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STUDY 4

Technical project of a new purification solution of water infested with oil
(First part)

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ABSTRACT

This first part of the 4TH study presents: Calculations for sizing the main elements of the water treatment plant and Basic Theories to develop the technical project of a new purification solution of water infested with oil.

4.1 Calculations for sizing the main elements of the water treatment plant

In this phase of the project is proposed based on theoretical relationships, to dimension the main elements that goes into the system. Also in this phase will follow that by the CFD (Computational Fluid Dynamics) simulation to determine function of the characteristics of pollutants the optimal operation program of the installation.

4.2 Basic Theories

As with every specialist subject, it is easy for the centrifuge engineer to assume that his basic theories are universally understood. To ensure that this work is comprehensible to the widest possible readership, a few basic concepts will be covered. These will include g-force, differential speed, and mass balances across the decanter.

4.2.1 Acceleration force

The centrifugal acceleration force, commonly known as g-force, is the basic motive force for separating the solids from the liquid in any sedimenting centrifuge. Thus, in a handbook about the decanter no apologies are needed when covering g-force as the first basic concept.

Consider Figure 4.2. A particle, of mass m, rotates at a tangential velocity, v_{ct} and angular velocity, u;, in a circle of radius, *r.* After a time, t, the particle has moved to a point on the circle radius, r, which subtends an angle, 7, where 7 = u;t, from its position at time t=0, the extreme right of the horizontal diameter of the circle.

At time t, the horizontal distance of the particle from the centre of the circle is s, and at time *t=0* it was *r:*

$$s = r\cos(\gamma) \qquad (4.2)$$

$$s = r\cos(\omega t) \qquad (4.2')$$

The horizontal acceleration of the particle towards the centre is the second differential of s:

$$\frac{d^2 s}{dt^2} = \frac{d\{-\omega r \cdot \sin(\omega t)\}}{dt} \qquad (4.3)$$

$$= -\omega^2 \cdot r \cdot \cos(\omega t) \qquad (4.4)$$

At time t=0:

$$\frac{d^2S}{dt^2} = -\omega^2 \cdot r \cdot \cos(0) \qquad (4.5)$$

$$= -\omega^2 \cdot r \qquad (4.6)$$

Thus, anything moving in a circle of radius r, at an angular velocity u;, will experience an acceleration towards the centre of the circle of u;^{2} r.

In the centrifuge, it is the liquid that moves round in a circle, and the particles in suspension are free to move relative to the liquid. Thus, relative to the liquid, the suspended particles experience an acceleration, u;^{2}r, radially outwards.

Thus, the gravitational force, F, on a particle of mass m, is the product of its mass and acceleration, where:

$$F = m \cdot r \cdot \omega^2 \qquad (4.7)$$

In centrifuge parlance the term "g" (or g-level) is often used. This is the number of times the acceleration in the centrifuge is greater than that due to gravity alone. To differentiate between "g", which is dimensionless, and the acceleration due to gravity, having dimensions of LT~^{2}, the ratio of the two accelerations "g" will be denoted here by #$_c$, and that due to gravity simply by *g.* Thus:

$$g = \frac{\omega^2 \cdot r}{g} \qquad (4.8)$$

Note that the g-level within a centrifuge will thus vary, proportional to the radius, throughout the depth of the liquid, and is proportional to rotational bowl speed squared. (To calculate g_c using a simple expression, use rpm^2 x diameter in mm/1 . 789 x 10^6; for example, rotating at 3000 rpm at 450 mm diameter, #$_c$ would be 3^{2}x450/l. 789=2264.

