ = Perforation Factor.

$$\text{The total length of the reactor} = L_{tube} + D_s \quad \text{(Equation 20)}$$

$$\text{Diameter of the bed } (D_{bed}) = \left(\frac{4V_{cat}}{L_{bed}}\right)^{0.5} \quad \text{(Equation 21)}$$

Input Data extracted from Aspen-plus modeling. (Karki *et al.*, 2022)

Table 1. Characteristics data of bed reactor

	First Bed Reactor	**Second Bed Reactor**	**Third Bed Reactor**
Inlet Temperature	673 K	696 K	698.5 K
Outlet Temperature	746 K	736 K	732.96 K
Operating Pressure	226.981 bar	225.981 bar	224.981 bar
Mass of Catalyst	16973.4 kg	16973.4 kg	16973.4 kg
Conversion	11%	11%	11%

4.8 Urea synthesis plant

Urea, an organic white compound, is manufactured worldwide in various shapes and sizes, from ammonia and carbon dioxide. The Snamprogetti ammonia stripping process involves two main reactions in the synthesis of urea from carbon dioxide and ammonia: the formation of ammonium carbamate from carbon dioxide and ammonia, and the conversion of ammonium carbamate into urea and water.

Now, for the volume of the reactor (V), the used formula is given below (Muhammad Muhaimin Binabd Rahman, 2013)

$$\frac{V}{v_0} = \frac{[-ln(1-x_a)]}{k} \quad \text{(Equation 22)}$$

Where, v_0 = Volumetric flow rate

X_a = Composition of carbamate

Number of stages Required

$$N = \frac{H}{spacing} \qquad \text{(Equation 23)}$$

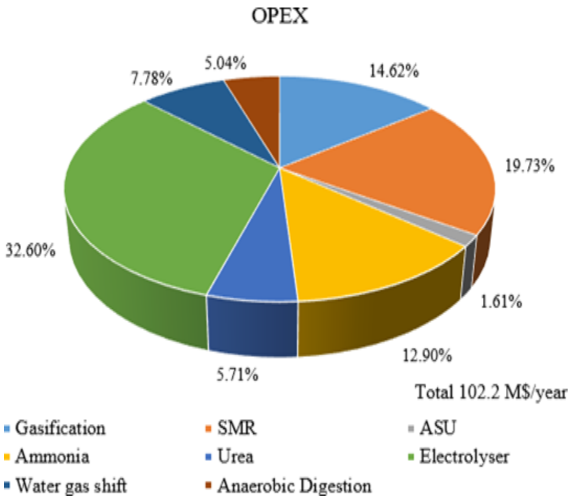

Figure 6. Pie-chart of Operating cost

5. Cost Estimation

In cost estimation, different parameters like total investment cost (CAPEX), and annual operating expenditure (OPEX) are determined (Salman *et al.,* 2017)

$$C = C_o \times \left(\frac{S}{S_o}\right)^d \times \left(\frac{CEPCI_y}{CEPCI_x}\right) \qquad \text{(Equation 24)}$$

C = Cost of equipment at present year
Co= Cost of equipment at reference year
S = Capacity of equipment at present year
So = Capacity of equipment at reference year
CEPCIy = Chemical engineering plant cost index of the preset year
CEPCIx = Chemical engineering plant cost index of the reference year

The total capital investment is found to be 431.45 M$, where gasification, air separation, water gas shift unit, steam methane reforming, urea synthesis, electrolyzer, ammonia synthesis, and anaerobic digestion contribute 8.18%, 3.18%, 1.43%, 15.16%, 29.89 %,15.30%, 18.46%, and 8.39% respectively. Similarly, the Operating cost comes to 102.2 M$/year, where 14.62%, 1.61%, 7.78%, 19.73%, 5.71%, 32.6%, 12.9%, and 14.62% gasification, air separation, water gas shift unit, steam methane reforming, urea synthesis, electrolyzer, ammonia synthesis, and anaerobic digestion contributes respectively. The payback period is calculated as 10.87 years, and the levelized cost of urea is $778.93 per metric ton.

