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Anaerobic Treatment for Palm Oil Mill Effluent Using Covered In-the Ground Anaerobic Reactor (CIGAR)

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Highlights:

- Anaerobic treatment for POME using CIGAR was investigated and Monod, Contois, Moser, and Shuler kinetic models were evaluated.
- The Shuler kinetic model performed best with a specific growth rate (μ_{max}) of 0.052 d^{-1} and a saturation constant (K_{so}) of 0.119.
- The first order reaction model performed best for the substrate utilization kinetic, with a k_s of 2.183 d^{-1} and a $Y_{x/s}$ of 0.024 kg/kg.
- The maximum average efficiency of anaerobic degradation (34.4%) occurred at a feeding rate 100 L/d with CH_4 yield 0.120 Nm^3/kg of removed COD.

Abstract. Wastewater from crude palm oil mills contains high organic matter, which potentially produces biogas through anaerobic digestion processes. The design and operation of an anaerobic bioreactor require a good understanding of the reaction kinetic in the bioreactor. This study aimed to evaluate the biogas production from POME and to determine the kinetic parameters of microbial growth and the substrate utilization rates in a CIGAR. An experiment was conducted using a 5-m^3 bioreactor with a working volume of 4.4 m^3. Wastewater from the Bekri palm oil mill was stored in a 5-m^3 tank. After stabilization, the wastewater was loaded into the reactor at a rate of 100 to 250 L/d, corresponding to a COD loading rate of 1.373-3.097 $kg \cdot m^{-3}.d^{-1}$, and an HRT of 18-44 days. Monod, Contois, Moser, and Shuler kinetic models were evaluated. The results showed that the Shuler model performed best for microbial activities, while the first order reaction model performed best for the substrate utilization kinetic. The maximum specific growth rate (μ_{max}) for the Shuler model was 0.052 d^{-1} and the saturation constant (K_{so}) was 0.119. The maximum substrate utilization rate constant (k_s) was 2.183 d^{-1} and biomass yield ($Y_{x/s}$) 0.024 kg/kg. The maximum average efficiency of anaerobic degradation (34.4%) occurred at a feeding rate of 100 L/d with methane yield of 0.120 Nm^3/kg of removed COD. This value is relatively low compared to the maximum potential of 0.350 Nm^3/kg COD_r.

Keywords: *anaerobic; biogas; Indonesia; kinetic; palm oil; wastewater.*

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1 Introduction

During oil palm extraction, a lot of wastes are released, namely mesocarp fiber (12.3%), shell (5.4%), empty fruit bunch (21.5%), and palm oil mill effluent, or POME (0.84 m^3) [1]. Palm oil mill effluent is a viscous, brown, bad smelling liquid with high organic matter content generated mainly from sterilization, clarification and purification of CPO and hydro-cyclone processes with a share of 36, 60, and 4%, respectively [2]. Most CPO industries treat POME using conventional open ponds with a hydraulic retention time (HRT) of 50 to 70 days [3]. Such treatment has a long residence time [4], needs a large area, and the collection and utilization of methane (CH_4) gas is difficult [5]. A more environmentally friendly way of treating POME is using anaerobic treatment in a closed tank or pond to produce biogas. One type of digester that is now widely used in POME management is the Covered In-Ground Anaerobic Reactor (CIGAR) digester. This reactor is the same as a covered lagoon digester, where wastewater is accommodated in a covered pond. This reactor is widely applied for tapioca wastewater treatment in Lampung [6,7] and Thailand [8] as well as for POME in Indonesia. The CIGAR reactor has the advantages of a high methane yield, moderate capital investment, and low construction and operational costs [9]. The technology of a CIGAR digester is simple enough to make it is easy for local developers to master.

Conversion of POME into biogas by means of anaerobic digestion reduces waste pollution parameters such as organic solids, microbial pathogens, and toxicity [10]. This process also produces biogas that can be used as an alternative energy source, resulting in significant reduction of greenhouse gas emissions [11] and supporting the sustainability of oil palm plantation [12]. Additionally, slurry digestates released from biogas digesters are a good source of organic fertilizer [13,14].

