PHYSICAL TECHNOLOGY

Non-Newtonian Fluids – Agitation

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Introduction

ALTHOUGH there is a negativism in the name, the non-Newtonian fluids are very common; so much so that we cannot spend our daily lives without constantly coming across them: starting from butter, honey, ketchup, taking through most of our liquid eatables, most cosmetics, liquid medicines and even the writing ink. Yes, so many liquids are non-Newtonian – in the sense that they are not Newtonian. They do not follow the simple Newton's law where the flow is directly proportional to the force. Processing of non-Newtonian fluids is in itself an involved topic. Here, to introduce our readers to the subject, we shall briefly describe the process of 'Agitation' in a rather general way with an idea to bring out some of the crucial points regarding this class of fluids.

Non-Newtonian fluids

"Viscosity" is a parameter that is defined as a ratio between shear stress and shear rate, when a Newtonian liquid is sheared in a precisely defined shear flow. Non-Newtonian liquids do not have a constant viscosity. Different types of non-Newtonian fluids are briefly described below (the details may be referred to in any text book on 'Rheology').

Pseudoplastic or Shear thinning fluids

For these types of fluids, the viscosity decreases as shear rate increases (Fig. 1 and 2). Polymer solutions, melts, greases and some pharmaceutical preparations are pseudoplastic.

Dilatant or Shear thickening fluids

For these types of fluids, the viscosity increases with increasing shear rates. Starch suspensions, gum solutions, aqueous suspension of titanium dioxide, etc., show a dilatant behaviour.

Fluids with yield stress

There are certain materials which do not flow unless the stress applied exceeds certain minimum value. This minimum stress is called "yield stress" and generally is connected to the structure of the fluid. Fig. 2 compares the behaviour of such fluids with pseudoplastic and dilatant fluids.

Tooth paste, emulsion, paints, inks, jelleys, fermentation broth, grease, food preparations, and blood exhibit yield stress.

Thixotropic Fluids

These fluids exhibit a reversible decrease in shear stress with time at a constant rate of shear and fixed temperature. The shear stress will, of course, approach some limiting value.

Oil well drilling muds, some polymer melts, tomato ketchup or some other food preparation, some greases, etc., show this type of behaviour.

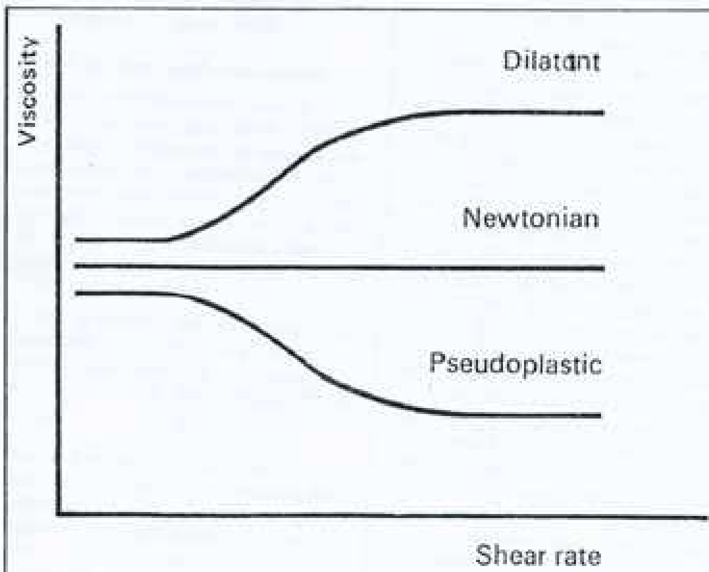


Fig. 1: Typical viscosity curves.

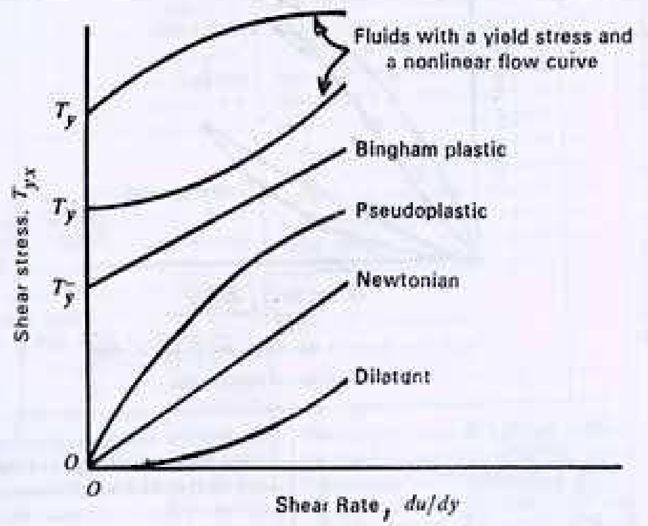


Fig. 2: Flow curves on arithmetic coordinates for various types of time-independent fluid.