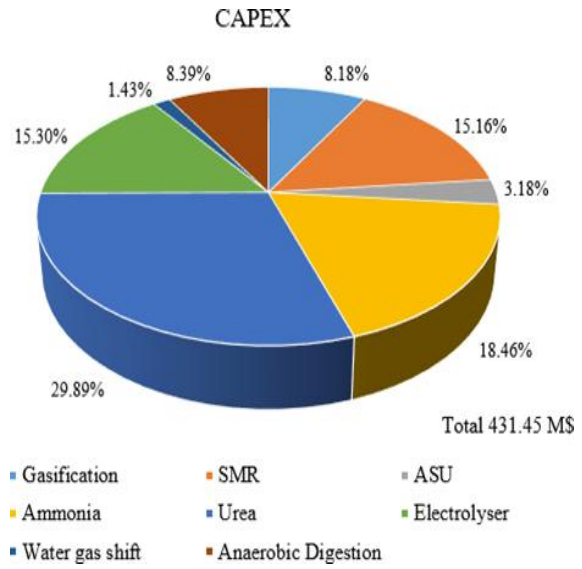

Figure 7. Pie-chart of Capital cost

Table 2. CAPEX and OPEX calculations

S.N.	Particulars		Gasification	SMR	ASU	Ammonia	Urea	Electrolyser	Water gas shift	Anaerobic Digestion	Total
A	Capital investment										
1	Inside Battery Limit Cost (ISBL)	% of equipment cost	$22,831,982.98	$38,997,896.39	$8,865,476.00	$49,031,100.00	$77,201,908.30	$42,692,601.80	$3,995,942.71	$23,420,101.12	$267,037,009.30
1.1	Equipment cost		$14,542,664.32	$24,839,424.45	$5,646,800.00	$31,230,000.00	$49,173,190.00	$27,192,740.00	$2,545,186.44	$14,917,261.86	
1.2	Buildings	2.50%	$363,566.61	$620,985.61	$141,170.00	$780,750.00	$1,229,329.75	$679,818.50	$63,629.66	$372,931.55	
1.3	Equipment foundation and support	13.50%	$1,963,259.68	$3,353,322.30	$762,318.00	$4,216,050.00	$6,638,380.65	$3,671,019.90	$343,600.17	$2,013,830.35	
1.4	Piping	8.00%	$1,163,413.15	$1,987,153.96	$451,744.00	$2,498,400.00	$3,933,855.20	$2,175,419.20	$203,614.92	$1,193,380.95	
1.5	Electrical	9.00%	$1,308,839.79	$2,235,548.20	$508,212.00	$2,810,700.00	$4,425,587.10	$2,447,346.60	$229,066.78	$1,342,553.57	
1.6	Instrumentation	5.00%	$727,133.22	$1,241,971.22	$282,340.00	$1,561,500.00	$2,458,659.50	$1,359,637.00	$127,259.32	$745,863.09	
1.7	Process insulation	2.00%	$290,853.29	$496,788.49	$112,936.00	$624,600.00	$983,463.80	$543,854.80	$50,903.73	$298,345.24	
1.8	Ocean fright	6.00%	$872,559.86	$1,490,365.47	$338,808.00	$1,873,800.00	$2,950,391.40	$1,631,564.40	$152,711.19	$895,035.71	
1.9	Port and handling	2.00%	$290,853.29	$496,788.49	$112,936.00	$624,600.00	$983,463.80	$543,854.80	$50,903.73	$298,345.24	
1.1	Custom duty and imports	5.00%	$727,133.22	$1,241,971.22	$282,340.00	$1,561,500.00	$2,458,659.50	$1,359,637.00	$127,259.32	$745,863.09	
1.11	Site preparation	0.50%	$72,713.32	$124,197.12	$28,234.00	$156,150.00	$245,865.95	$135,963.70	$12,725.93	$74,586.31	
1.12	Inland fright and insurance	3.50%	$508,993.25	$869,379.86	$197,638.00	$1,093,050.00	$1,721,061.65	$951,745.90	$89,081.53	$522,104.17	
2	Outside Battery	% of	$4,223,916.8	$11,114,	$1,640,1	$12,012,	$21,616,5	$7,898,1	$739,24	$4,332,718	$63,57