Operation of anaerobic wastewater treatment often fails because of too high organic loading. On the other hand, very low substrate feeding results in very long HRT and a large bioreactor is needed to convert the organic compounds into biogas [15], even though the efficiency of organic compound degradation is high. Increasing the organic loading rate will increase the biogas production to a certain rate, at which methanogenic bacteria can no longer convert acetic acid into CH_4 [4]. However, this may cause washout of microbes from the bioreactor exceeding their growth rate. As a result, contact intensity between the substrate and microbes decreases, interrupting the anaerobic degradation stability [15].

The performance of the wastewater treatment process in an anaerobic reactor is affected by the dynamics of microbial growth and substrate utilization. A suitable kinetics model is required to provide optimal conditions for the organic matter to

decompose due to microorganism activity [16]. A kinetics study is also useful for estimating the suitability of feedstocks as a substrate in the biogas process [17]. The objective of this research was to present anaerobic treatment of POME to generate biogas and to determine the kinetic parameters of substrate utilization and microbial growth in a mini scale CIGAR digester with semi continuous feeding using selected kinetic models.

2 Materials and Methods

2.1 Digester and Substrate Preparation

An experiment was conducted using a mini scale CIGAR bioreactor at the Wastewater Treatment Lab, Department of Agro-industrial Technology, University of Lampung (Figure 1). The bioreactor was locally fabricated from resin fiber and had a capacity of 5 m^3 and a working volume of 4.4 m^3. In order to facilitate the biological activated sludge to settle [18], the reactor was partitioned using baffles into three equally sized chambers, as pictured in Figure 1. The last chamber is responsible for settling the biological activated sludge to be used for seeding. The reactor was buried in soil at a depth of 1.5 m with 20 cm rising above the ground for easy maintenance. The bioreactor base was slightly tilted downward (around 1%) in order to facilitate sludge sedimentation. The bioreactor was operated with semi continuous feeding. The pump was connected to the stock tank and was turned on as required to circulate the wastewater with the purpose of homogenizing the wastewater before feeding into the digester.

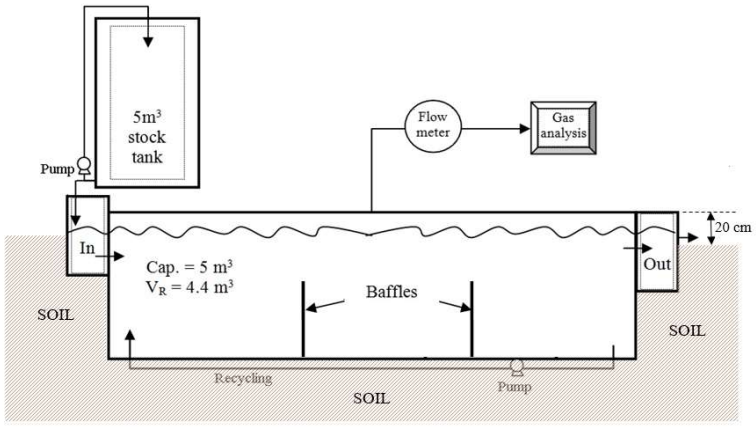

Figure 1 Schematic of pilot scale CIGAR bioreactor used in the experiment.

Palm oil mill effluent was taken from a cooling pond of the wastewater treatment plant of PTPN VII Bekri palm oil mill (Central Lampung) and was trucked to the laboratory and stored in a 5-m^3 plastic tank for substrate supply. After two weeks of use, the tank was emptied and refilled with new POME. Based on pH measurements (4.65-4.98 with a standard deviation of 0.094), it was assured that the change in the POME's state was insignificant during two weeks of storage. Table 1 shows the characteristic of the POME used in this experiment along with values from literature as comparison.

Table 1 Characteristic of POME used in experiment and literature comparison.

Characteristic	This work	Teng [19]	Sarono [3]
pH	4.65-4.98	4.15-4.45	5.63-5.64
COD (g/L)	37.4-60.4	44,5-65	41.25-52
VS (g/L)	23.23-23.90	27.3-30.15	-
TS (g/L)	25.17-26.46	33.79-37.23	-

