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Performance Evaluation of Flow Digesters Design for Optimum Production of Biogas from the Decomposition of Glucose

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ABSTRACT

In a quest to reduce the high reliance on the traditional non-renewable energy source (oil, coal and gas), which poses great environmental, social and human costs, this research is geared towards ensuring that energy of the future must be regenerative (renewable) and sustainable for global economic advancement. The research considered the performance evaluation of flow digesters design that is, the continuous stirred tank reactor (CSTR) and plug flow reactor (PFR) for optimum production of biogas from anaerobic decomposition of glucose. The design models were obtained from the application of the conservation law of mass and energy of the process. The developed performance models were simulated using MATLAB R2023a version at same initial feed rate of 166.752kg/day and operating conditions of the digesters. The performance evaluation of the CSTR and PFR was based on the yield of the biogas and energy efficiency of the process. At a conversion rate of 90%, the CSTR and PFR volume was 10.50m^3 and 2.80m^3 respectively while the quantity of heat generated per unit volume of the CSTR and PFR was 0.50 kK/m^3 and 1.75 kK/m^3 respectively. The performance evaluation of the design results showed that more yield of the biogas is produced in the CSTR as indicated by digester volume while the PFR showed a better performance characteristic in terms of energy efficiency and conservation as shown in the quantity of heat generated. The digester design technology is crucial for renewable energy production and sustainability.

1. INTRODUCTION

Biogas is a renewable source of energy just like bio fuels and bio diesels and constitutes about 70% methane and 45% carbon dioxide and other trace of gases (Abanades *et al*., 2021). It originates from biogenic materials and a type of biofuel produced anaerobically from a wide range of available organic

or biodegradable materials such as animal manure, feaces, sewage sludge, municipal organic waste, plant materials, straw, sugarcane and by products from agricultural and industrial processes (Raja and Wazir, 2017; Durogbade *et al*., 2024; Tambuwal *et al*., 2020; Uheuegbu and Onuora, 2014; Sambo *et al*., 2015).

The discovery of biogas technology as a renewable, clean and environmentally friendly source of energy for domestic and industrial applications can be described as a waste to wealth program for economic advancement. This technology helps to mitigate the adverse effect such as the release of harmful greenhouse gases which causes global warming, environmental pollution from burning or combustion which releases harmful chemicals into the air, including carbon dioxide and sulfur which causes respiratory issues and climatic changes all of which emanates from the use of non-renewable energy sources such as oil, natural gas, nuclear fuel, electricity and coal (Igoni and Harry, 2017; Abdulsalam and Yusuf, 2015; Tamburini *et al*., 2023; Malico *et al*., 2016).

The techno-economic importance of biogas, has attracted several researches on its production and applications and thus Themelis and Ulloa, (2007) researched on the fundamental processes involved in anaerobic biogas production and stated that the process basically involves four steps after the biomass disintegration.

The four steps basically includes; hydrolysis, acidogenesis, acetogenesis and methanogenesis. In the hydrolysis step, the bacteria break or convert large organic polymers (fat, carbohydrate, proteins) into fatty acids, simple sugar and amino acids. The acidogenesis steps involve fermentation process where the low alcohol, volatile fatty acid and organic acids are produced. During the acetogenesis process, the acetogenic bacteria's converts the alcohol and organic acids to acetic acid hydrogen and carbon. While in the methanogenesis step, the acetogenesis products are converted to methane, carbon dioxide and other gases by methanogens bacteria.

Raja and Wazir (2017) stated that the large quantity of agricultural and municipal organic wastes that are discharged into the environment can be beneficial in the production of biogas which is a renewable source of energy and at the same time be an environmental waste management strategy in the world. According to Raja and Wazir, this technology may further optimize the promotion and development of agricultural and animal husbandry in rural areas which will improve the standard of living. Raharjo *et al*., (2021) in their research stated that homemade bio-activator in semi-continuous digesters can be utilized for biogas production from household food waste and discovered that the home bio-activator showed a better performance characteristics in terms of methane yield or recovery when compared to other commercial activators under the same process condition. Organic waste can be treated for biogas production using anaerobic digestion technology (Ismail and Talib, 2016; Antoine *et al*., 2018; Luea and Cossu, 2015). It is important to note biogas production should depend basically on food waste and should not be a threat to food production (Angelidaki *et al*., 2018; Scarlat *et al*., 2018).