4.2.2 Differential

The difference in rotational speed between the bowl and the conveyor is commonly referred to as the conveyor differential speed, N. Conveyor differential speed is calculated from a knowledge of the rotational bowl speed, S, the gearbox pinion speed, S_p, and the gearbox ratio, jR_{GB}:

$$N = \frac{(S - S_p)}{R_{GB}} \qquad (4.9)$$

When an epicyclic gearbox is used, the conveyor rotates at a speed less than the bowl speed, while with a Cyclo gearbox the conveyor rotates at a speed above the bowl speed. This fact can have an effect on process performance with short bowls, if the feed is not fully accelerated to bowl speed on entry. With the conveyor rotating faster than the bowl, the liquor has to get up to speed to find its way to the liquor discharge hub. With conveyors rotating slower than the bowl, the liquor could wind its way around the helix of the conveyor to the centrate discharge ports, without ever getting up to bowl speed. Note, to effect scrolling in the required direction, the flight helix on a conveyor using an epicyclic gearbox would be left-hand, and with a Cyclo gearbox it would be right-hand (assuming conventional equipment and operation).

With modern technology, the speeds of the bowl and gearbox pinion can be continuously measured with tachometers or proximity probes, and their signals fed to a simple PLC to work out, and even control, the differential. The PLC would need the gearbox ratio programmed in to execute this duty.

4.2.3 Conveyor torque

Conveyor torque, T, is a measure of the force exerted by the conveyor in moving the separated solids through the bowl, up the beach and out of the decanter. It equals the pinion torque, T_p, times the gearbox ratio:

$$T = R_{GB} \times T_p \qquad (4.10)$$

It is not easy to measure conveyor torque directly, whereas pinion torque can usually be obtained from instrumentation on the pinion braking system. This signal is often given to a PLC for control purposes. Conveyor torque is a vital measure in the control of modern decanter systems.

4.2.4 Process performance calculations

Consider the two-phase decanter separation system in Figure 4.2. Input is the feed at a rate of Q_f, of density p_f, with a solids fraction x_f, and an additive, often a flocculent, of density p_p, at a rate of Q_p, with a solids fraction x_p. There are two products, cake at a volumetric flow rate Q_s, at density p_s, and solids fraction x_s, and a centrate at flow rate Q_b at density p_b with a solids fraction of *x*. From measurements of some of the eight process parameters mentioned, it is required to assess the performance of the decanter. It is normal to monitor the feed rate, Q_f, and the additive rate, Q_p, with flow meters.

Periodically, gravimetric analyses are conducted on samples of feed, cake, centrate and, if necessary, the additive. Performance is judged by how high is the solids recovery, R, and how low is the flocculent dose, P_D, when this is used. Recovery is the percentage of solids in the feed that reports to the cake discharge. Flocculent dose, sometime referred to as polymer dose, is the amount of dry polymer used per unit dry solids in the feed, usually expressed as kg/t db (kilograms per tone dry basis).

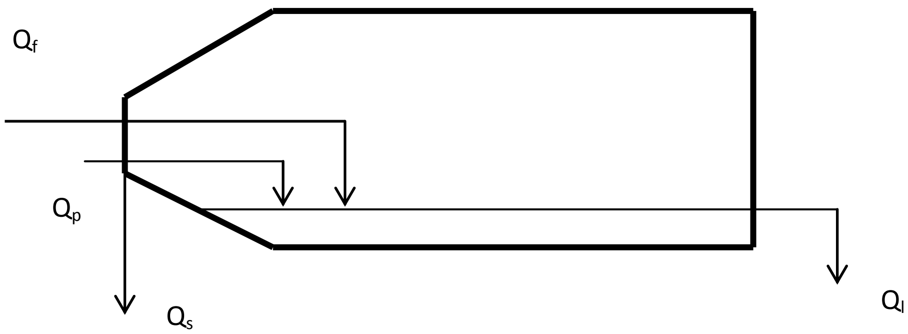

Figure 4.2. Decanter Mass Balance

As an intermediate parameter in the calculations, it is necessary to calculate the centrate rate, *Q*\, by conducting a total and a solids mass balance across the decanter. Total mass balance:

$$Q_f \cdot \rho_f + Q_p \cdot \rho_p = Q_s \cdot \rho_s + Q_l \cdot \rho_l \qquad (4.11)$$