Rheopectic Fluids

These exhibit a reversible increase in shear stress with time at a constant rate of shear at a fixed temperature. These materials are relatively less in number. Materials which show this behaviour are bentonite clay suspensions, V_2O_5 suspensions, gypsum suspensions, dilute suspensions of ammonium oleate, etc.

(If stress is measured, when shear rate is steadily increased from zero to a maximum value and then immediately decreased steadily towards zero, a hysteresis loop would be obtained as shown in Fig. 3).

Viscoelastic Fluids

Fluids which show partial elastic recovery upon the removal of a deforming stress and exhibit process properties of both liquids (visco-) and solids (-elastic) fall into this category.

These fluids exhibit the properties of a viscous fluid and elastic solid. In a purely Hookean solid, the stress corresponding to a given strain is independent of time, whereas, for viscoelastic fluids, the stress will gradually dissipate. On the other hand, in contrast to purely viscous fluids, viscoelastic fluids flow when subjected to stress but a part of their deformation is gradually recovered upon removal of the stress.

Due to elasticity, these fluids show a markedly different behaviour. The rod climbing effect or Weissenberg effect or recoiling tendency or thread formation are all due to elastic properties. These fluids exhibit normal stress differences. How the simpler situations are dealt with is given below.

Flow of Non-Newtonian Fluids

The use of rheological models based upon rheomatric data help in deciding the range of shear rates or throughputs over which the applicability of design equations may be valid. Thus, it is not enough to know the viscosity at one shear rate. One should always have the values of viscosity in the vicinity of shear rate of interest.

Let us consider the horse power calculations for pipe line transport. Usually one plots wall shear stress $\tau_w = \frac{\Delta p}{4L}$ against $\frac{8v}{D}$ where 'v' is linear average velocity. On a large scale, if the flow remains laminar, then one has to read τ_w from the graph or calculate the wall shear rate and estimate corresponding τ_w. Since diameters are different pressure gradient, $\Delta p/L$ will be different.

For pipe flow the wall shear rate is given by

$$\dot{\gamma}_w = \frac{8v}{D}\left[\frac{3n+1}{4n}\right] \quad ...(1)$$

For power-law type of fluids

$$\frac{D\,\Delta P}{4L} = \tau_w = K\left[\frac{8v}{D}\left(\frac{3n+1}{4n}\right)\right]^n \quad ...(2)$$

The above equation shows that, if n = 0.5, the pressure drop would go up by a factor of 1.4 if the flow is doubled. But if n = 1.3, the pressure drop would increase by a factor of 2.46.

Flow patterns and velocity distribution in agitated vessels

The rheological complexities associated with the liquid being agitated can very significantly influence the flow patterns and velocity distribution. The nature of the rheological complexity will, of course, govern the net influence on the hydrodynamics. The influence of shear dependent viscosity is most easy to understand and analyse. For instance, one would expect that, for pseudoplastic (shear thinning) fluids, the apparent viscosity of the fluid in the near impeller region should be rather low and increase progressively as one moves away from the impeller. This will consequently result in high velocities and velocity gradients in the near impeller region, which will die away rapidly from the impeller. The early photographic studies did confirm this. The singularly important result which were able to obtain on the basis of their velocity distribution measurement was that the average shear rate in the vessel was linearly proportional to RPM. We shall see later that this result plays an extremely important role in the prediction of power requirements in non-Newtonian fluids. It should be noted here that near solidification of a dilatant fluid in the region of the impeller is observed. The size of this core increases rapidly with rotational speed. The effect becomes particularly important as the ratio of the tank diameter to the agitator diameter is increased. The angular velocity distribution measurements made on anchor agitators are also in agreement with the trends observed by considerably flatter velocity profiles for a shear thinning fluid in comparison to a Newtonian fluid.