2.2 Experiment

Previously we have reported the effect of a feeding rate in the range of 50 and 350 L/d on the biogas yield [20]. The bioreactor showed increasing biogas yield at a feeding rate from 50 to 250 L/d. At a feeding rate of 350 L/d the biogas yield started to decrease drastically due to washout of active microorganisms and decreasing pH. In this work, therefore, we carried out the experiment by varying the feeding rate from 100 to 250 L/d. Initially, the digester was acclimatized at a feeding rate of 50 L/d for 3 weeks. During the acclimatization stage, biogas production was measured to monitor the stability of the digestion process. After that, the feeding rate was gradually increased to 100, 150, 200, and 250 L/d. Each feeding rate was kept for about one month before changing to a different rate. The POME was stirred for around one hour prior to its loading into the digester. This step was conducted to homogenize the POME. Daily observation of pH of the substrate and the biogas yield was done. The chemical oxygen demand (COD), total suspended solid (TSS), and volatile suspended solid (VSS) were analyzed biweekly.

2.3 Analysis

The acidity of the POME was measured using a pH meter (DKK-TOA Corporation, Japan). Total suspended solid (TSS) was analyzed using 50 ml of fresh and spent POME. The sample was centrifuged at 3000 rpm for 15 min. The solid was removed using a pipette and then dried in an oven at 105 °C for two hours. The sample was weighted to get the TS content after cooling in a desiccator. The volatile suspended solid (VSS) content was analyzed by burning the dried sample in a furnace (Barnstead Thermolyne 1300) at 550 °C for 2 hours

[21]. The chemical oxygen demand (COD) was analyzed based on standard methods from APHA (American Public Health Association) for the examination of water and wastewater [19]. The analysis used the closed reflux method (Hach DRB 200) followed by spectrophotometry (Hach DR/4000 U). The COD removal (COD_r) was calculated as the difference between COD input and COD output.

The volume of biogas production was monitored daily using a gas flow meter (Itron ACD G1.6) with the capacity to measure a minimum flowrate of 0.016 m^3/h up to maximum of 3 m^3/h. The flow meter works by connecting the biogas line to the inflow port of the flow meter. The biogas yield was calculated by subtracting the previous recorded value from the last. The biogas composition was measured using a gas chromatograph (GC-2014 Shimadzu) using helium as carrier gas at a flow rate of 40 mL/min. The GC employs a thermal conductivity detector and a column (Shincarbon, 3 mm inner diameter and 4.0 m length) with temperature set at 200 °C, injection time at 1 minute, temperature at 100 °C and injection pressure at 100 kPa.

2.4 Growth Kinetic Models

In this work, Monod, Contois, Moser, and Shuler kinetic models were evaluated because they are simple, i.e. the input and output parameters can be represented by organic concentration. In this way, interpretation and field implementation can easily be done.

An empirical relationship to describe the specific microbial growth rate (μ, unit) as a function of the growth-limiting substrate concentration (S, g/L) has been proposed as in Eq. (1) [22]:

$$\mu = \frac{\mu_{max} \cdot S}{K_s + S} \tag{1}$$

where μ_{max} is the maximum specific microbial growth rate (d^{-1}) and K_s is the half-saturation constant (g/L). The mass balance in a continuous flow reactor system is presented as:

$$QX_o - QX + V_R\mu X = V_R[dX/dt] \tag{2}$$

where Q is the substrate loading rate (L/d), X_o and X are the influent and effluent cell concentration (g/L), respectively, and V_R is the bioreactor working volume (L, assumed to be constant). Defining the dilution rate $D = Q/V_R = 1/HRT$, where HRT is the hydraulic retention time. Assuming that the influent biomass concentration can be neglected and that the steady state condition prevails ($[dX/dt] = 0$) it can be shown that Eq. (2) yields $\mu = 1/HRT$. Substituting μ into

the growth kinetic equations (Eq. (1)), we can represent the Monod kinetic model as follows:

$$HRT = \frac{1}{\mu_{max}} + \frac{K_s}{\mu_{max}} \frac{1}{S} \tag{3}$$

The other suggested kinetic models, Contois [23], Moser [24], and Shuler [25], are presented in Eqs. (4) to (6) respectively:

