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A Parametric Study of Performance Characteristics of Upflow Anaerobic Sludge Blanket Bioreactors

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Abstract

Attempts have been made to perform a parametric study of the performance characteristics of UASB (upflow anaerobic sludge blanket) bioreactors. Bioreactor performance is first simulated mathematically using PFDRs – in – series model that assumes the sludge bed and the sludge blanket each equivalent to a PFDR (plug flow dispersion reactor) but with different values of axial dispersion coefficient (D_L). Experimental values of D_L are used in the computations. Alternate models that employ PFR – CSTR approach and the one proposed by Narayanan and Narayan are also considered for comparison. Dependence of total required height of reaction zone (L) with different system / operating parameters such as fractional conversion of substrate (α) desired, feed concentration (C_{S0}) and feed flow rate have been studied separately. Computed values of design parameters are compared with experimental data collected from industrial units / pilot plants. It is observed that the sludge bed is not a tightly packed bed, but is a partially expanded bed. Axial dispersion and resistance to substrate transfer to sludge granules are significant and cannot be neglected. Among the different kinetic pathways analysed, data based on Andrews' kinetic model that accommodates substrate inhibition to microbial growth agree more closely with experimental data.

Keywords: UASB, PFDR, PFR-CSTR, Bioreactors.

Introduction

Upflow anaerobic sludge blanket (UASB) bioreactors have found large scale commercial acceptance particularly in the areas of anaerobic digestion of liquid wastes. Though they demand significantly large startup time (3 – 8 months) and are restricted to anaerobic operation with complex microbial cultures, they do exhibit attractive features such as simple construction, capability of handling high strength feedstocks and providing more than 98% BOD removal at substantially large capacities. However, mathematical analysis of performance of these bioreactors is rather complex. The sludge bed is composed of sludge granules and the sludge blanket that gets formed above this bed is composed of gas bubbles (bubbles of biogas produced during the anaerobic decomposition of biomass) that carry sludge granules with them in the form of a wake or tail and the upflowing substrate solution. The blanket thus resembles a three phase fluidized bed. The most established theory on the formation of sludge granules[1] is that it is an evolutionary, biological process arising from the natural tendency of different categories of microbes to come together and arrange themselves to form colonies so as to optimize their activities and interactions. Each granule is, thus, such a colony of microbes. Being a revolutionary process, it is slow and that explains the large startup time demanded by these bioreactors.

Mathematical analysis of performance of UASB reactors has been attempted by Kalyuzhnyi and Fedorovich[2] and by Narayanan and Narayan[3]. Kalyuzhnyi and Fedorovich have proposed a dispersed flow model for the bioreactor. They have, however, computed the dispersion coefficient from an empirical correlation (though considered its variation along the height of the reactor due to change in gas velocity) and have neglected resistance to substrate transfer into sludge granules. They also have confined to Monod – type kinetics. The simulation model reported by Narayanan and Narayan[3] is more rigorous. Apart from axial dispersion and its variation in the axial direction, they have considered different kinetic pathways and resistance to transport of substrate into sludge granules has been adequately accounted for by incorporating appropriately defined effectiveness factors. Their package is thus more versatile though it demands large computational load.

In the present paper, attempts have been made to analyse the performance characteristics of UASB bioreactors and study the reactor's response to variations in different process / operating parameters (feed concentration, feed flow rate, fractional conversion desired, kinetic pathways etc.) based on a PFDRs – in – series approach and also on the simplified PFR – CSTR approach. The former assumes that axial dispersion is prevalent in both sludge bed as well as in the sludge blanket, though the degree of dispersion differs from one to another. The sludge bed and the sludge blanket each is thus equivalent to a PFDR (plug flow dispersion reactor), but the magnitude of axial dispersion coefficient shall be different for both. The latter is the rudimentary model which assumes true plug flow in sludge bed and complete back mixing in the sludge blanket. Such an approach has been discussed by Narayanan and Narayan[3] also. The same is considered here for sake of comparison. Elaborate

experimental data compiled from pilot plants / industrial units have also been employed to substantiate the inferences.

Simulation Models: We shall start with the PFR – CSTR approach (simulation model – 1). As stated earlier, this is the simplest approach to the analysis of UASB bioreactor's performance. Once true plug flow has been assumed in the sludge bed, the performance equation for the same shall be

$$\tau_b = (AL_b\ \varepsilon_{bL}\ /Q_0\ ) = \int_{C_{Sb}}^{C_{S0}} [1/\eta(-r_s)(int)]\ dC_S \qquad (1)$$

where C_{S0} , C_{Sb} = substrate concentration in feed solution and that in solution leaving the sludge bed and entering the sludge blanket respectively, g / L, ε_{bL} = fractional liquid holdup in sludge bed

Similarly, assuming complete backmixing in the sludge blanket (though it is a three phase zone),

$$\tau_f = (\ AL_f\,\varepsilon_L / Q_0\ ) = (C_{Sb} - C_{Se}\ ) / [\ \eta\ (-\mathbf{r}_S\ )(int)\ ] \qquad (2)$$

where C_{Se} is the substrate concentration in the final product solution and ε_L is the fractional liquid holdup in the sludge blanket. It is observed subsequently (on execution of the simulation packages and also from the industrial / pilot plant data compiled) that the volumetric flow rate of feed solution (Q_0) in UASB reactors is necessarily large such that the superficial velocity, U(sup) , of the solution remains larger than the minimum fluidization velocity (U_{mf}) of each sludge granule. The sludge bed thus is not a tightly packed bed, but is a partially expanded bed. It has been assumed that the fractional gas holdup in the sludge bed is relatively low and accordingly, we may assume with allowable error that $\varepsilon_b = (\varepsilon_{bL} + \varepsilon_{bg}\ ) \approx \varepsilon_{bL}$. We have also assumed that the total voidage of sludge bed is more or less constant and equal to 0.40.