Fluid elasticity can very significantly influence the flow patterns around agitators. In fact, a fundamental approach to the understanding of velocity distribution around agitators may be to study the rotational flows around simple bodies, such as spheres and discs. The advantage here is that, at least, the shape complexities of the agitators (which preclude the possibility of any theoretical study) are avoided and the velocity distribution can be viewed from a theoretical angle.

If one considers a third order viscoelastic fluid which portrays the characteristics of a finite shear thinning viscosity and elasticity (manifesting itself in the form of finite normal stress difference in viscometric flows), then, for a sphere rotating at a given angular velocity, one can calculate the velocity profile.

The projections of streamlines on a plane containing the axis of rotation have been shown in Fig. 4. It is evident that different flow situations will arise for different values of the ratio of inertial stresses to normal stresses. This ratio plays an important part in determining the flow field around spheres, and is likely to play an equally important part in describing the flow field around agitators, and consequently in determining the circulation and mixing times.

In fact, such a flow reversal has also been in the case of a disc rotating in viscoelastic fluids. The changes in flow patterns due to the interaction of elasticity and inertia in the case of commercial agitators and found qualitatively similar flow patterns.

An extensive investigation of the influence of rheological properties on flow patterns, especially with a view to investigate its implications in bulk polymerising styrene to polystyrene, is undertaken. Spheres, discs and turbines are used as agitators. The observed flow patterns are in good agreement. A screw-propeller agitator is also used. In Newtonian fluids, the action of the screw propeller is to push the fluid vertically in the tank. In the case of viscoelastic fluids, it was found that normal stress effects reinforced the screw propulsion mechanism. When flow patterns during bulk polymerization of styrene were observed, it was found that, at low conversions, the flow patterns around the sphere and other agitators involved a primary axisymmetric flow and secondary flow, consisting of a fluid being drawn in at the poles axially and expelled at the equator. As the conversion increased, the secondary flow field around a sphere divided into two regions, one adjacent to the sphere consisting of closed circulating motions and the second at larger distances from the sphere. Similar flow patterns are observed with other agitators.

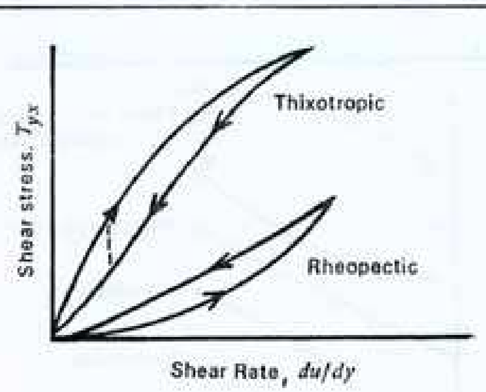


Fig. 3: Flow curves for thixotropic and rheopectic fluids in single continuous experiments.

The implications of such changes in the secondary flow patterns on the performance of a mixing vessel should be clearly appreciated. The photographic study is particularly important in this connection. Their velocity distribution measurements in viscoelastic fluids showed quite clearly that, although the secondary flow patterns can be changed quite significantly, the primary flow remains practically unaffected. Since the major contribution to the shear stresses in the vicinity of the rotating body comes from the variation of the primary flow velocity component with radial distance, the power consumption appears to be relatively unaffected by the modifications brought in due to elasticity. However, the secondary flow patterns will have a significant effect on residence-time distribution, molecular-weight distribution (in the case of polymerization reactors), mixing times, circulation times and related kinematically controlled processes.

The influence of fluid elasticity on the velocity distribution in vessels stirred with helical agitators has not been extensively studied. Some qualitative observations showed that, for highly elastic liquids and atleast for some geometrical configurations, the angular flow increased, whereas the axial flow was damped considerably. The detailed velocity distribution studies support this. Fig. 5 shows some typical data obtained. For a 2% CMC solution (which showed negligible elasticity), the strong axial upflow and downflow can be clearly seen, whereas, for a 1% PAA solution (which was strongly elastic), the axial flow can be seen to be significantly damped. The mechanisms responsible for such a phenomenon are not quite clear. However, it is suspected that the peculiar behaviour of elastic liquids in kinematic fields of rapidly changing deformation is responsible.

Discharge rates, circulation capacities and circulation times

From the velocity measurements, an important integral hydrodynamic quantity can be easily obtained. It is the overall discharge rate or the average time elapsed between the successive passes of a fluid element through the impeller. A comprehensive summary is given in the literature of the investigation of turbines and propellers where the calculations for the overall radial and axial flow are done from the velocity measurement.