Based on the economic importance of biogas as a renewable energy source which offers multi-purpose solution for transportation and generation of heat, power and electricity which have significantly and conspicuously promoted its global demand as well as sustainability, this research considered the performance evaluation of flow digesters design for optimum production of biogas from the decomposition of glucose. The co-digester design basically involves the application of the conservation law of mass and energy for the development of the flow digesters (continuous stirred tank reactor and plug flow reactor) performance model for size determination. The conservation law of mass and energy is the first principle that governs all equipment design and is crucial for all industrial processes involving the transformation of raw materials to finished products (Wosu, 2024a; Wosu, 2024b; Wosu, 2024c; Wosu *et al*., 2024).

2. MATERIALS AND METHODS

2.1. Materials

The research materials are computer set, thermodynamic data, calculated and literature data, the simulation tool used is MATLAB R2023a Version.

2.2. Methods

The research methodology is both quantitative and analytical. The procedures involved are:

i. Process description

ii. Development of reaction chemistry

iii. Development and simulation of the co-digester models.

2.2.1. Process description

The production of biogas from the decomposition of glucose (feed material) obtained from anaerobic processes on biodegradable waste materials resulting to the formation and disintegration of biomass. The biomass is anaerobically converted to glucose through a succession of steps involving hydrolysis which is carried out by bacteria's, acidogenesis through fermentation and acetogenesis caused by acetogenic bacteria. The glucose feed material at initial condition of temperature T_0= 300 K, concentration C_{AO} = 0.04 mol/m^3 and molar flow rate F_{AO} = 0.002 mol/s was feed into the flow digesters operating at a temperature T=310 K where methanogeneic bacteria convert the glucose to carbon dioxide, methane and other trace of gases at an isothermal condition. The methane and carbon dioxide produced are the major constituents of biogas which can be utilized in the generation of energy, power and electricity. The decomposition process is described hypothetically in Figures 2 and 3.

2.2.2. Development of the Decomposition Chemistry

The methanogenic bacteria decomposition process of glucose for biogas production is given in Equation (1).

$$C_6H_{12}O_{6_{(L)}} \xrightarrow{K} 3CO_{2_{(g)}} + 3CH_{4_{(g)}} \tag{1}$$

The rate expression for the methanogenesis process is a pseudo-first order reaction is given as:

$$\left(-r_A\right) = \frac{-dC_A}{dt} = KC_A \tag{2}$$

where $-r_A$ is the depleting rate of glucose in (mol/m^3/s), C_A is the concentration of specie A in mol/m^3, t is the reaction time in seconds and K is the reaction rate constant in sec^{-1}.

Equation (2) can be expressed in terms of fractional conversion (X_A) as;

$$\left(-r_A\right) = K\,C_{Ao}\left(1 - X_A\right) \tag{3}$$

Where C_{A0} is the initial concentration of species A in mol/m^3 and X_A is the fractional conversion.

The rate constant or pre-exponential factor K is given by the modified Gompertz model and substituted into Equation (3) to give;

$$\left(-r_A\right) = Y(t) = A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t) + 1\right)\right] C_{Ao}\left(1 - X_A\right) \tag{4}$$

Where Y(t) is the cumulative of specific biogas production in (mol/m^3/s), A is the pre-exponential or frequency factor in (sec.$^{-1}$), μ_e is the maximum biogas production rate in (sec.$^{-1}$) and λ is the lag phase period in (seconds).

Equation (4) represents the reaction kinetic scheme of the glucose decomposition for biogas production in the co-digester types.

2.2.3. Development of the CSTR design models

Consider the CSTR hypothetical diagram showing the glucose decomposition process in Figure 1.