Contois:

$$HRT = \frac{1}{\mu_{max}} + \frac{K_{ss}}{\mu_{max}} \frac{X}{S} \tag{4}$$

Moser:

$$HRT = \frac{1}{\mu_{max}} + \frac{K}{\mu_{max}} \frac{1}{S^n} \tag{5}$$

Shuler:

$$HRT = \frac{K_{so} S_o}{\mu_{max} S} + \frac{1}{\mu_{max}} \tag{6}$$

where S_o and S are the substrate concentration (g/L) in the influent and the effluent respectively, X is the biomass concentration in the effluent (g/L), K_{ss}, K $(g/L)^n$, K_{so} are the constants for the Contois, Moser, and Shuler models respectively. The n in the Moser model is an adjustable parameter. Volatile solid (VS) is the parameter most commonly used to follow biomass growth (X) in full scale biological wastewater treatment systems [26]. The COD values of the substrate in the influent and effluent are represented by S_o and S, respectively. Kinetic growth parameters μ_{max} and K can be obtained by plotting $1/S$, X/S, $1/S^n$, and S_o/S vs HRT for the Monod, Contois, Moser, and Shuler models, respectively.

2.5 Substrate Utilization Kinetic Models

The substrate utilization rate (r_{su}) is expressed using the Michaelis-Menten model [26]:

$$r_{su} = -\frac{dS}{dt} = \frac{-k \cdot S \cdot X}{K_s + S} = \frac{S_o - S}{HRT} \tag{7}$$

where k is the maximum specific utilization rate constant. Negative sign is used to show that substrate mass decreases with time. Rearranging Eq. (7) we have:

$$\frac{HRT \cdot X}{S_o - S} = \frac{1}{k} + \frac{K_s}{k}\frac{1}{S} \tag{8}$$

Other models evaluated in this work are the first order, modified Stover-Kincannon, and Grau second order models. Using the first order kinetic model, $r_{su} = dS/dt = -k_s \times S$. Applying this to a process in a pond (modeled as a completely mixed digester without recycling) and after integration we have:

$$(S_o/S) - 1 = k_s \times HRT \tag{9}$$

where k_s is the first order rate coefficient obtained from plotting (S_o/S) vs *HRT*.

The modified Stover-Kincannon [27] and Grau second order kinetic models [28] are presented in Eqs. (10) and (11) respectively:

$$\frac{HRT}{S_o - S} = \frac{1}{U_{max}} + \frac{K_B}{U_{max}}\frac{1}{S_o} \tag{10}$$

$$\frac{S_o \times HRT}{S_o - S} = a + b \cdot HRT \tag{11}$$

where U_{max} is the maximum specific substrate utilization rate (g/L.d) and K_B is the half saturation constant (g/L), $a = S_o/k_s.X$ and b are constants obtained from plotting $S_o \cdot HRT/(S_o - S)$ vs *HRT*. The substrate utilization kinetic constants are obtained similarly from Cartesian plots as above. Figure 2 presents a block diagram of the steps performed in this experiment.

3 Results and Discussion

3.1 Digester Performance

Table 2 provides the steady state condition of the CIGAR bioreactor characterized by loading rate, COD_{in}, COD_{out}, and VSS. The table also lists biogas and methane yield, COD removal (COD_r), and process efficiency. In the Table 2, process efficiency is calculated from:

$$\text{Efficiency} = [CH_4 \text{ yield} / (350 \times 303/273)] \times 100\% \tag{12}$$

where 350 ($LCH_4/kgCODr$), equivalent to 0.25 ($kgCH_4/kgCOD_r$), is the stoichiometric conversion of COD_r into methane at standard condition, and 303 is the average working temperature (in K, equal to 30 °C) of the bioreactor. This conversion is recommended by the Intergovernmental Panel on Climate Change (IPCC) for calculating methane emitted from wastewater [29].