Solids mass balance:

$$Q_f \cdot \rho_f \cdot x_f + Q_p \cdot \rho_p \cdot x_p = Q_s \cdot \rho_s \cdot x_s + Q_l \cdot \rho_l \cdot x_l \qquad (4.12)$$

Eliminating $Q_s p_s$ from equations (4.21) and (4.22):

$$Q_l \cdot \rho_l = Q_f \cdot \rho_f \cdot \frac{(x_s - x_f)}{(x_s - x_l)} + Q_p \cdot \rho_p \cdot \frac{(x_s - x_p)}{(x_s - x_l)} \qquad (4.13)$$

Recovery of solids is calculated by subtracting the percentage loss of solids in the centrate from 100. Thus:

$$R = 100 \cdot \left(1 - \frac{Q_l \cdot \rho_l \cdot x_l}{Q_f \cdot x_f \cdot \rho_f}\right) \qquad (4.14)$$

Polymer dosage is given by:

$$P_D = \frac{Q_p \cdot x_p \cdot \rho_p}{Q_f \cdot x_f \cdot \rho_f} \qquad (4.15)$$

Polymer dose levels are frequently expressed in kg/t db, with the dry basis measure applying to both solids rate and polymer rate.

During these calculations, one must take care with the units used. Volumetric flow rates are invariably measured and the density terms are often ignored as they are usually close to unity. However, the density terms must be used when density values are significantly above unity. The gravimetric analyses of the samples should all be total solids (i.e. samples are evaporated to d the centrate generally are. However, centrates are generally analyzed in terms of suspended solids. Any dissolved solids in the centrate cannot be considered a measure of a decanter's inefficiency, as it is suspended solids which it separates, and any dissolved solids attached to the cake by virtue of its moisture content represents a bonus. Dissolved solids in the centrate are usually low and can be ignored, but when they are not, then they should be included in the equations when calculating centrate rate.

4.2.5 Particle Size Distribution

Very few process slurries contain particles of uniform size. A large proportion of slurries, processed by decanters, contain solids which have a particle size distribution which conforms closely to a logarithmic probability distribution. The logarithmic probability equation was derived by Hatch and Choate in 1929:

$$\frac{dz}{d\{\ln(d)\}} = \frac{Z}{\sqrt{2\pi}\ln(\sigma_g)} \cdot \exp\left[-\frac{\{\ln(d)-\ln(d_g)\}^2}{2\{\ln(\sigma_g)\}^2}\right] \qquad (4.16)$$

where Z is the total number of particles; *d* is the particle diameter; cr_g is the geometric standard deviation; d_g is the geometric mean diameter; and *z* is the number of particles less than diameter *d.*

Integrating this equation gives the formula for a cumulative number distribution:

$$C_n = \frac{1}{2} + \frac{1}{2} erf\left[\frac{\ln\left(d/d_g\right)}{\sqrt{2}\ln(\sigma_g)}\right] \qquad (4.17)$$

where C_n is the cumulative fraction of the number of particles below size *d*\ and erf is a tabulated integral from -1 to +1.

It can be shown, by using equation (4.26), that the equation for the cumulative weight distribution is similar:

$$C_w = \frac{1}{2} + \frac{1}{2} erf\left[\frac{\ln\left(\frac{d}{d_g}\right)}{\sqrt{2}\ln(\sigma_g)} - \frac{3\cdot\ln(\sigma_g)}{\sqrt{2}}\right] \qquad (4.18)$$

where C_w is the cumulative weight or volume of particles below size *d.* In verting and simplifying equation (4.28):

$$\ln(d) = a_1 + a_2 erf^{-1}(2C_w - 1) \qquad (4.19)$$

where a_1 and a_2 are constants, functions of d_g and <r_g.