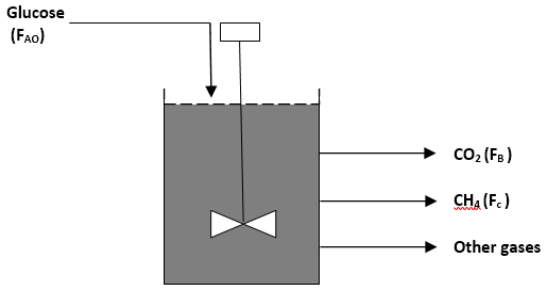

Figure 1: Glucose Decomposition Process in a CSTR

For the CSTR, the following assumptions can be applied while integrating the general conservation of mass balance equation for the development of the design models;

i. The reactor operates at steady state condition with continuous flow of reactant and product
ii. The reactant feed assumes uniform composition (constant density)
iii. Pressure drop along the reactor is negligible
iv. The reaction mixture is composed of 1 mole of glucose reactant species and 3 moles of products (carbon-dioxide and methane gas).
v. The feed assumes a uniform composition throughout the reactor meaning that the reactant mixture is well stirred
vi. Balance can be made about the entire volume of the reactor.
vii. Shaft work by the impeller or the stirrer is negligible.
viii. The process occurs isothermally (constant temperature)

The mass balance over the reactor volume is given in equation (5) .

$$\left[\begin{array}{l}\text{Rate of}\\ \text{Accumulation}\\ \text{of Product}\\ \text{within the}\\ \text{Elemental}\end{array}\right] = \left[\begin{array}{l}\text{Rate of Input}\\ \text{of Feed into}\\ \text{Elemental}\\ \text{Volume}\end{array}\right] - \left[\begin{array}{l}\text{Rate of}\\ \text{Outflow of}\\ \text{Feed from}\\ \text{Elemental}\\ \text{Volume}\end{array}\right] - \left[\begin{array}{l}\text{Rate of}\\ \text{Depletion of}\\ \text{Feed due to}\\ \text{Chemical}\\ \text{Reaction}\end{array}\right] \quad (5)$$

The terms in Equation (5) can be defined as follows;

$$\left[\begin{array}{l}\text{Rate of}\\ \text{Accumulation of}\\ \text{Product within the}\\ \text{Elemental Volume}\end{array}\right] = \frac{d}{dt}\left(C_A V_R\right) \quad (6)$$

$$\left[\begin{array}{l}\text{Rate of Input of Feed into Elemental Volume}\end{array}\right] = F_{AO} \tag{7}$$

$$\left[\begin{array}{l}\text{Rate of Output of Feed from Elemental Volume}\end{array}\right] = F_A = F_{AO}(1-X_A) \tag{8}$$

$$\left[\begin{array}{l}\text{Rate of Depletion of Feed due to Chemical Reaction}\end{array}\right] = (-r_A)V_R \tag{9}$$

Substituting Equation (6) to (9) into Equation (5) yields:

$$\frac{d}{dt}(C_A V_R) = F_{AO} - [F_{AO}(1-X_A)] - (-r_A)V_R \tag{10}$$

At steady state, the accumulation term is equal to zero, that is:

$$\frac{d}{dt}(C_A V_R) = 0$$

$$\therefore \quad 0 = F_{AO} - [F_{AO}(1-X_A)] - (-r_A)V_R \tag{11}$$

Expanding and simplifying Equation (11) yields;

$$V_R = \frac{F_{AO}X_A}{(-r_A)} \tag{12}$$

Substituting equation (4) into equation (12) yields;

$$V_R = \frac{F_{AO}X_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t)+1\right)\right]C_{Ao}(1-X_A)} \tag{13}$$

Where V_R is the volume of the CSTR in (m^3), F_A and F_{A0} are the final and initial molar flow rate of feed in (mol/s)

Equation (13) gives the volume of the CSTR for biogas production.