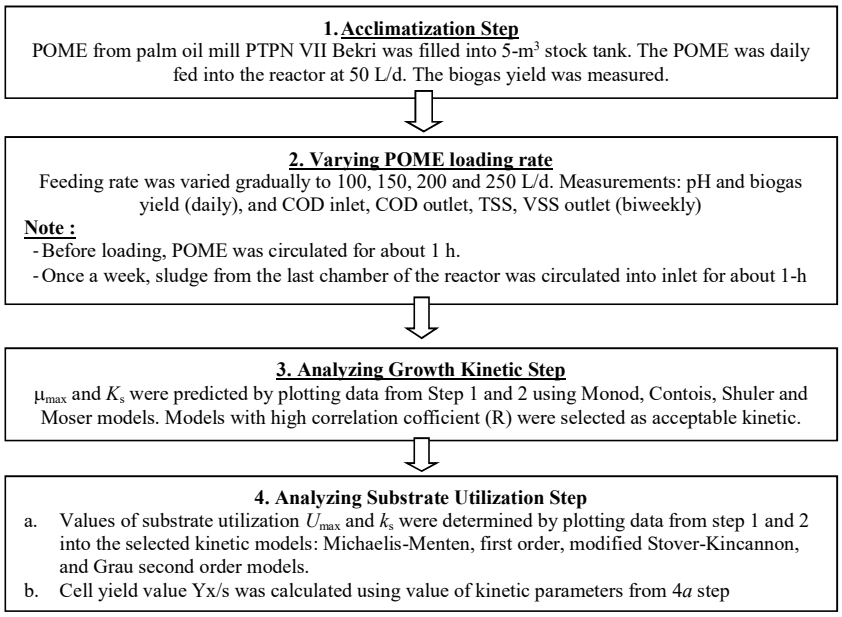

Figure 2 Block diagram of the steps performed during the experiment.

Table 2 Steady state condition of CIGAR bioreactor.

Loading Rate, F (L/d)	HRT (d)	COD_{in} (S_o) g/L	COD_{out} (S) g/L	Daily Biogas Yield (NL/kgCOD_r)	Daily CH_4 Yield (NL/kgCOD_r)	VSS, X (g/L)	Process Efficiency (%)
100	44	59.91±1.29	4.20±0.75	198.4±30.3	120.5±18.4	0.91±0.27	34.4±5.2
150	29	39.60±20.80	6.71±0.85	194.7±114.0	105.8±60.0	1.71±0.09	30.2±17.4
200	22	49.31±5.61	2.14±0.67	166.9±47.6	91.6±28.0	0.45±0.07	26.2±8.0
250	18	54.50±0.06	2.10±0.13	138.3±49.1	86.8±30.8	0.53±0.23	24.8±8.8

It was shown that the CIGAR digester worked well, as can be seen from the large difference between COD_{in} and COD_{out}. Our calculation revealed that COD removal was between 78% at a feeding rate of 150 L/d and 96% at a feeding rate of 250 L/d. This is comparable to the performance of the Upflow Bioreactor with Central Substrate Dispenser (UBCSD) with an enhanced mixing process in [30], which had a COD removal of up to 95.2%. However, the high standard deviation of some of the data in Table 2 indicates a fluctuating variation in the data. This indicates the instability of the anaerobic decomposition process in the CIGAR digester system that was used, which could be caused by fluctuation of the COD input, sedimentation and washout problems.

3.2 Biogas and Methane Yield

It can be seen from Table 2 that the daily biogas yield ranged from 138.3±69.5 to 198.4±30.3 NL/kgCOD_r with feeding rate varied from 100 to 250 L/d. Our measurement of the biogas composition revealed a methane content of 48.4 to 62.77% (v/v). As given in the table, the methane yield was in the range of 86.8±43.6 to 120.5±30.3 NLCH_4/kgCOD_r a day, with the same loading rate variation. The table shows that the biogas yield decreased with feeding rate (*F*) from the highest yield of 198.4 at 100 L/d feeding rate to 138.3 L/kgCOD_r for an *F* value of 250 L/d. Figure 3 depicts a graphical representation of the effect of the feeding rate on the biogas and methane yield as well as the process efficiency. An increase in *F* resulted in a decrease in biogas yield, which could be caused by the occurrence of suspected washout removing some active bacteria from the reactor through the effluent so that the process efficiency decreased.

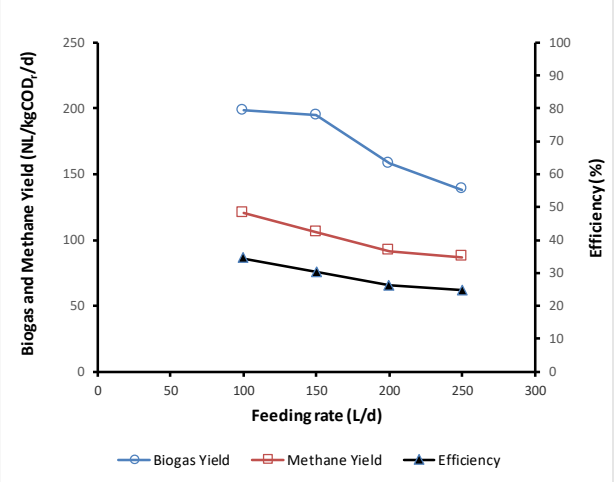

Figure 3 Average biogas and methane yield as a function of loading rate.