Since Hatch and Choate first published their equation, special graph paper has been developed and printed whereby plotting the cumulative percent of particles by number or weight, oversize or undersize, against particle size, results in a straight line. The mathematics of the distribution are such that one can readily transfer between weight and number distributions, and even area and diameter distributions.

The diameter at the 50% line on the graph gives the geometric mean diameter for the type of distribution plotted, be it number, weight or whatever. The geometric standard deviation, which is the same for all types of distribution, is given by the size for which 84.23% of the number, weight, etc., is smaller, divided by the geometric mean size:

$$\sigma_g = \frac{d_{84.13}}{d_{50}} = \frac{d_{50}}{d_{15.87}} \qquad (4.20)$$

The relationship between the various means is given by:

$$d_{gw} = d_g e^{3(\ln\sigma_g)^2} \qquad (4.21)$$

$$d_{gs} = d_g e^{2(\ln\sigma_g)^2} \qquad (4.22)$$

$$d_{gl} = d_g e^{(\ln\sigma_g)^2} \qquad (4.23)$$

where d_{gw} is the geometric mean for a weight distribution; d_{gs} is the geometric mean for an area distribution; d_{gl} is the geometric mean for a diameter distribution; and d_g is the geometric mean for a number distribution.

Figure 4.3 shows a typical distribution plotted as a number, area, and weight distribution on the specially scaled graph paper. From a graph such as this, the two basic parameters, cr_g and d_g, can readily be obtained and from these more pertinent information can also be obtained.

For example, the total surface area, A_T, of the solids in the slurry can be calculated:

$$A_T = \frac{6}{d_{gw}} \cdot e^{\frac{1}{2}\{\ln(\sigma_g)\}^2} \qquad (4.24)$$

This parameter is useful for paints and pigments, giving the covering power of the solids. It is also useful in assessing relative flocculent demand, as this is proportional to the surface area of the particles.

In general, the decanter removes the coarsest or densest particles from the slurry, leaving only the finest or least dense in the centrate. Knowing the required percentage solids removal from the slurry, the recovery, one can read the desired cut point from the distribution graph. This then gives an appreciation of the feasibility of the desired separation. Experience will tell whether the decanter would be capable of achieving the required cut point. For instance, a cut point of 2-5 |im would be feasible on a decanter with most slurries, but, say, less than 0.1 ^im would probably be impossible, however high the density of the particles.