For a cylindrical reactor, the height, diameter, space time and space velocity of the digester are given in Equations (14), (15), (16) and (17) respectively as:

$$H_R = \left[\frac{16\ F_{AO}X_A}{\pi.A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda-t)+1\right)\right]C_{Ao}(1-X_A)}\right]^{1/3} \tag{14}$$

$$D_R = \frac{\left[\frac{16\ F_{AO}X_A}{\pi.A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda-t)+1\right)\right]C_{Ao}(1-X_A)}\right]^{1/3}}{2} \tag{15}$$

$$\tau_{CSTR} = \frac{X_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda-t)+1\right)\right]C_{Ao}(1-X_A)} \tag{16}$$

$$S_V = \frac{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda-t)+1\right)\right]C_{Ao}(1-X_A)}{X_A} \tag{17}$$

The quantity of heat generated per unit volume of the digester (q) is given as:

$$q = \frac{\Delta H_R\ F_{AO}X_A}{V_R} \tag{18}$$

Where q is the quantity of heat generated per unit volume of the CSTR digester in (kW/m^3) and ΔHR is the change in enthalpy of reaction in (j/mol)

2.2.4. Development of the PFR design model

Consider the hypothetical diagram of a PFR digester in Figure 2.

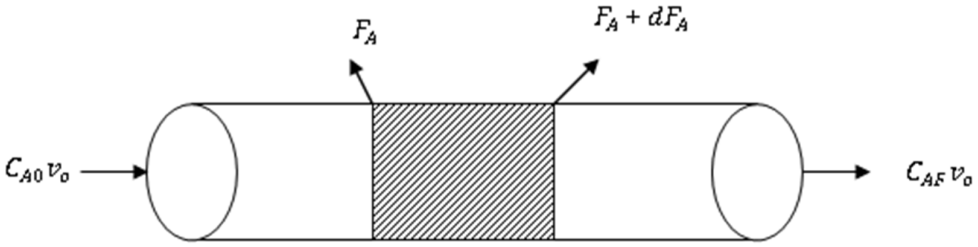

Figure 2: Plug flow reactor schematic

For the PFR material balance, the terms in Equation (5) can be defined as follows;

$$\begin{bmatrix}\text{Rate of Accumulation of Material with the Volume}\end{bmatrix} = 0 \quad (19)$$

$$\begin{bmatrix}\text{Rate of input of feed io the Volume}\end{bmatrix} = F_A \quad (20)$$

$$\begin{bmatrix}\text{Rate of output of feed from the Volume}\end{bmatrix} = F_A + dF_A \quad (21)$$

$$\begin{bmatrix}\text{Rate of depletion of feed due to chemical reaction}\end{bmatrix} = (-r_A)V_R \quad (22)$$

Combining equation (19) to (22) into (5) yields;

$$F_A - F_A + dF_A + (-r_A)dV = 0 \quad (23)$$

Equation (23) can be simplified to give the following;

$$dF_A + (-r_A)dV = 0$$

But $F_A = F_{AO}(1 - X_A)$

$$dF_A = F_{AO}dX_A$$

$$-F_{AO}dX_A + (-r_A)dV = 0$$

$$F_{AO}dX_A = (-r_A)dV \quad (24)$$

Equation (24) can be re-arranged and integrated to give;

$$V_R = F_{AO}\int_O^{X_{AF}} \frac{dX_A}{(-r_A)} \quad (25)$$

But $F_{AO} = C_{AO}\,V_O$ (26)

$$V_R = C_{AO}V_O\int_O^{X_{AF}} \frac{dX_A}{(-r_A)} \quad (27)$$

Substituting equation (5) into (27) yields;

$$V_R = C_{AO}V_O \int_O^{X_{AF}} \frac{dX_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t)+1\right)\right] C_{Ao}(1-X_A)} \tag{28}$$

Equation (28) gives the volume of plug flow digester.