In terms of methane yield and process efficiency, the digester system is unsatisfactory. From Table 2, the process efficiency of the anaerobic reaction was considerably low, ranging from 24.8 to 34.4%. The efficiency decreased almost linearly with the feeding rate (*F*) from a maximum value of 34.4% for an *F* value of 100 L/d to 24.8 for a *F* value of 250 L/d. This may be due to the occurrence of precipitation or sedimentation. Other factors that cause low efficiency are washout and overload of COD at higher feeding rates. If this happens, the pH will drop. Our observations show that the outlet pH was relatively constant in a narrow range of 7.0 to 7.9 with a standard deviation of 0.23. This implies that the precipitation factor was more responsible for the low efficiency of the anaerobic

process. This means that the circulation of sludge from the bottom of the reactor once a week is not sufficient to produce efficient conversion of COD into biogas.

3.3 Growth Kinetic Models

Based on Table 1, we can calculate the respective variables to produce kinetic parameter plots based on the equations previously described. Kinetic plots of the four models are given in Figure 4. It should be noted that at a loading rate of 250 L/d, wash-out is suspected to occur, which removes active microorganisms from the reactor. Therefore, the data at this loading rate was excluded from the kinetic analysis. The data, however, was included in the evaluation of the reactor performance.

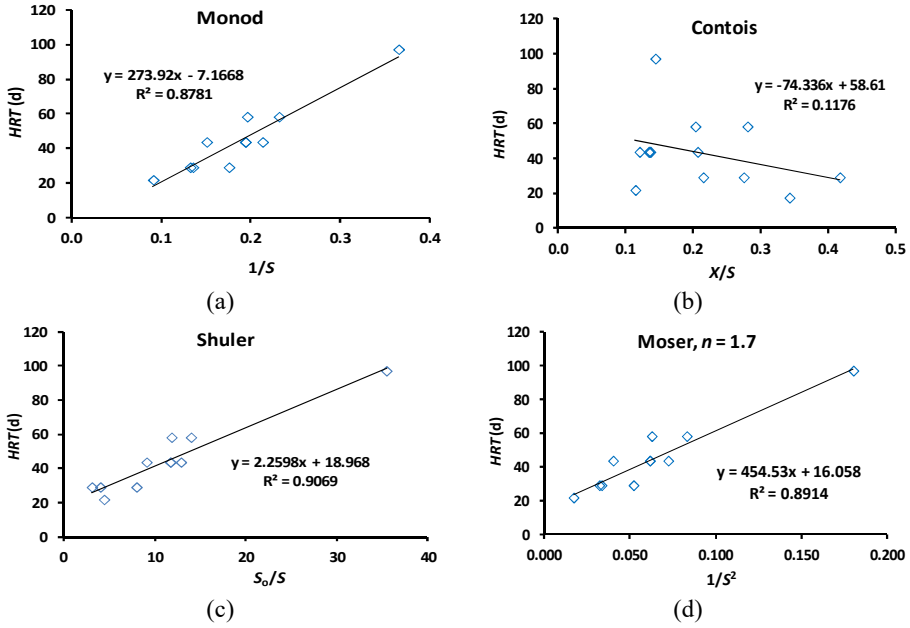

Figure 4 Plots of kinetic models: (a) Monod, (b) Contois, (c) Shuler, and (d) Moser (n = 1.67).

3.3.1 Monod Model

The plot for the Monod model is given in Figure 4(a) along with its regression equation. In this case, S is represented by COD_{out}, i.e. the COD value of the effluent. In the resulted equation, y stands for hydraulic retention time (HRT) and x stands for inverse of S or $1/S$. The intercept of the regression equation (-7.167) is $1/\mu_{max}$, while the coefficient of the equation (273.92) is K_s/μ_{max}. From the figure

it can be seen that the plot for the Monod kinetic model shows good correlation of *HRT* and 1/*S*, with a determination coefficient (R^2) of 0.88. However, this model provided a negative value for $1/\mu_{max}$, meaning a negative value for μ_{max}. Therefore, this model is unacceptable. According to Mahanta *et al.* [31], the Monod model is accurate for pure, homogenous cultures, but not for heterogeneous cultures or complex substrates.