The fractional gas holdup (ε_g) and liquid holdup (ε_L) in the sludge blanket are estimated from Dakshinamurthy et.al's correlation[4] and Kato et.al's correlation[5] respectively that have been originally developed for three phase fluidized beds. The former that has been employed to predict the total voidage ($\varepsilon_f = \varepsilon_L + \varepsilon_g$) of the sludge blanket is of the form

$$\varepsilon_f = k\ [U(sup)\ /\ U_t]^m\ [U_g\,\mu_L\ /\ \sigma]^{0.08} \qquad (3)$$

where the correlation constants (k, m) depend on the particle Reynolds number (Re_P) and U_t stands for the terminal free settling velocity of each granule and is estimated as

$$U_t = [(4/3)\ (\rho_P - \rho_L)\ g\ d_P / (f_D\ \rho_L)]^{1/2} \qquad (4)$$

The drag coefficient (f_D) depends on the particle Reynolds number (Re_P) and hence, equation (4) is solved by trial to obtain the value of U_t. The iterative procedure, that is performed with the help of the standard f_D versus Re_P plots, has been found to be fast converging. In the above equation, ρ_P is taken equal to ($x_f\ \psi$), where x_f is the cell mass concentration in each granule and ψ is the correction factor incorporated to account for gas entrapment within the granule. Typically, the value of x_f varies from 1030 to 1050 kg / m^3 and ψ = 0.96 – 0.98. The correlation proposed by Kato and coworkers[5] is of the form

$$\varepsilon_L = k_1\ [\ Usup)\ /\ U_t\ ]^n \qquad (5)$$

where the correlation constants (k , n) depend on the particle Reynolds number (Re_P) and the dimensionless parameter (N) which is defined as , $N = [(\ U_g\ )^4\ \rho_L / (\sigma\ g)]$. As stated earlier, the gas flow rate and thereby the gas velocity (U_g) increases along the height of the reactor. As the substrate solution flows upward, it undergoes more and more bioconversion and more and more biogas is produced. However, in equations (3) and (5), for the computation of fractional fluid holdup in the sludge blanket, an average value of gas velocity has been employed such as

$$U_g = Q_0\ Y_g\ (\ C_{S0} - C_{Se}\ )\ /\ A \qquad (6)$$

where Y_g is the average gas yield in m^3 of gas produced per kg of BOD destroyed. Typically, Y_g = 0.5 to 0.7 m^3 per kg of BOD.

In the simulation model – 2, the performance equation shall be the same for the sludge bed as well as for the sludge blanket, except that the magnitude of axial dispersion coefficient shall be different:

$$-\ U(sup)(dC_S/dz)\ +\ D_L(d^2C_S/dz^2)\ =\ \eta\ (-r_{Sp})(int) \qquad (7)$$

Starting from the top of the sludge blanket where $C_S = C_{Se}$, the boundary conditions for sludge blanket and those for sludge bed shall be

B.C.1 : At z = 0, $C_S = C_{Se}$ (8)

B.C.2 : At $z = L_f$, $C_S = C_{Sb}$ (9)

B.C.3 : At $z = L = (\ L_f + L_b\ )$,

$$U(sup)\ C_{S0} = U(sup)\ C_S\ (\text{at } z \rangle 0\ ) - D_L\ dC_S/dz\ (\text{at } z \rangle 0) \qquad (10)$$

BC – 1 and BC – 2 are for sludge blanket and those for sludge bed are BC – 2 and BC – 3. A value of sludge blanket height (L_f) is first specified. Equation (7) is then solved numerically after substituting the expressions for effectiveness factor (η) and intrinsic rate, ($-\ r_S\ )$ (int) , using a modified form of fourth order Runge – Kutta method. The iterations are continued downward (starting from i = 1, z = 0) until i = n, when L_f (computed) exceeds L_f. To note that , L_f (computed) = (n) (Δz). Then, $C_{Sb} = C_S$ (i +1). Once the value of C_{Sb} has been thus determined, equation (7) is further solved using the same numerical algorithm iteratively starting from $C_S = C_{Sb}$ (when i = 1) to $C_S = C_{S0}$ (when i = m). Then, the height of the sludge bed = L_b = (m) Δz). The total height of the reaction zone (L) is then given by

$$L = L_f + L_b \qquad (11)$$

The kinetic pathways considered are those proposed by Monod, Teissier and Andrews and are given respectively as

$$(-r_s)(int)\ =\ \mu_m(app)C_S\ /\ (K_S + C_S) \qquad (12)$$

$$=\ \mu_m(app)[\ 1 - \exp\ (-C_S/(K_S)] \qquad (13)$$

$$=\ \mu_m(app)C_S\ /\left(K_S + C_S + {C_S}^2/K_{Si}\right) \qquad (14)$$

Monod's kinetic equation (equation – 12) is the simplest of all, while Teissier's equation (equation–13) predicts an exponential relationship between substrate concentration and the intrinsic rate and Andrews' kinetic model (equation– 14) accounts for substrate inhibition to microbial growth as well (K_{Si} = substrate

inhibition coefficient, g / L). In the above equations, μ_m(app) is defined as

$$\mu_m(app) = \mu_m x_f(1-\varepsilon_b)/(Y\varepsilon_b) \quad (15)$$

(for sludge bed)

$$= \mu_m x_f(1-\varepsilon_f)/(Y\varepsilon_L) \quad (16)$$

(for sludge blanket)

For simulation model – 2, experimental values of axial dispersion coefficient (D_L) have been used in the present study. Typically, D_L = 0.007 m^2 /s for the sludge bed and D_L = 0.0315 m^2 /s for the sludge blanket. Attempts to utilize D_L – values computed from experimental correlations reported in literature yielded inconsistent results.