The height, diameter, space time and space velocity can obtained from further simplification of the mathematical relationship of the cylindrical reactor thus;

$$H_R = \frac{F_{AO}}{\pi D^2} \int_O^{X_A} \frac{dX_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t)+1\right)\right] C_{Ao}(1-X_A)} \tag{29}$$

Equation (29) shows the mathematical relationship between the PFR height and diameter;

$$\tau = \frac{F_{AO} \int_O^{X_A} \frac{dX_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t)+1\right)\right] C_{Ao}(1-X_A)}}{V_O} \tag{30}$$

$$S_v = \left\{ \frac{F_{AO} \int_O^{X_A} \frac{dX_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t)+1\right)\right] C_{Ao}(1-X_A)}}{V_O} \right\}^{-1} \tag{31}$$

The potential heat involved or generated per unit volume of the PFR during the process of biogas production is given as;

$$q_h = \frac{Q}{V_R} = \frac{\Delta H_r F_{AO} X_A}{V_R} \tag{32}$$

Where Q is the quantity of heat generated in (kW).

Substituting equation (28) into (32) and further simplification yields;

$$q_h = \frac{\Delta H_r X_A}{\int_O^{X_{AF}} \frac{dX_A}{A\exp\left[-\exp\left(\frac{\mu e}{A}(\lambda - t)+1\right)\right] C_{Ao}(1-X_A)}} \tag{33}$$

The pressure drop along the plug flow digester can be computed for a laminar flow as;

$$\Delta P = 0.0013 f \rho v^2 \frac{F_{AO}}{\pi D^3} \int_{O}^{X_A} \frac{dX_A}{\mathrm{A} \exp\left[- \exp\left(\frac{\mu e}{A}(\lambda - t) + 1 \right) \right] C_{A0}(1 - X_A)} \quad (34)$$

2.2.5. Solution techniques

The design models of the digesters were solved numerically using Runge Kutta Agorithm in built in the MATLAB Code.

3. RESULTS AND DISCUSSION

The results of the flow digesters (CSTR and PFR) design performance evaluation for biogas production are presented in Table 1 and Figure 3 to 8.

Table 1: Design results of CSTR and PFR

Reactor design parameters (Unit)	@ 90% Fractional conversion		Difference (%)
	CSTR	PFR	
Volume (m^3)	10.500	2.800	28.947
Height (m)	3.800	2.900	6.716
Diameter	1.900	1.120	12.914
Space Time (s)	8.000	5.600	8.824
Space Velocity (s^{-1})	0.125	0.175	8.882
Quantity of Heat Generated per unit volume of the Reactor (kW/m^3)	0.500	1.750	27.778

Table 1 shows the design performance evaluation of the co-digesters (CSTR and PFR) during biogas production from the decomposition of glucose obtained from biodegradable waste materials. The MATLAB simulation of the flow digester models was done at same initial feed and operating condition. At maximum conversion of 0.9, the CSTR and PFR volume was 10.500 m^3 and 2.800 m^3 respectively with while the quantity of heat generated per unit volume of the CSTR and PFR was 0.50 kW/m^3 and 1.75 kW/m^3 respectively. The analysis of design results showed that more yield of biogas is produced in the CSTR as shown in the volume while the PFR performed better in terms of energy efficiency of the process as indicated in the quantity of heat generated per unit volume of the digester.

Figure 3 is a graphical relationship between the CSTR and PFR volume and fractional conversion obtained from the MATLAB simulation of both digesters at same initial feed and operating temperature of 300K and 310K with varying fractional conversion of $X_A \geq 0 \leq 0.90$ at an interval of 0.1. According to the profile, the volume of both digesters increases exponentially as the fractional conversion increases. However, at maximum conversion of 0.90, the volume of the CSTR and PFR was 10.50m^3 and 2.80m^3 respectively. This result showed that more yield of biogas is produced in the CSTR compared to that of the PFR during the glucose decomposition process, this result is greatly influenced by certain factors such as reaction kinetics, mass transfer limitations and reactor configuration. The profile is in agreement with steady state CSTR and PFR trend or behavior (Wosu *et al.*, 2024; Wosu 2024c)