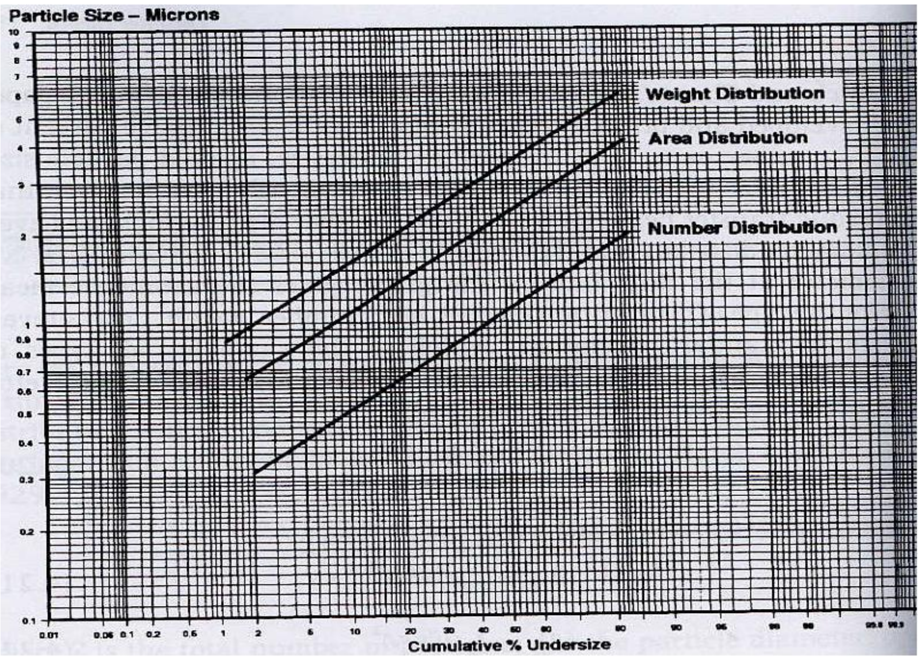

Figure 4.3. Cumulative weight, area and number log probability distributions

The cut point size is the smallest particle size that has to be settled in the decanter. Technically 50% of particles of that size settle and 50% are lost in the centrate; above that size the separation efficiency increases and below it vice versa. In consequence, the size distributions in both the cake and the centrate will also exhibit logarithmic probability distributions.

Figure 4.4 depicts examples of weight frequency distributions for feed, cake and centrate. Note that the centrate and cake lines intersect at the cut point size, and at a frequency level half that of the feed at that size. Hence, there is a 50% split between cake and centrate at the cut point.

These frequency distributions are plotted as cumulative weight distributions in Figure 4. 5.

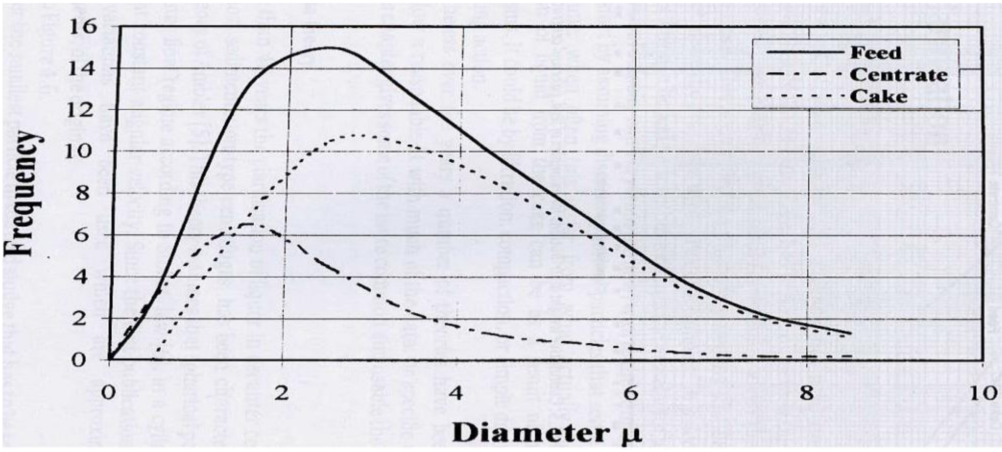

Fig. 4.4 Frequency distributions by weight for examples of feed, cake and centrate