The effectiveness factor (η) that accounts for the resistance to substrate transfer into sludge granule is another important process parameter. With intrinsic rate following Monod's kinetics (equation – 12), η is computed as[6]

$$(1/\eta)^2 = (1/\eta_d)^2 + exp[6(\phi')^2/\{5(1+\beta)^2\} - (1/\eta_d)^2] \quad (17)$$

where

$$\eta_d = (\sqrt{2}/\phi')[(1+\beta)/\beta]\{\beta - ln(1+\beta)\}^{1/2} \quad (18)$$

$$\phi' = (d_p/6)[\mu_m(app)/D_e K_S]^{1/2} \quad (19)$$

$$\beta = (C_{Sp}/K_S) \approx C_S/K_S \quad (20)$$

If the intrinsic rate follows Teissier's equation (equation – 13) or Andrews' kinetic model (equation – 14), then η is estimated from

$$\eta = [3\phi \coth(3\phi) - 1]/(3\phi^2) \quad (21)$$

where ϕ = Thiele – type modulus

$$= (\phi'/\sqrt{2})\,[1 - \exp(-\beta)] \,/\, \{(\beta - 1) + exp(-\beta)\}^{1/2} \quad (22)$$

(for Teissier's kinetic model)

$$= \left(\frac{d_p}{6}\right)(-r_{sp})(int)/\,[2D_e \int_0^{C_{sp}} (-r_s)(int)dC_s]^{1/2} \quad (23)$$

(for Andrews' kinetic model)

The resistance to transport of substrate from fluid bulk to granule surface shall necessarily be small and we have safely neglected the same. Accordingly, C_{SP} (namely, substrate concentration at granule – fluid interface) has been taken equal to substrate concentration in the fluid bulk (namely, C_S). After replacing C_{SP} by C_S, the integral of equation (23) is evaluated numerically using Simpson's rule.

Material and Methods

Experimental data are collected from a number of industrial units and pilot plants. All of them deal with anaerobic treatment of waste water (process effluents) to reduce the BOD. The feed solutions are those discharged from paper and pulp industries, fertilizer plants, food processing units and petrochemical industries. BOD of feed solutions varied from 7.5 to 20 g/L and the feed flow rate (capacity of the unit) from 200 to 300 m^3 / hr. Though laborious, the data collection had been massive and successful. The data had been filtered by inspection and those found reliable have only been utilized for comparison with computed results.

Results and Discussion

The variation of total height of reaction zone (L) with final substrate concentration (substrate concentration in product solution, C_{Se}) is shown in figures-1 to 3. Computations have been performed with a specified value of L_b (height of sludge bed) and based on this value of L_b , the substrate concentration in the exit stream from sludge bed (namely, C_{Sb}) is computed from equation (1) through an iterative procedure. A value of C_{Sb} is first assumed and the integral of equation (1) is evaluated numerically using Simpson's rule from $C_S = C_{S0}$ to $C_S = C_{Sb}$. For this, the value of effectiveness factor (η) at each value of C_S is computed from equations (17) to (23). In case of Andrews' kinetic model, computation of η further demands numerical evaluation of integral of equation (23), once again by Simpson's rule. It is now checked whether the value of L_b computed from equation (1) agrees with that specified at the outset. If not, computations are repeated with an alternate value of C_{Sb}. Substrate concentration at sludge bed exit is thus finalized by trial. Convergence could be achieved within 10 – 15 iterations. Thereafter, the required height of sludge blanket is estimated from equation (2) and the total height of reaction zone required (L) from equation (11). The fractional liquid holdup (ε_L) and fractional gas holdup (ε_g) in the sludge blanket are determined from equations (3) to (6), which also involves an iterative procedure as described earlier. Results at L_b = 1.0 m are presented in figures-1 to 3 as illustrative examples, though computations have been performed at different values of L_b such as L_b = 0.5 m, 0.6 m, 0.75 m and 0.8 m. Figure-1 displays data when intrinsic rate follows Monod's kinetics (equation – 12), figure-2 data based on Teissier's kinetic equation (equation – 13) and figure-3 provides data computed based on Andrews' kinetic model (equation– 14). The plots demonstrate data computed from the simulation packages and also the experimental values. Results from software package reported by Narayanan and Narayan[3] are also included for comparison.

It can be seen from plots (figures – 1,2,3) that the value of L decreases with increase in C_{Se}. This is understandable since a large value of final substrate concentration implies lower fractional conversion and consequently, the volume of reaction zone required shall be lower. The pattern of variation is similar with all the three forms of kinetic equation. It is observed that Andrews' kinetic model typically demands larger height of reaction zone for the same fractional conversion when compared to data on other kinetic models. This brings to the conclusion that substrate inhibition does play a role in systems of this kind (anaerobic digestion of organic matter). It is also found that experimental data collected from industrial units / pilot plants agree more closely with data computed based on Andrews' model.

Variation of size of reaction zone (L) with feed concentration (C_{S0}) and that with feed flow rate (Q_0) are illustrated in figures-4 to 9. Plots are given for all the three kinetic pathways considered. As anticipated, the required height of reaction zone

(L) increases with increase in feed concentration (C_{S0}) and feed flow rate (Q_0). Though the variations are similar in nature, it is seen that data based on Andrews' kinetic model are in better agreement with experimental data. These plots also demonstrate that simulation model – 2 (PFDRs – in – series model) predicts the required height of reaction zone (L) for attaining a specified value of fractional conversion (α) at any specified feed concentration / feed flow rate more accurately than simulation – 1. This is particularly true at large values of α (low values of C_{Se}) and when the reactor handles high strength feedstocks (high values of C_{S0}) and operates at high capacities (high values of Q_0). The software proposed by Narayanan and Narayan[3] also predicts values of L comparable with those predicted by Simulation model – 2.

This yields the inference that axial dispersion does influence the reactor's performance and the same cannot be neglected. The sludge is thus better modeled as a PFDR (plug flow dispersion reactor) with a specific value of axial dispersion coefficient (D_L). It also helps us conclude that the sludge bed is not a tightly packed bed but is a partially expanded bed. This is further substantiated by the fact that within the range of flow rate (Q_0) considered, the superficial velocity of feed solution does exceed the minimum value required to fluidise the sludge bed.

Though the PFDRs – in – series model and the software package proposed by Narayanan and Narayan[3] predict comparable results (at least within the range of process parameters considered here), the former is computationally simpler and it employs two separate values of axial dispersion coefficient (D_L) for the sludge bed and the sludge blanket. However, it demands experimental values of D_L and it does not consider variation of D_L in the axial direction (due to change in gas flow rate). The simulation package reported by Narayanan and Narayan[3] is computationally more complex and it considers the entire bioreactor as equivalent to a single PFDR. Their package, however, takes into account the variation of gas flow rate along the height of the reactor and accordingly, D_L (which is computed as a function of gas velocity) is also taken as a function of axial distance (z). At low values of α (high values of C_{Se}) and while handling low strength feedstocks (low values of C_{S0}) at lower capacities (lower values of Q_0), simulation model – 1 also predicts L – values that agree within 12 – 15 % deviation with experimental values.