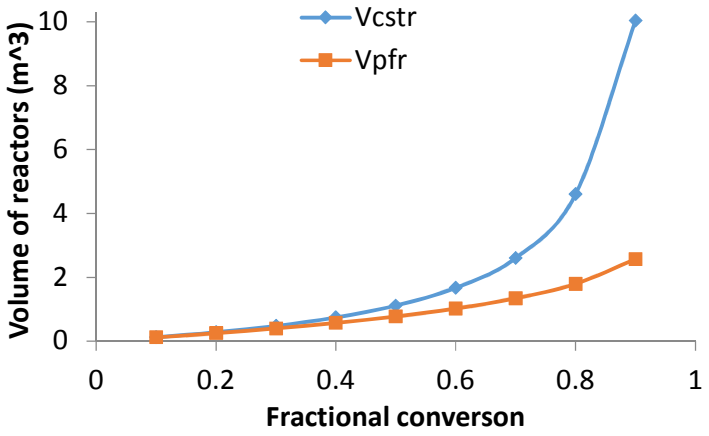

Figure 3: Profile of CSTR and PFR volume and fractional conversion

Figure 4 is a profile relationship of CSTR and PFR height and fractional conversion obtained from the MATLAB simulation of flow digesters performance models for height during the decomposition process. The simulation was performed at same initial feed and operating temperature of 300K and 310K with fractional conversion variation of $X_A \geq 0 \leq 0.90$ at an interval of 0.1. From the profile behavior, the digesters height was increased exponentially as the fractional conversion increases and at a maximum conversion of 0.90, the height of the CSTR and PFR design was 3.800m and 2.900m respectively. This result is mathematically justified since the digester volume is also a function of its height and the profile trend is in line with the results obtained by Wosu *et al.*, (2024) and Wosu, (2024c).

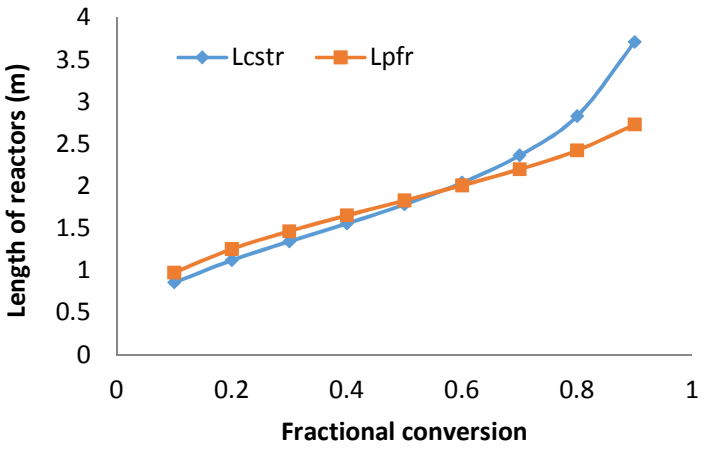

Figure 4: Profile of CSTR and PFR height and fractional conversion

Figure 5 shows a variation of the CSTR and PFR diameter with fractional conversion for the production of biogas from glucose decomposition. This profile was obtained from the MATLAB simulation of the steady state performance model of the flow digesters diameter at same initial feed and operating temperature of 300K and 310K with fractional conversion variation of $X_A \geq 0 \leq 0.90$ at an interval of 0.1. According to the plot, the diameter of both digesters was increased exponentially as the fractional conversion increases in both digesters. At a maximum fractional conversion of 0.90, the CSTR and PFR diameter was 1.90m and 1.12m respectively. This is justified by the high significant difference between the digesters volume. The developed profile is in agreement with steady state condition behavior of flow digesters (Wosu *et al.*, 2024 ; Wosu, 2024c).