	limit cost (OSBL)	ISBL	5	400.47	13.06	619.50	34.32	31.33	9.40	.71	7,683.65
2.1	Cooling water system	3.00%	$684,959.49	$1,169,936.89	$265,964.28	$1,470,933.00	$2,316,057.25	$1,280,778.05	$119,878.28	$702,603.03	
2.2	Water supply, storage and pretreatment	2.50%	$570,799.57	$974,947.41	$221,636.90	$1,225,777.50	$1,930,047.71	$1,067,315.05	$99,898.57	$585,502.53	
2.3	Yard piping	6.00%	$1,369,918.98	$2,339,873.78	$531,928.56	$2,941,866.00	$4,632,114.50	$2,561,556.11	$239,756.56	$1,405,206.07	
2.4	Auxillary facilities	4.00%	$913,279.32	$1,559,915.86	$354,619.04	$1,961,244.00	$3,088,076.33	$1,707,704.07	$159,837.71	$936,804.04	
2.5	General welfare	1.00%	$228,319.83	$389,978.96	$88,654.76	$490,311.00	$772,019.08	$426,926.02	$39,959.43	$234,201.01	
2.6	Construction facilities	2.00%	$456,639.66	$779,957.93	$177,309.52	$980,622.00	$1,544,038.17	$853,852.04	$79,918.85	$468,402.02	
2.7	Steam generation	10.00%	$0.00	$3,899,789.64	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	
2.8	Urea storage and handling	9.50%	$0.00	$0.00	$0.00	$0.00	$7,334,181.29	$0.00	$0.00	$0.00	
2.9	Ammonia storage and handling facilities	6.00%	$0.00	$0.00	$0.00	$2,941,866.00	$0.00	$0.00	$0.00	$0.00	
	Total cost (ISBL + OSBL)		$27,055,899.83	$50,112,296.86	$10,505,589.06	$61,043,719.50	$98,818,442.62	$50,590,733.13	$4,735,192.12	$27,752,819.83	$330,614,692.95
3	Others	% of (ISBL + OSBL)	$8,252,049.45	$15,284,250.54	$3,204,204.66	$18,618,334.45	$30,139,625.00	$15,430,173.61	$1,444,233.60	$8,464,610.05	$100,837,481.35
3.1	Non-plant buildings and township	3.00%	$811,677.00	$1,503,368.91	$315,167.67	$1,831,311.59	$2,964,553.28	$1,517,721.99	$142,055.76	$832,584.59	
3.2	Engineering fee	15.00%	$4,058,384.98	$7,516,844.53	$1,575,838.36	$9,156,557.93	$14,822,766.39	$7,588,609.97	$710,278.82	$4,162,922.97	
3.3	Project management charge	3.00%	$811,677.00	$1,503,368.91	$315,167.67	$1,831,311.59	$2,964,553.28	$1,517,721.99	$142,055.76	$832,584.59	
3.4	Land development fee	4.00%	$1,082,235.99	$2,004,491.87	$420,223.56	$2,441,748.78	$3,952,737.70	$2,023,629.33	$189,407.68	$1,110,112.79	
3.5	Contingency	5.00%	$1,352,794.99	$2,505,614.84	$525,279.45	$3,052,185.98	$4,940,922.13	$2,529,536.66	$236,759.61	$1,387,640.99	
3.6	Net commissioning expenses	0.50%	$135,279.50	$250,561.48	$52,527.95	$305,218.60	$494,092.21	$252,953.67	$23,675.96	$138,764.10	
	Total investment cost (CAPEX)		$35,307,949.28	$65,396,547.40	$13,709,793.72	$79,662,053.95	$128,958,067.62	$66,020,906.74	$6,179,425.71	$36,217,429.87	$431,452,174.30
B	Operating cost (OPEX)	% of CAPEX									
1	Fixed cost		$1,447,625.92	$2,681,258.44	$562,101.54	$3,266,144.21	$5,287,280.77	$2,706,857.18	$253,356.45	$1,484,914.62	$17,689,539.15
1.1	Cytalyst, chemicals and consumables	0.60%	$211,847.70	$392,379.28	$82,258.76	$477,972.32	$773,748.41	$396,125.44	$37,076.55	$217,304.58	
1.2	Salary, wages and overheads	2.00%	$706,158.99	$1,307,930.95	$274,195.87	$1,593,241.08	$2,579,161.35	$1,320,418.13	$123,588.51	$724,348.60	
1.3	Repair and	1.00%	$353,079.49	$653,9	$137,097	$796,620	$1,289,58	$660,209	$61,794	$362,174.3	

	maintenance			65.47	.94	.54	0.68	.07	.26	0	
1.4	Insurance	0.50%	$176,539.75	$326,982.74	$68,548.97	$398,310.27	$644,790.34	$330,104.53	$30,897.13	$181,087.15	
2	Variable cost										
2.1	Water (Cost in year)	2.5 USD/kL	$80,330.25	$429,627.00	$0.00	$0.00	$0.00	$187,481.25	$0.00	$461,109.00	$1,158,547.50
	Water input	kLPD	97.37	520.76	0	0	0	227.25	0	558.92	
2.2	Waste (Cost in year)	10.27USD/ton	$747,635.46	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$2,509,967.46	$3,257,602.92
	Waste input	TPD	220.6	0	0	0	0	0	0	740.6	
2.3	Electricity	491040$/MWyear	$12,673,742.40	$17,058,729.60	$1,085,198.40	$9,923,918.40	$545,054.40	$30,434,659.20	$7,704,417.60	$692,366.40	$80,118,086.40
	Electricity input	MW	25.81	34.74	2.21	20.21	1.11	61.98	15.69	1.41	