Conclusion

Performance analysis of upflow anaerobic sludge blanket (UASB) bioreactors has been performed by developing simulation packages and by verifying how far the package results agree with real – life experimental data. The simplified approach involved in simulation model – 1, which neglects axial dispersion and considers the bioreactor to be equivalent to PFR – CSTR combination, is reasonably reliable at low capacities, while handling low strength feedstocks and when the fractional conversion of substrate (α) desired is low. To keep in mind that UASB bioreactors usually handle high strength feedstocks and operate efficiently at high capacities. They also provide large scale BOD removal. The PFDRs – in – series model and the simulation package reported by Narayanan and Narayan[3] are more reliable at high capacities, with high – strength feedstocks and when α (required) is large. In other words, axial dispersion is an influencing parameter and for an accurate prediction of bioreactor's performance, its effect is to be duly accounted for. The sludge bed is to be taken equivalent to a partially expanded bed. The utility of the reported package of Narayanan and Narayan[3] heavily depends on the availability of a reliable correlation for the estimation of axial dispersion coefficient, D_L (z), though the approach in their package is more rigorous. On the other hand, the PFDRs – in – series model is reasonably accurate (whenever the change in gas velocity along the reactor height is not too controlling), but it demands dependable, experimental values of D_L. Among the kinetic pathways considered, Andrews' kinetic model that accounts for substrate inhibition to microbial growth, appears more applicable for the systems handled (all involving anaerobic bioconversion of organic matter). The package, however, is equally applicable to all alternate types of intrinsic kinetics of bioconversion. The versatility of the simulation packages is convincing, since the dependence of bioreactor's performance on different process parameters such as fractional conversion desired, feed concentration, feed flow rate and type of kinetic pathway involved can be reliably deduced by executing these packages. The package results are also in good agreement with real – life industrial / pilot plant data.

Nomenclature: A = cross – sectional area of reactor column, m^2 , C_S = substrate concentration in liquid bulk, g / L, C_{S0} = substrate concentration in feed, g / L, C_{Se} = substrate concentration in exit stream from reactor, g / L, C_{SP} = substrate concentration in liquid at granule – fluid interface, g / L, d_P = diameter of granule , m, D_e = effective diffusivity of substrate into the granule, m^2 / s, D_L = axial dispersion coefficient, m^2 / s, f_D = drag coefficient, dimensionless, K_S = kinetic constant, g / L, K_{Si} = substrate inhibition coefficient, g / L, L= total height of reaction zone, m, L_b = height of sludge bed, m section and that of packed bed section respectively, m, L_f = height of sludge blanket, m, Q_0 = volume flow rate of feed solution, m^3 /s, $(-\mathbf{r}_s)$(int) = intrinsic rate of bioconversion, g / (L.s), Re_P = particle Reynolds number, dimensionless, U(sup) = superficial velocity of substrate solution, m / s, U_g = average gas velocity, m / s, U_t = terminal free velocity of each granule, m / s, x_f = cell mass concentration in each granule, g / L, Y = overall yield coefficient for cell mass production, dimensionless, β = dimensionless parameter defined in equation (20), ε_b voidage (porosity) of sludge bed, dimensionless, ε_{bg} = fractional gas holdup in sludge bed, dimensionless, ε_{bL} = fractional liquid holdup in sludge bed, dimensionless, ε_f = voidage of sludge blanket, dimensionless, ε_L = fractional liquid holdup in sludge blanket, dimensionless, ε_g = fractional gas holdup in sludge blanket, dimensionless, η = effectiveness factor, dimensionless,

η_d = dimensionless parameter defined in equation (18), τ_b = space time for sludge bed , s, τ_f = space time for sludge blanket , s, μ_L = liquid viscosity, kg / (m.s), μ_m = kinetic constant, s^{-1}, ρ_L = liquid density, kg / m^3 , ρ_P = density of each granule, kg / m^3, σ = interfacial tension, N / m, ϕ = Thiele – type modulus, dimensionless, ϕ' = dimensionless parameter defined in equation (19), ψ = correction factor for gas entrapment, dimensionless.