3.3.2 Contois Model

The plot for the Contois model is given in Figure 4(b). In this case, *X* is represented by *VSS* as presented in Table 2. The Contois model is similar to the Monod model except that the term of 1/*S* in the Monod model is replaced by *X*/*S*. From the figure it is obvious that the Contois kinetic model failed to describe the relationship between *HRT* and *X/S*. This means that introduction of biomass concentration (*X*) into the model is not able to improve the Monod model. According to Tchobanoglous, *et al.* [26], VSS not only represents active cell concentration but also includes other particulate organic matter. Most wastewater also contains non-biodegradable VSS and perhaps affects VSS, which is slowly decomposed in the reactor. These solids are involved with biomass in the VSS measurement. Therefore, its use to represent cell concentrations of the active biomass requires correction for other components.

3.3.3 Shuler Model

Figure 4(c) shows the plot for the Shuler kinetic model. In this case, S_o is the COD value of the influent. The Shuler model is also similar to the Monod model; the difference is made by modifying 1/*S* with S_o/S. The figure clearly shows that the Shuler model produced the best fit to describe the relationship between *HRT* and S_o/S, with an R^2 value of 0.907. The maximum growth rate, μ_{max}, using this model was equal to 1/18.986 = 0.0527 d^{-1} and the saturation constant (K_{so}) was equal to 2.2598/18.968 = 0.119. Riffat [32] tabulated μ_{max} values in the range of 0.11 to 0.44 d^{-1}, meaning that our result was considerably lower than the values in the literature. A low μ_{max} implies that the bioreactor requires a much longer start-up time [33] to achieve stable process conditions.

3.3.4 Moser Model

Figure 4(d) shows that the Moser model with n = 1.67 is an acceptable model with an R^2 value of 0.891. The maximum growth rate, μ_{max}, using this model was equal to 1/16.058 = 0.062 d^{-1} and the saturation constant (*K*) was equal to 454.53×16.058 = 7298.84 $(L/g)^{1.7}$. The Moser model provides a degree of flexibility in fitting the data due to an adjustable *n* so that is able to predict dynamic behavior in the reactor [31].

3.4 Substrate Utilization Kinetic

Figure 5 shows that the Michalis-Menten model failed to represent the anaerobic reaction kinetic for POME degradation in the CIGAR bioreactor. Again, this could be due to the existence of a VSS that is inadequate to represent X. The Stover-Kincannon and Grau second order models showed very good fits of the respective kinetic parameters. However, both models provided a negative intercept (negative maximum specific substrate utilization rate U_{max} and negative of $S_o/k_s.X$). Therefore, both models were rejected. The first order substrate utilization kinetic model provided an acceptable result with a determination coefficient value of $R^2 = 0.908$. The value of k_s determines the slope of the plot (Figure 3(b)), which is 2.183 d^{-1}.