Such information, however, pertains to be $> 10^5$, i.e., for turbulent conditions. The circulation numbers on the basis of equal power consumption is then compared. The rotating disc required about 300 times more energy to give the same discharge rate as an eight-bladed turbine. The circulation numbers for propellers were typically in the region of 0.4 to 0.6, whereas for turbines they were in the region of 0.6 to 2.

In the literature, circulation capacities (and times) for high viscosity liquid agitation with helical impellers have been well summarized. The circulation number varies between 0.4 to 1 for ribbons and between 0.01 to 0.5 for screws in a draft tube. These are independent of Reynolds number (for $Re < 10$ or $Re < 50$) but are strongly influenced by the geometrical configuration. Studies with a large number of purely viscous liquids revealed that circulation numbers were not at all influenced by shear-thinning characteristics. The observations with strongly elastic liquids, however, showed that the circulation was damped as a result of elasticity and the circulation numbers were lowered quite significantly. A correlation which adequately describes the data for purely viscous fluids and method to correlate the data on elastic liquids for a given geometrical configuration is provided. The peculiar flow reversals and the two zone velocity distribution regions arising out of the presence of elasticity should strongly influence the circulation times. The enhanced vertical circulation in elastic liquids and for anchor agitated vessels and for propellers would imply that the circulation numbers may increase on account of elasticity rather than decrease as in the case of helical agitators.

Specific comments need to be made concerning the influence of circulation times and circulation time distribution in polymerization reactors. An interesting study in this context is mentioned with specific reference to condensation polymerization of bifunctional polymers. The study is with a specific reference to a turbine impeller which produces circulation loops of the polymer. The viscosities of the polymer mass are so high that material in the circulation loops remains segregated. The water of condensation is removed by diffusive transport to the surface of the polymerizing mass. The contribution of the diffusive process to the rate of reaction is approximated by assuming that the water concentration is proportional to the circulation time of the segregated fluid elements. The effect of mixing upon the product molecular weight distribution has been found and it has been shown that the assumed mixing pattern significantly alters it. Indeed, if the circulation time is low at the end of polymerization, the polymer may be expected to have poor uniformity. This example clearly illustrates the role of circulation loops in batch polymerization systems.

Power Consumption

The space available here is too limited to cover the enormous power consumption data which have been published over the years, nor is it the purpose of this review. The published reviews give a good appraisal of such information. At present, one can make a reliable estimate of the power consumption for various impeller designs and for a number of geometrical arrangements for Newtonian liquids. A great deal of information is also available on non-Newtonian liquids and multiphase systems. We will present here a viewpoint, so that the data may be critically analysed before the use for design.

Since power consumption was first measured for rotating impellers in the late 19th century, the data have often been correlated using dimensional analysis. It may be expected to be a function of the following variables:

(i) Geometrical Variables: Impeller diameter (d), tank diameter (t), Liquid depth (h)
(ii) Material Properties: Density (ρ), Viscosity(μ)
(iii) Process Variables: Rotational Speed of the impeller (N), gravitational acceleration (g)

The relationship can simply be written as,

$$P = f(d, D, h, \rho, \mu, N, g) \quad \ldots\ldots \quad (3)$$

Dimensional analysis give,

$$\frac{P}{d^5N^3\rho} = \bar{K}\left(\frac{d^2N\rho}{\mu}\right)^{c_1}\left(\frac{dN^2}{g}\right)^{c_2} \ldots\ldots (4)$$

Where K is a constant for particular set of geometrical variables and could be obtained as

$$\bar{K} = \bar{K}_1\left(\frac{t}{d}\right)^{a_1}\left(\frac{h}{d}\right)^{a_2} \ldots\ldots\ldots (5)$$

The dimensionless groups $(P/d^5N^3\rho)$, $(d^2N\rho/\mu)$ and (dN^2/g) are known as power number (Po), Reynolds number (Re) and Froude number (Fr), respectively.

The above analysis is valid only under the following conditions: (a) either a single liquid or two miscible liquids having similar properties are present in the vessel, (b) the temperature changes due to energy dissipation are so small that variation in fluid properties because of the temperature changes is negligible, (c) the flow behaviour of the liquid can be characterized by a single parameter, i.e., viscosity. Thus the liquid is Newtonian.