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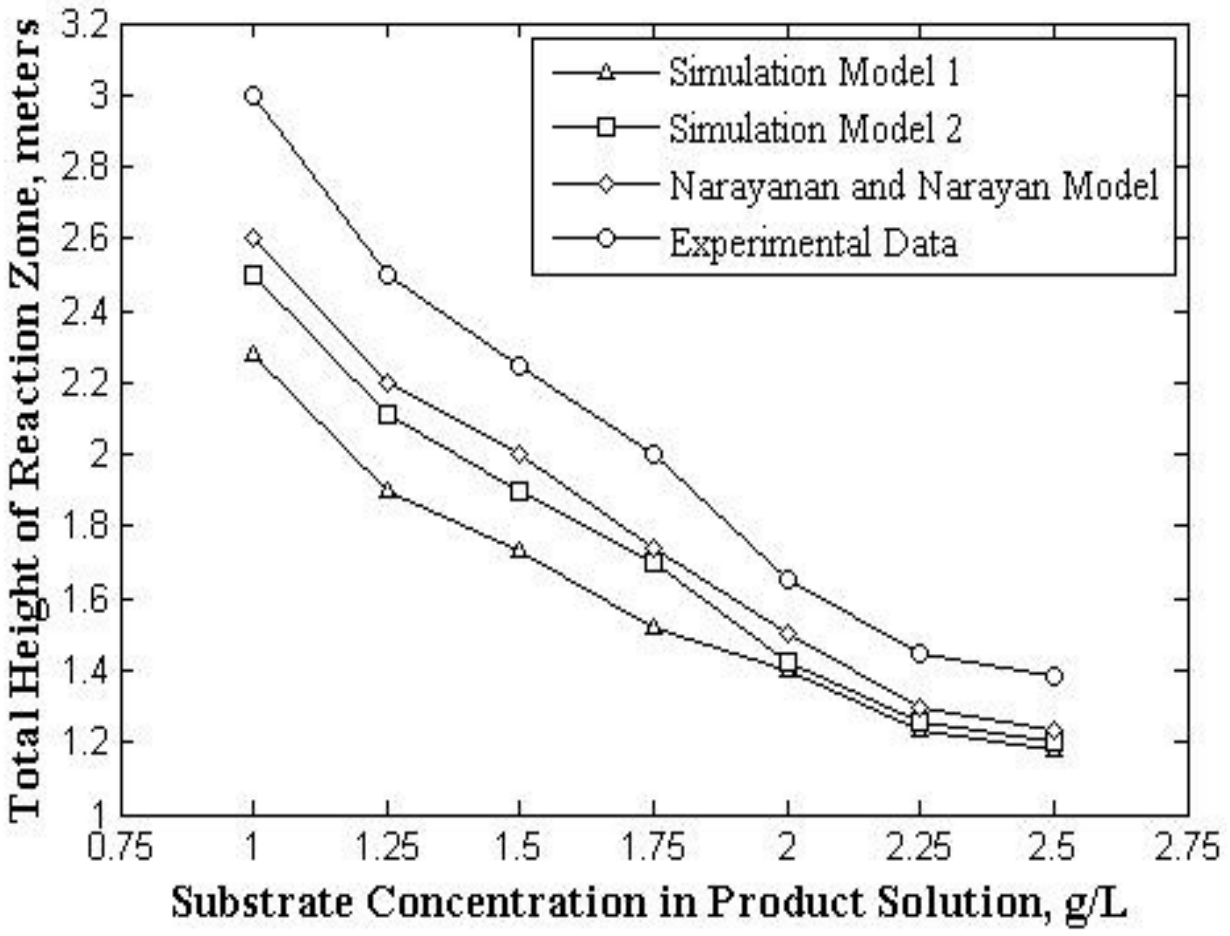

Figure-1
Variation of Total Height of Reaction Zone (L) with Substrate concentration in Product Solution, C_{Se} (Monod's Model)

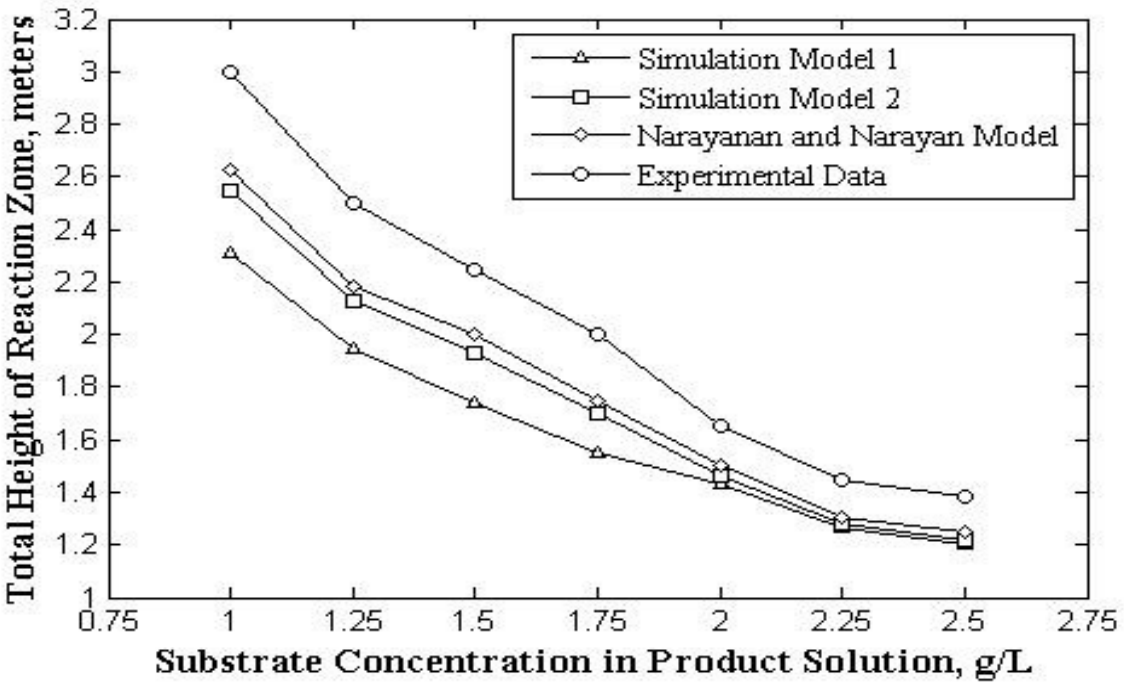

Figure-2
Variation of Total Height of Reaction Zone (L) with Substrate concentration in Product Solution, C_{Se} (Teissier's kinetic model)