6. Results

The objective of the project work has been completed, and after completion of the entire work, the following conclusions are derived:

1. The approximate riser diameter and height of the circulating fluidized bed gasifier are determined as around 3.5 m and 21 m with a terminal velocity of 4.5 m/s. The total length, the diameter of the gas exit, and the diameter of the dust outlet of the cyclone separator are calculated and found to be 4 m, 0.5 m, and 0.2 m. The length of the vortex finder is 0.625 m.

2. The number of furnaces required to process the given amount of syngas is 2 with 2322 total tubes 1161 tubes in each. The volume of the reactor and mass of the catalyst used is 164.1 m^3 and 101404.9 kg respectively.

3. Maximum allowable velocities for nitrogen and oxygen were identified as 1.22 m/s and 1.285 m/s, respectively, with the distillation column dimensions determined to be 1.3 m in diameter and 8 m in height.

4. The urea reactor size parameters, derived from a volumetric flow rate of 86.34 m3/hr, resulting in a reactor volume of 15.66 m3, with specific dimensions including a diameter of 1.58 m and a height of 7.9 m. The residence time within the reactor is set at 10 minutes, and the reactor design involves 10 stages.

5. Analyzing the operating conditions of 80^0C temperature, the number of electrolysis stacks required to meet our hydrogen production is determined to be 2266 & the power consumed by the stack during electrolysis is 45.44 MW. For the temporary storage of hydrogen gas at a maximum 150 bar pressure, the thickness of HSLA 15CDV6 steel is determined to be 55.476 mm. The diameter and the length of reactor one is 2.274 m & 4.274 m. Similarly, reactor two is 2.54 m & 5.04 m, and likewise for reactor three is 2.641 m & 5.34 m.

6. The CSTR digester system consists of 15 units, with a combined substrate volume totaling 44,436 m^3. Each digester has a final volume of 53,323.2 m^3, with individual dimensions of 24 m in diameter and 8 m in height. The total amount of gas produced is measured at 41,473.2 m^3. Additionally, the feeding tank, designed to support the anaerobic digestion process, boasts an internal volume of 59.248 m^3, with a diameter of 5 m and a height of 3 m. The tank is positioned 1 m above the ground level.

7. The water gas shift reactor system is characterized by distinct parameters for the high-temperature (HT-WGS) and low-temperature (LT-WGS) stages. The volumetric flow rates of the reactant gas are 56,667.62 m^3/hr. for HT-WGS and 566,667.62 m^3/hr. for LT-WGS. The reactor dimensions vary, with HT-WGS having

a volume of 10.29 m^3, a diameter of 2 m, and a height of 3.27 m, while LT-WGS features a larger volume of 51.5 m^3, a diameter of 3 m, and a height of 7.28 m. The tube configurations include 54 tubes for HT-WGS and 190 tubes for LT-WGS, each with lengths of 2.7 m and 5 m, and diameters of 0.15 m. Additionally, the boiler feed water inlet and super-saturated steam outlet pipes have diameters of 1.9 m and 0.38 m, respectively.

8. The capital expenditure (CAPEX) needed for constructing the plant, along with the ongoing operating expenditure (OPEX) for labor, maintenance, and raw materials, were assessed. They were found to be 431.45 M$ and 112.65 M$/year, respectively. The payback period for the plant is 10.87 years, with a levelized cost of urea at $778.93 per metric ton.

7. Conclusions

This research study offers 2D and 3D modal visualization and calculation of sizing of major components used in green urea production plant. This project entails a total estimated capital investment of 431.45 million dollars and estimated operating cost 112.65 million dollar per year.

8. Recommendation

Based on the above conclusion, the following recommendations are set forward:

- Process simulation could be done to obtain precise results.
- Collaborating with experts and professionals in the urea production field to gain deeper insight into the technical requirements and challenges associated with establishing a green urea plant.
- Conduct a thorough environmental impact assessment, including potential emissions and their effects on the local ecosystem.
- A thorough feasibility study of the green urea plant can be conducted by examining all minor components comprehensively.
- Identify potential risks and challenges associated with establishing a green urea plant and develop mitigation strategies.

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