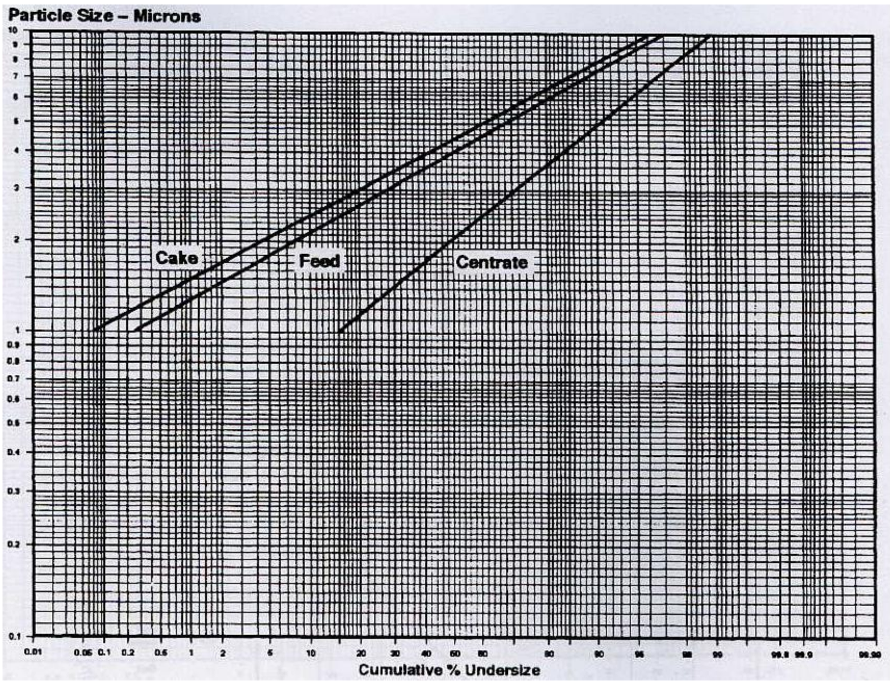

Fig 4.5 Cumulative weight distributions for examples of feed, cake and centrate

4.3 <u>Clarification</u>

The separation of solid particles and agglomerates from the suspending liquors, within a decanter centrifuge, has invited numerous theories. Few of these theories have provided exact results, which have given the opportunity for many more. This is no reflection on those providing the theories, as the system, let alone the processes used, are quite complex. In a decanter, liquid and solids flow in a helical path within a cylindrical vessel with a conical end. The rotational velocity can vary from the bowl wall to the pond surface. Many theories start by assuming discrete spherical particles that settle in a laminar flow regime, when often this is far from what actually happens. The expression of liquid from the cake can be as a result of one or more mechanisms. It could be by filtration, compaction, or simple drainage against the scrolling action.

Nevertheless over the years a number of theories have been proposed, which allow a reasonable fit with much of the data, or specific categories of data. This chapter gives some of the more common and usable theories.

4.3.1 Sigma theory

For more than 40 years the clarification of liquor in decanter centrifuges, in fact in most sedimentation-type centrifuges, has been characterized by the Sigma theory of Ambler. This theory assumes that spherical particles settle in a laminar flow regime according to Stokes' law, in a cylindrical bowl rotating at constant angular velocity. Since the first publication by Ambler, several variations have been used which are approximations, or developments, of the original.

Refer to Figure 4.6

Consider the smallest particle in the feed sludge that has to be separated, the cut point size, d_c. This particle has a density, p_s, and settles in a liquor of density, PL, and viscosity, TJ_L. The feed slurry enters the decanter at a rate of Q_f, at a pond radius, r_{lf} at point X at time *t=0.* By the time the particle traverses the clarifying length of the centrifuge, L, in time $t=£_e$, the particle must settle to a radius r_2, at point *Y,* the bowl internal radius, if it is to be collected by the conveyor. The centrifuge rotates at a constant angular velocity, *u>.* It is assumed that the fluid in the bowl also rotates uniformly at angular velocity *u,* and travels along the bowl in plug flow. It is further assumed that the particle being considered is homogeneous and spherical, settling in a laminar flow regime.

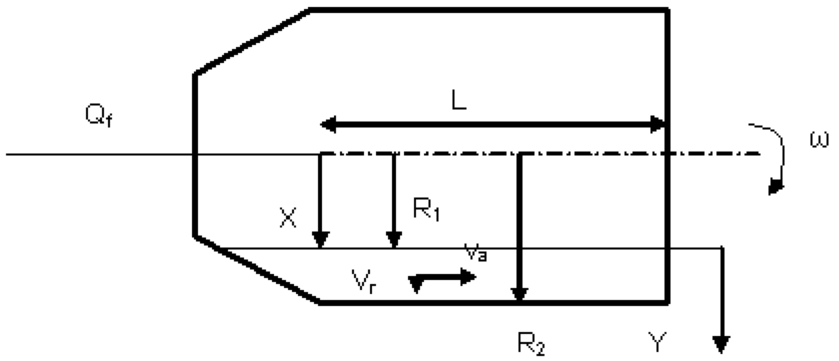

Figure 4.6. A particle settling in a decanter

At time *t=t,* the particle has a radial velocity v_r and a constant axial velocity of v_a. It is also assumed that the particle travels a negligible distance from the pond surface before it reaches its terminal velocity.