The values of $\bar{K}$, c_1 and c_2 have been determined by many for several different geometrical variables. c_1 assumes different values in the range of -1 to 0 depending upon the operating range of Reynolds numbers. It has been found to have a value of -1 below a certain critical value of Reynolds number. This critical Reynolds number varies between 10 to 100 and it seems to be influenced by the geometry of the system. At a very high Reynolds number, ($Re > 10^4$) power number becomes independent of Reynolds number. The region where c_1 is -1 is commonly understood as laminar region. The effect of Froude number on the power consumption would be apparent only when some power is consumed in producing significant waves on the surface of the liquid or in sustaining a vortex in liquid around the impeller shaft. It is found that, for unbaffled vessels, these effects are negligible below Reynolds number of 300. This limit is exceeded for $Re > 10^3$ when the baffle system is adequate or when the impeller is suitably off-centred. The effect of Froude number is so negligible as to be indeterminate, except by a very accurate dynamometer. When such effects are present, a practical method of analysis is available. For standard designs, $\bar{K}$ and c_1 in equation (4) can be easily obtained. However, to obtain a generalized correlation, a series of experiments with wide variation in geometrical variables will be required.

The first attempt to incorporate the parameters which describes the deviation of any inelastic fluid from the Newtonian characteristics involve the use of results obtained in a pipe flow. This analogy was first suggested and it has been shown that such a Reynolds number does not give a unique power curve for a wide range of flow behaviour index. It is stated that this is not a failure of the analogy but of the power-law approximation. It is suggested that an alternative method involving the use of models has conclusively shown that, even with such models, pipe flow analogy does not offer a unique curve, even in the laminar region.

In order to obtain a power correlation in terms of the rheological parameters, the knowledge of shear rates (atleast in the immediate vicinity of the impeller) is necessary. In an extremely significant contribution, the average shear rate in the vessel can be assumed to be directly proportional to the rotational speed, atleast in the laminar region. This shear rate can be obtained as follows. First, it is necessary to obtain the plot of Po-Re for Newtonian fluids for the system under consideration. Then the power number is calculated from the power data for the non-Newtonian liquids. For this power number the corresponding Reynolds number could be obtained from the Newtonian plot. The average viscosity given by the Reynolds number would give the corresponding shear rate from the viscometric data. A relationship such as

$$\text{Av. Shear rate} = k_s N \quad \text{.........} \quad (6)$$

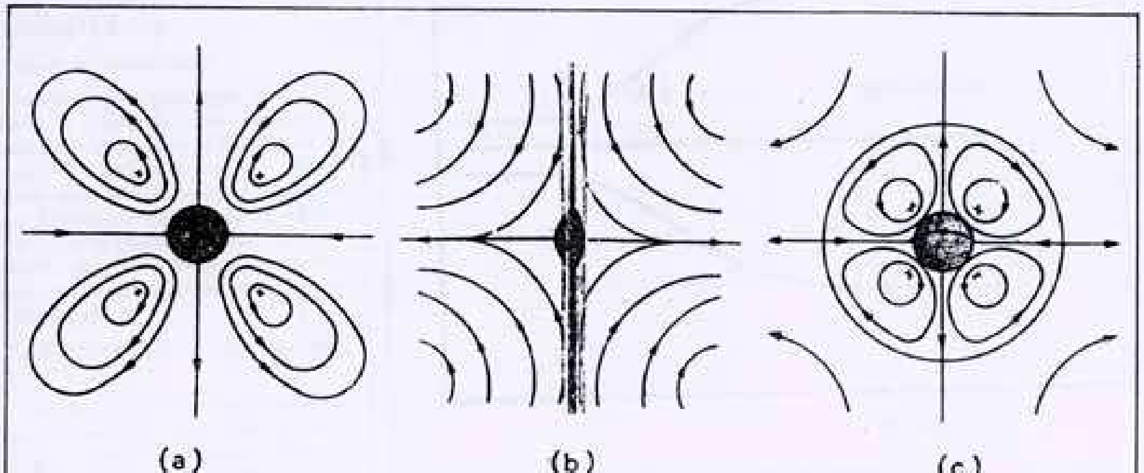
has been obtained for various impellers and the values of k_s have been reported. For the agitation of pseudoplastic liquids by turbines, propellers and paddles k_s assumes a value between 10 and 13. For dilatant liquids a linear relation (upto $n = 1.5$) is obtained. The validity of the linear relationship in equation (C) was subsequently established. From the available information, it appears that one can accept a linear relationship under creeping flow conditions. However, there is some uncertainty about the extension of this method for higher Reynolds number and also about whether k_s can have a universal value even in the low Reynolds number range. The results on turbine show that the generalized Reynolds number $Re = \frac{d^2N}{\mu_a}$ (Where μ_a corresponds to a viscosity at 'k_sN') can give a unique curve upto $Re = 160$ (Laminar region ends at $Re = 10$) within $\pm$ 20 per cent. However, this is not observed in every case, that such extension is not justified on account of insufficient evidence. The constant of proportionality, k_s, can otherwise be a constant for a particular pseudoplastic liquid and geometrically similar system. In other words, k_s should be generally a function of geometry and rheology. The problem can best be overcome by considering a power-law relationship. In creeping-flow (or laminar) region one can obtain a relationship