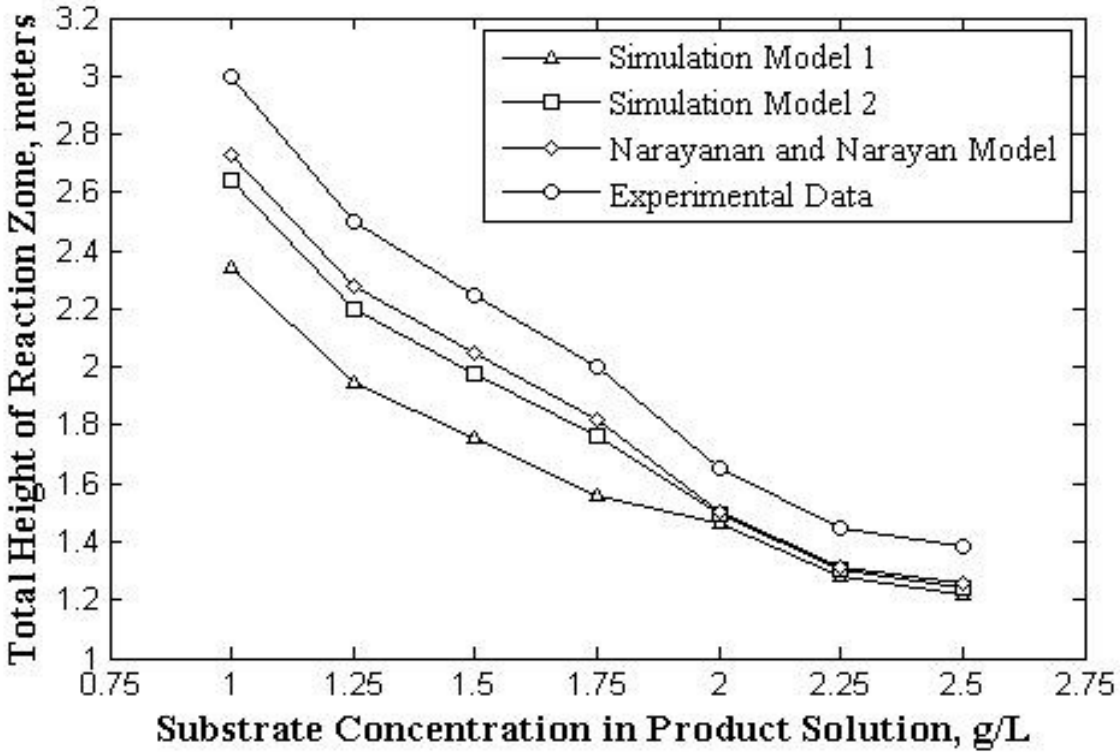

Figure-3
Variation of Total Height of Reaction Zone (L) with Substrate concentration in Product Solution, C_{Se} (Andrews' kinetic model)

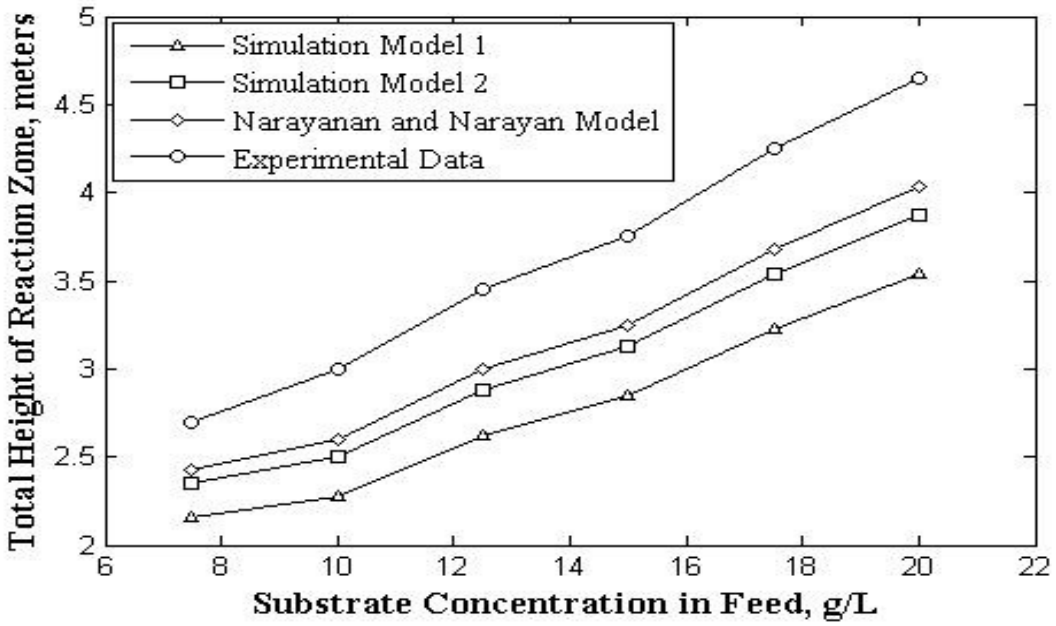

Figure-4
Variation of Total Height of Reaction Zone (L) with Substrate concentration in Feed, C_{S0} (Monod's Model)

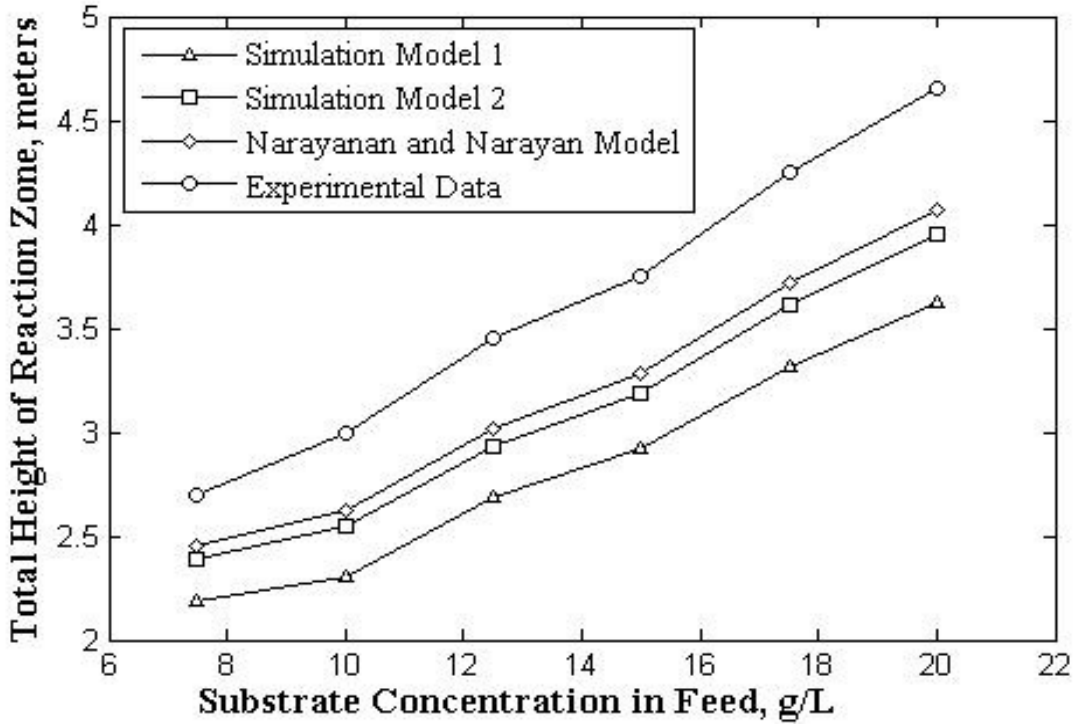

Figure-5
Variation of Total Height of Reaction Zone (L) with Substrate concentration in Feed, C_{S0} (Teissier's kinetic equation)