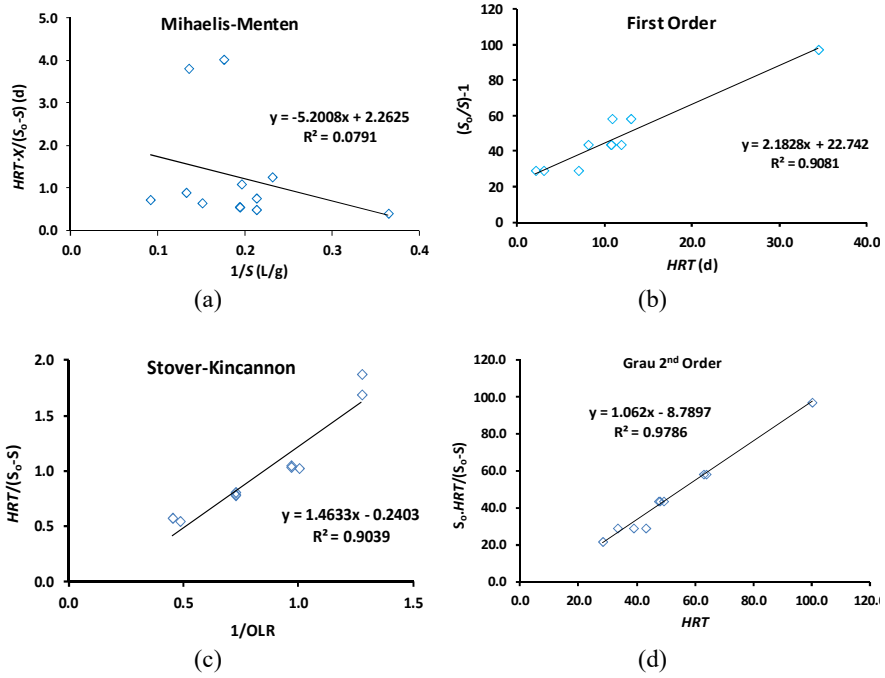

Figure 5 Substrate utilization kinetic plots for the (a) Michaelis-Menten, (b) first order, (c) Stover-Kincannon, and (d) Grau second order models.

Based on the kinetic parameters obtained using Shuler model, we calculated the value of the biomass cell yield from $Y_{x/s} = \mu_{max}/k_s = 0.053/2.183 = 0.024$. A list of yield coefficient values in the range of 0.023 to 0.054 kg biomass over kg COD was reported in [32]. It can be surmised that the Shuler model results in a biomass cell yield that is very close to the values from the literature.

As a comparison, we collected biokinetic constants from several other studies, which are summarized in Table 3. The microbial growth rate constant (μ_{max}) in our study was 0.053 d-1 lower than the results of other studies that used the same POME substrate. Our value is a quarter of that reported by Setiadi (1996) for POME using an anaerobic baffled reactor (ABR) under stable conditions [34]. The substrate utilization kinetic follows the first-order model with maximum substrate utilization rate $k_s = 2.183$ d^{-1} and is significantly lower than that reported by Chan [35].

Table 3 Kinetic parameters of growth rate and substrate utilization for POME.

Reactor type	Kinetic Model	Kinetic Parameters	R^2	Reference
Growth Kinetic				
Stirred reactor, lab scale.	First order	$\mu_{max} = 0.477$ $K_s = 0.098$	0.974	[36]
ABR, unstable condition	Monod	$\mu_{max} = 2.42$ $K_s = 11.56$	--	[34]
ABR, stable condition	Monod	$\mu_{max} = 0.20$ $K_s = 0.34$	--	[34]
IAAB	Monod	$\mu_{max} = 0.103$ $K_s = 8.17$	0.741	[35]
UASBR	Monod	$\mu_{max} = 0.988$	0.998	[37]
Modified ABR	Monod	$\mu_{max} = 0.304$ $K_s = 0.313$	--	[38]
CIGAR, pilot scale	Shuler	$\mu_{max} = 0.053$ $K_s = 0.119$	0.959	This work
Substrate Utilization Kinetic				
CIGAR, pilot scale	First order	$k_s = 2.183$	0.908	This work
IAAB	Modified Stover-Kincannon	$U_{max} = 23.1$ $k_s = 14.7$	0.973	[35]
Stirred reactor, lab scale.	First order	$U_{max} = 0.868$	0.923	[36]

Note: UAPB (Up-Flow Anaerobic Packed Bed); CIGAR (Covered In-Ground Anaerobic Reactor); IAAB (Integrated Anaerobic-Aerobic Bioreactor); ABR (Anaerobic Baffled Reactor); UASBR (Upflow Anaerobic Sludge Blanket Reactor)

4 Conclusion

Anaerobic treatment of palm oil mill effluent using a pilot scale CIGAR bioreactor was conducted. The efficiency of the anaerobic process using the CIGAR bioreactor was low, with a maximum of 34.4% at a feeding rate of 100 L/d (corresponding to a COD loading rate of 1.37 $kg/m^3.d$) with a methane yield of 0.2 m^3CH_4/kg of removed COD. The cell growth kinetic in the CIGAR bioreactor was best represented by the Shuler model with a μ_{max} of 0.053 d^{-1} and a half-saturation constant K_{so} of 0.119. Another acceptable model was Moser's

with n = 1.67, resulting in μ_{max} of 0.062 d^{-1} and a saturation constant K of 7298.84 $(L/g)^{1.7}$. The substrate utilization kinetic followed the first order model with maximum substrate utilization rate k_s = 2.183 d^{-1}. Biomass yield coefficient $Y_{x/s}$ was obtained to be 0.024 kg/kg.

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