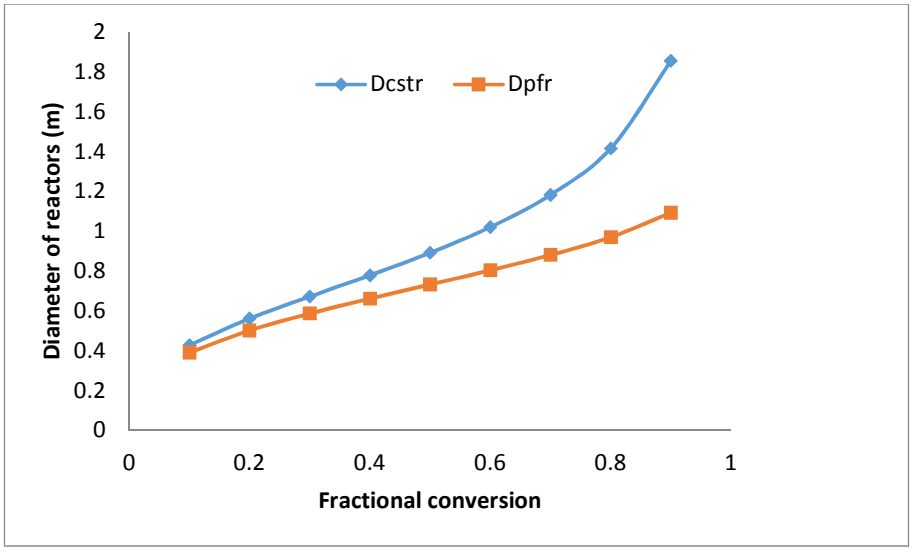

Figure 5: Profile of CSTR and PFR (D_R) and Fractional Conversion (X_A)

Figure 6 shows an exponential increase of the CSTR and PFR space time as the fractional conversion increases. This profile was obtained from the MATLAB simulation of the space time steady state model for the decomposition process. At a maximum fractional conversion of 0.90, the CSTR and the PFR space time values taken were 8.00 seconds and 5.60seconds respectively. The difference between the CSTR and PFR space time is an indication that more time is required in the CSTR due to its design configuration and nature reactants involved in the glucose decomposition process for biogas production. The profile behavior showed a similar trend for CSTR and PFR steady state operation process by Wosu *et al*., (2024) and Wosu, (2024c).

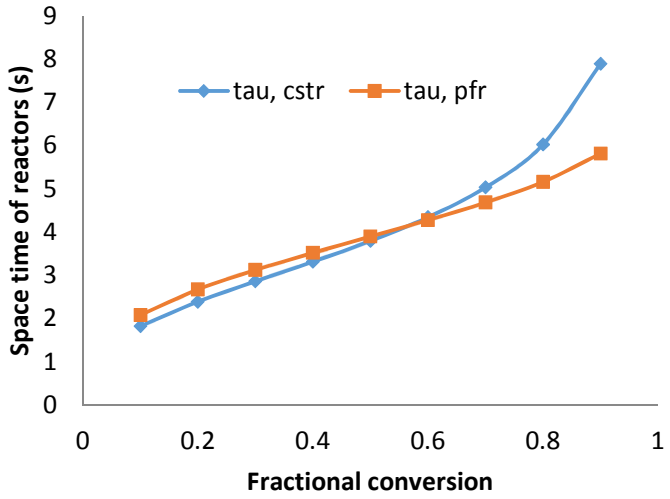

Figure 6: Profile of the CSTR and PFR Space Time (τ) and Fractional Conversion (X_A)

Figure 7 shows the CSTR and PFR space velocity variation with fractional conversion during the biogas production in both digesters. This profile was developed from the MATLAB simulation of the steady state performance models of the digesters space velocity at same initial feed and operating temperature of 300K and 310K with change in fractional conversion of $X_A \geq 0 \leq 0.90$ at 0.1 intervals. According

to the plot, the space velocity of the digesters decreases exponentially as the fractional conversion increases. At higher fractional conversion above 0.9, the space velocity value in both reactors tends towards negative infinity. This profile behavior is justified by the mathematical relationship between the space time and the space velocity. At a maximum conversion of 0.90, the CSTR and PFR space velocity was $0.125s^{-1}$ and $0.179s^{-1}$ respectively. Here, the space velocity of the PFR is higher because of its configuration for fast reaction within a short residence time compared to that of the CSTR.