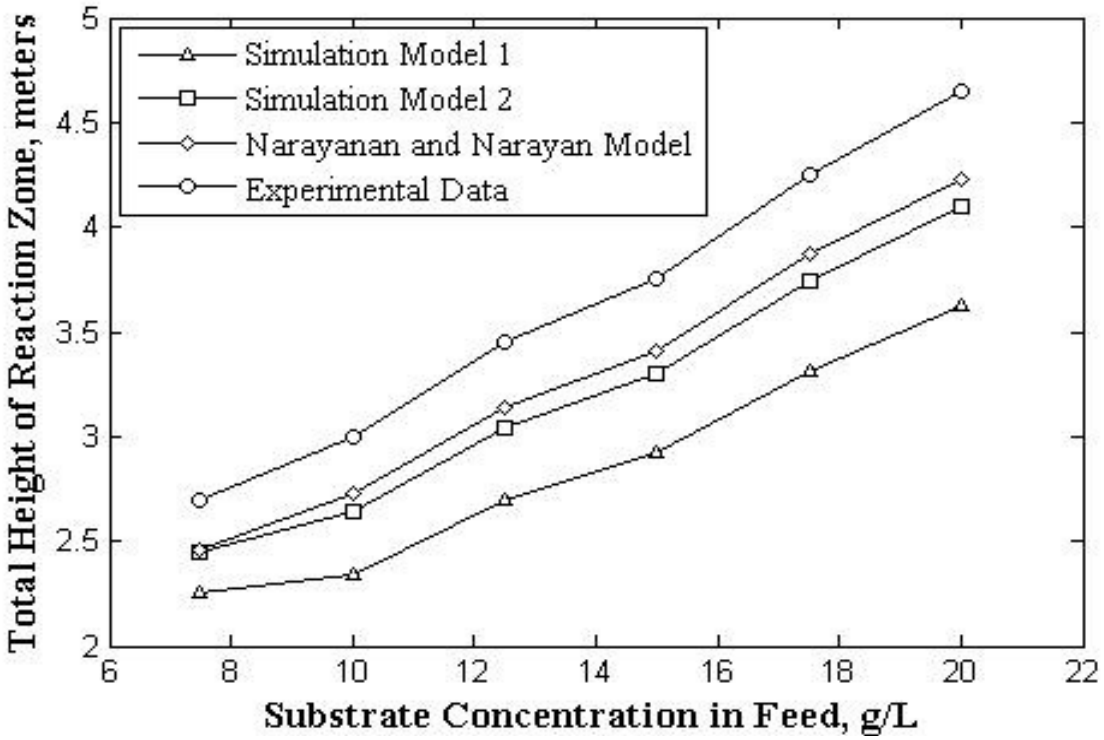

Figure-6
Variation of Total Height of Reaction Zone (L) with Substrate concentration in Feed, C_{S0} (Andrews' kinetic model)

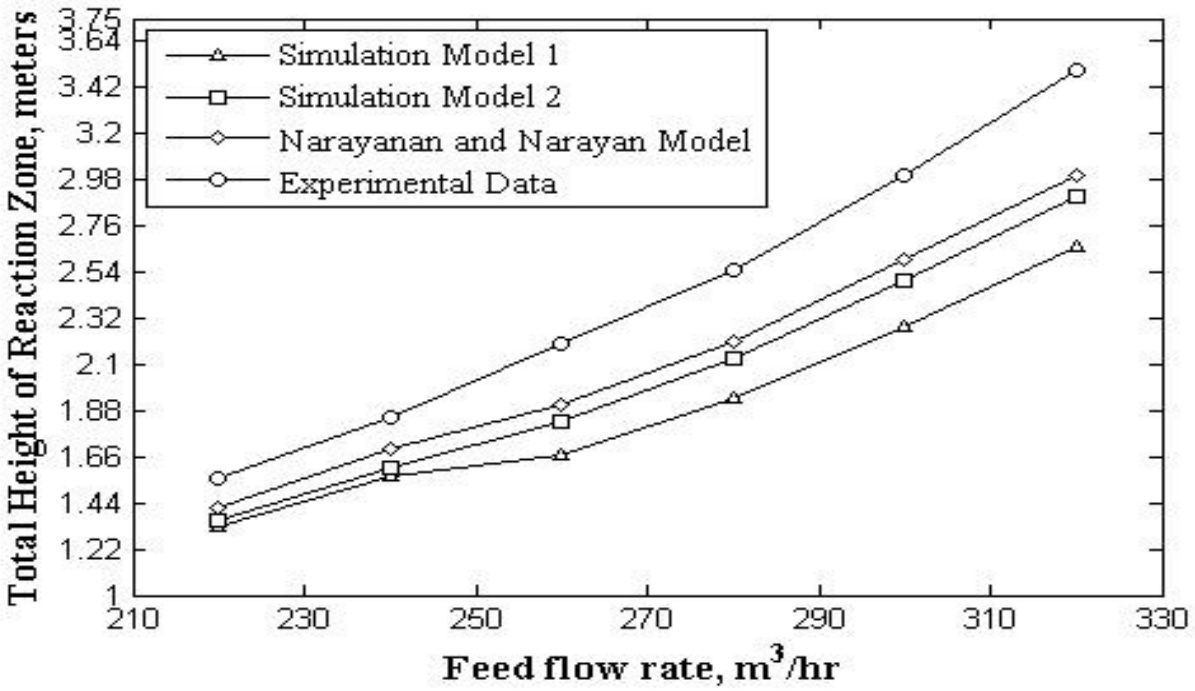

Figure-7
Variation of Total Height of Reaction Zone (L) with Feed flow rate (Monod's kinetics)