The axial velocity is given by:

$$v_a = \frac{Q_f}{\pi\left(r_2^2 - r_1^2\right)} = \frac{L}{t_e} \tag{4.25}$$

Thus:

$$t_e = \frac{\pi L}{Q_f}(r_2^2 - r_1^2) \tag{4.26}$$

The radial velocity at any time *t* is given by:

$$v_r = v_s \frac{\omega^2 r}{g} \tag{4.27}$$

where v_s is the Stokes settling velocity, given by:

$$v_s = \frac{d_c^2\left(\rho_s - \rho_L\right)g}{18\eta} \tag{4.28}$$

where d_c is the particle size at the cut point; p_s is the density of the particle; p_L is the density of the liquid; and η is the viscosity of the liquid. Now:

$$v_s = \frac{dr}{dt} \tag{4.29}$$

Substituting equation (4.29) into (4.27) and integrating between the limits of r=ri to r_2 and *t=0* to t_e:

$$\ln\frac{r_2}{r_1} = \frac{\omega^2 v_s}{g} \cdot t_e \tag{4.30}$$

Eliminating t_e from equations (4.26) and (4.30):

$$Q_f = \frac{\pi L \omega^2}{g} \cdot \frac{\left(r_2^2 - r_1^2\right)}{\ln\left(\frac{r_2}{r_1}\right)} \cdot v_s \tag{4.31}$$

The terms on the right-hand side of equation (4.31) consist of v_s, which is solely a function of the process material, and the remainder of the terms, that are solely functions of the centrifuge. These latter terms are collectively known as Sigma, E, the clarification capacity of the centrifuge:

$$\Sigma = \frac{\pi L \omega^2}{g} \cdot \frac{\left(r_2^2 - r_1^2\right)}{\ln\left(\frac{r_2}{r_1}\right)} \tag{4.32}$$

Sigma has units of area, which is consistent with non-centrifugal clarifying equipment.

From equations (4.31) and (4.32):

$$Q_f = v_s \cdot \Sigma \tag{4.33}$$

Equation (4.32) is the equation for Sigma preferred today, and is particularly recommended when deep pond decanters ($ri/r_2<0.65$) are used. When Ambler first derived his formula, only shallow ponds ($ri/r_2>0.75$) were used, and he used different starting assumptions for his derivation.

With a shallow pond it is assumed that the incoming feed distributes itself evenly throughout the depth of the pond, in the annular plane of the point of entry. The theory then develops the same equations in the same way, assuming that half of the particles of the smallest size that have to be separated will be removed. This is consistent with the definition of cut point.

The last particle of the half of the smallest particles to be settled will start at a radius r_x at the feed point, at which half of all particles will start inward and half outward of this point in this plane.

So:

$$r_2^2 - r_x^2 = r_x^2 - r_1^2 \tag{4.34}$$

When ce:

$$r_x^2 = \frac{r_2^2 - r_1^2}{2} \tag{4.35}$$

Now substituting equation (4.29) into equation (4.27) again, but this time integrating between the limits of $r=r_1$ and $r=r_x$ and t=0 to $t=t_s$:

$$\int_{r_1}^{r_x} \frac{dr}{r} = v_s \frac{\omega^2}{g} \int_0^{t_e} dt \qquad (4.36)$$

Solving equation (4.36), eliminating r_x using equation (4.35) and eliminating t_e using equation (4.26):

$$Q_f = 2\pi L \frac{\omega^2}{g} \frac{r_2^2 - r_1^2}{\ln\{2r_2^2/(r_1^2 + r_2^2)\}} v_s \qquad (4.37)$$

$$Q_f = 2v_s \Sigma \qquad (4.38)$$

where ∑ this time is the true Ambler Sigma given by:

$$\Sigma = \pi L \frac{\omega^2 L}{g} \left(\frac{3}{4} r_2^2 + \frac{1}{4} r_1^2\right) \qquad (4.39)$$

Compare ∑ of equation (4.39) with ∑ of equation (4.32). Note the extra numeral 2 in equation (4.38) compared with equation (4.33). This is to be expected if all the particles that have to be separated have the advantage of starting at half pond depth!