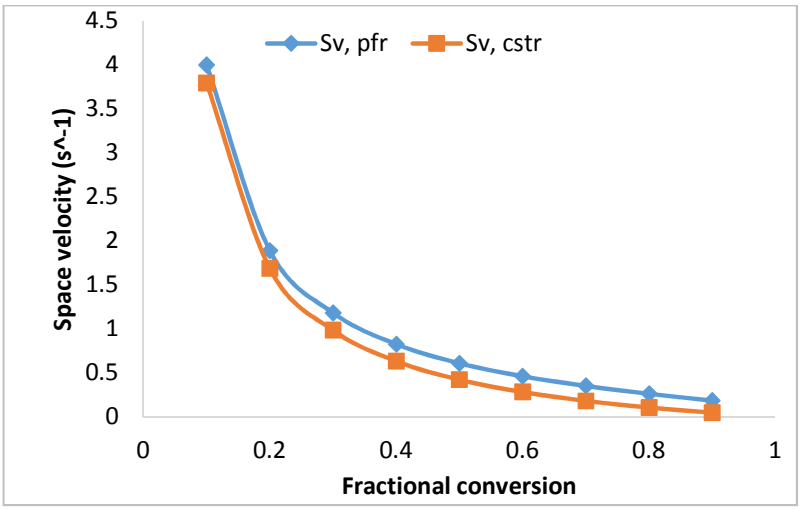

Figure 7: Profile of the CSTR and PFR Space Velocity (S_V) and Fractional Conversion (X_A)

Figure 8 shows the relationship between the CSTR and PFR quantity of heat generated per unit volume of the reactors and fractional conversion during the decomposition process. This profile was developed from the MATLAB simulation of the process using the same operating condition in both digesters. According to the profile, the quantity of heat generated per unit volume of both reactors decreases exponentially as the fractional conversion increases. At a maximum conversion of 0.90, the quantity of heat generated per unit volume of the CSTR and PFR was $0.50kW/m^3$ and $1.75kW/m^3$ respectively. The results showed that the PFR design is better in terms of energy efficiency of the process than the CSTR.

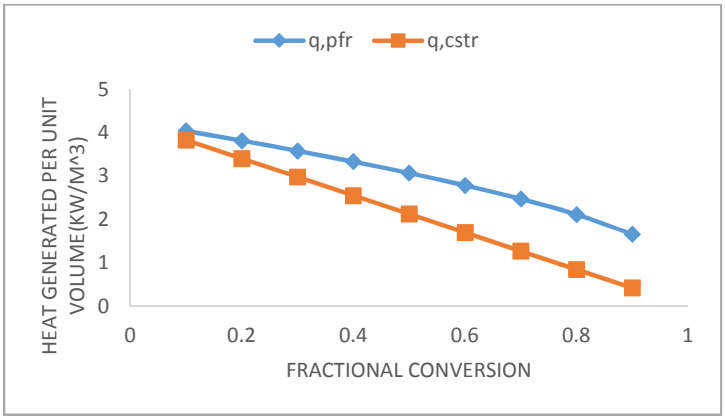

Figure 8: Profile of the quantity of heat generated per unit volume of the CSTR and PFR (q) with fractional conversion

4. CONCLUSION

Design models of the flow digesters for biogas production from the decomposition of glucose were developed using the principle of conservation of mass and energy. The developed models were

simulated using MATLAB R2023a version at same initial feed and operating conditions of the co-digesters. At a conversion of 90% the co-digester (CSTR and PFR) specification for volume that will process 166.752kg/day of the glucose was 10.500m^3 and 2.800m^3 respectively while the quantity of heat generated per unit volume of the CSTR and PFR digesters was 0.50kW/m^3 and 1.75kW/m^3 respectively. The performance evaluation of the co-digester design showed that the CSTR displayed a better performance in terms of biogas yield as indicated by the high volume of the digester while the PFR performed better in terms of energy efficiency and conservation as shown in the quantity of heat generated. The design of co-digester for biogas production is a waste management strategy and the technology contribute substantially to the global demand of renewable and sustainable energy source which is cheap, available and environmentally friendly compared to other non-renewable like oil, gas, coal and electricity.

5. CONFLICT OF INTEREST

There is no conflict of interest associated with this work.

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