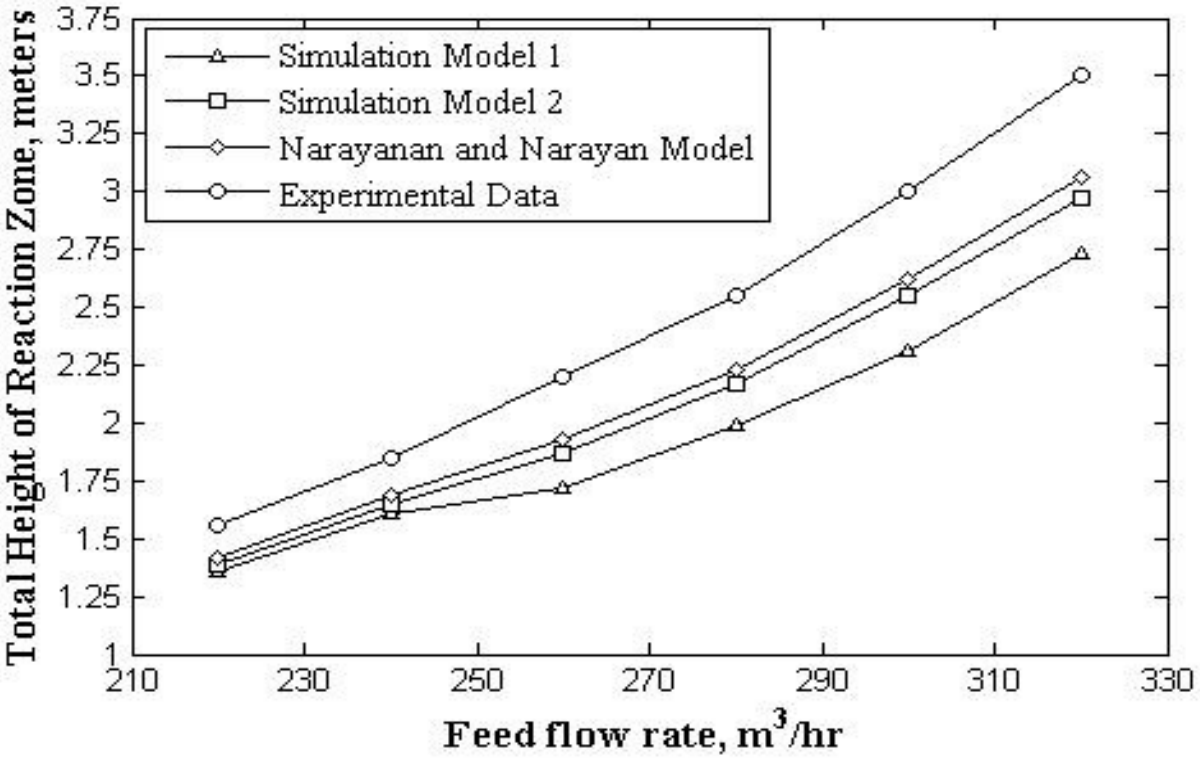

Figure-8
Variation of Total Height of Reaction Zone (L) with Feed flow rate (Teissier's kinetic model)

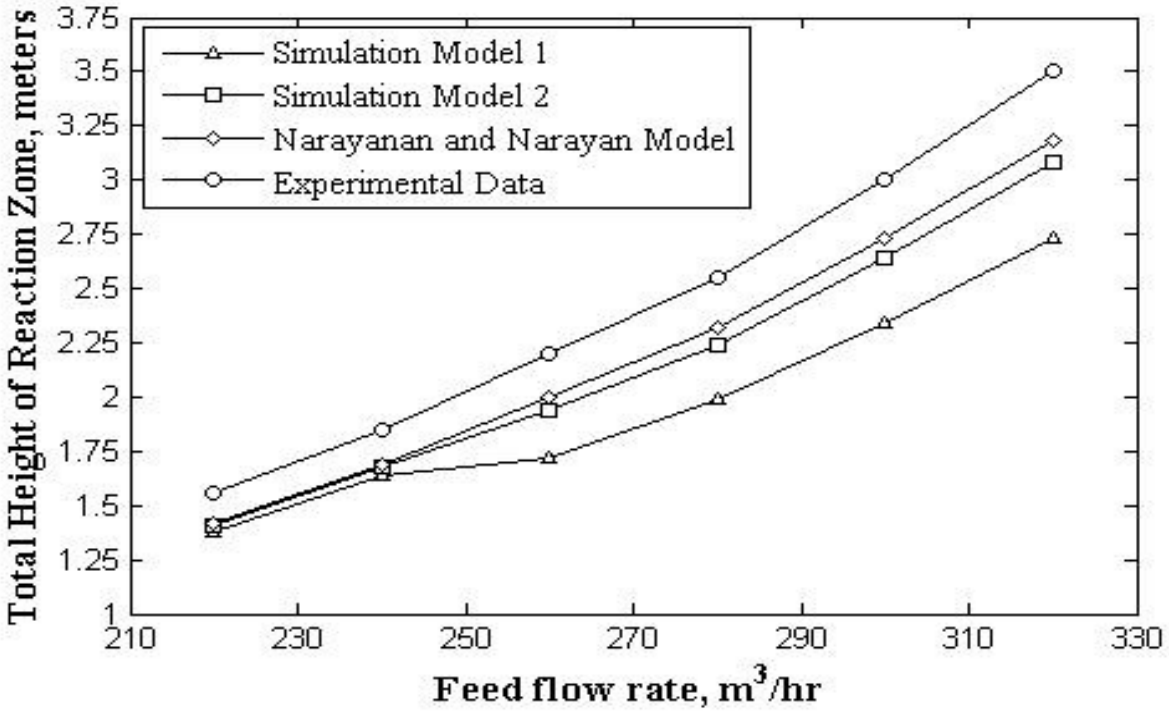

Figure-9
Variation of Total Height of Reaction Zone (L) with Feed flow rate (Andrews' kinetic model)