$$P_o = \bar{K}(k_s)^{1-n}\left(\frac{d^2N^{2-n}}{K}\right)^{-1} \quad \text{.........} \quad (7)$$

From experimental data on power consumption on Newtonian and inelastic liquids and from the flow curve, k_s can be measured (as described earlier). Also, the range of shear rates will be known to give K and n. k_s is obtained in this way and related to the geometry and rheology.

After knowing the operating rotational speeds and the values of k_s from literature an approximate estimate of the shear rates can be obtained. When k_s is dependant on n, a trial and error procedure may be needed to estimate k_s. However, a rough estimation could be possible from the previous information in similar situation. Thus, for example, for helical impellers, k_s varies from 10 to 100, then if rotational speeds range from 0.1 to 10 sec^{-1} then a range of shear rates of 1 to 1000 sec^{-1} may be appropriate.

Note that equation (6) does not give an average shear rate in the entire vessel in all the regimes. The shear rates thus obtained can only characterize the flow near the impeller and that too



(a) (b) (c)

Fig. 4: Secondary flow around a sphere rotating in a viscoelastic liquid.
(a) Normal forces dominating. (b) Centrifugal forces dominating.
(c) Normal and Centrifugal forces comparable

only in the laminar region. Even in the laminar region, the shear-rates at the tip of the impeller are as much as seven times higher than those given by equation (6). At higher Reynolds numbers, the maximum shear rates in the vortices behind the blade have been found to be of an order of magnitude higher.

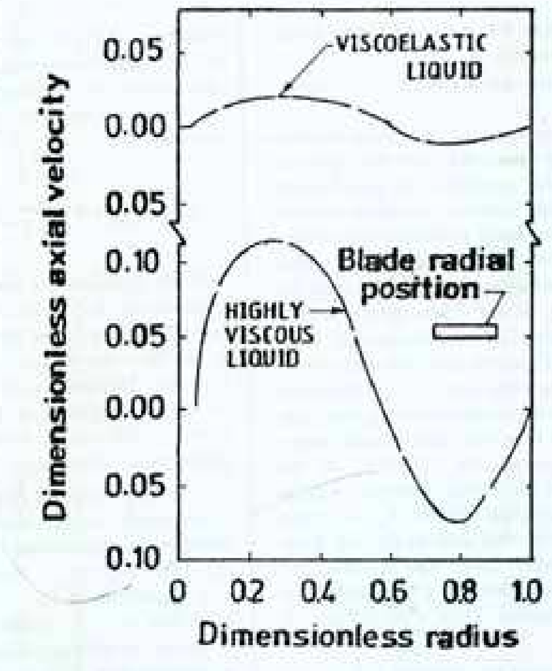


Fig. 5: Axial velocity distribution in a ribbon agitated vessel.

The influence of fluid elasticity on power consumption is negligible under creeping flow conditions. At higher Reynolds numbers, however, it appears that elasticity suppresses the secondary flows and one obtains a reduction in power consumption in comparison to a purely viscous liquid.

Symbols (Not given in the text)

Δ P = Pressure drop
L = Length
D = Pipe or vessel diameter
K = Consistency Index
n = Flow Index
k_s = Agitation Stearate constant.

Concluding Remarks

Process Development, Process Engineering and Design of Equipment are fast developing into a requirement of Chemical Engineering practice in this country. Thus in our attempt to update the "Physical Technology" background we are introducing such topics. The information given here is by no means exhaustive. However, it is a "refresher" to the practising engineers, and also a reminder to many for not getting into design without proper information. It is hoped that this purpose would be served. □