It will also be seen in the literature that approximations are sometimes made for the logarithmic term in equation (4.32) to give:

$$\Sigma = 2\pi L \frac{\omega^2}{g} \left(\frac{3}{4} r_2^2 + \frac{1}{4} r_1^2\right) \qquad (4.40)$$

With shallow ponds it is sometimes considered that the g-force is constant, in which case equation (4.27) would be rewritten:

$$v_r = v_s \frac{\omega^2 (r_2 - r_1)}{2g} \qquad (4.41)$$

Substituting equation (4.29) into (4.41) and integrating:

$$(r_2 - r_1) = v_s \frac{\omega^2 (r_2 + r_1)}{2g} t_e \qquad (4.42)$$

Eliminating t_e from equations (4.42) and (4.26) and rearranging gives:

$$Q_f = \frac{\pi L \omega^2 (r_2 + r_1}{2g} \cdot \frac{(r_2^2 - r_1^2)}{(r_2 - r_1)} v_s \qquad (4.43)$$

$$= \pi L g_c^{'} (r_2 - r_1) v_s \qquad (4.44)$$

$$= \pi L g_c^{'} D_{AV \cdot v_s} \qquad (4.45)$$

where $g_c^{'}$ is the mean g-level in the pond; and D_{AV} is the average pond diameter. Alternatively equation (4.43) may be written:

$$Q_f = \frac{V}{\Delta r} g_c^{'} \cdot v_s \qquad (4.46)$$

where Δr is the pond depth; and *V* is the pond volume.

For this derivation of clarification capacity, it is readily deduced that:

$$\Sigma = \pi L \frac{\omega^2}{g} \frac{r_2^2 - r_1^2}{r_2 - r_1} \qquad (4.47)$$

In the graph of Figure 4.7 the various formulae for Sigma developed so far (equations (4.32), (4.39), (4.40) and (4.47)) are compared for various ratios of r_1/r_2.. The common factor $\pi L \omega^2 / g$ is removed and r_2 is taken as unity, for the graphical comparison.

By means of Figure 4.7, a number of observations may be made. The expansion of the logarithmic term to give an easier formula for Ambler's Sigma is a very acceptable approximation. The even simpler formula last developed above is also acceptable for shallow ponds (radii ratio greater than 0.75). However, there is a significant difference for the formula used for deep ponds. Notice that with zero pond radius, the shallow pond versions of Sigma have finite Sigma values while the deep pond version has a zero value. This is because a particle starting at the centre line will experience no *g* to initiate its fall, while those which by definition start half way into the pond, or are subjected to a mean *g* throughout the pond will always have a finite settling rate. However for practical designs the radius ratio will always be appreciably over zero, generally in the range 0.4 - 0.8.

It will be seen from the various Sigma formulae that increasing the length of the bowl increases Sigma pro rata. Thus, in this respect Sigma is additive. Some like to include the Sigma value of the beach in their formula, especially when feeding on the beach. For this, using equations (4.46) and (4.40):

$$\Sigma = \frac{2\pi\omega^2 g}{g} \left\{ L_c \left(\frac{3}{4} r_2^2 + \frac{1}{4} r_1^2 \right) + \frac{L_k}{8} \left(r_2^2 + 3 r_1 r_2 + 4 r_1^2 \right) \right\} \qquad (4.48)$$

where L_k is the wetted beach axial length; and L_c is the cylindrical length of the bowl.

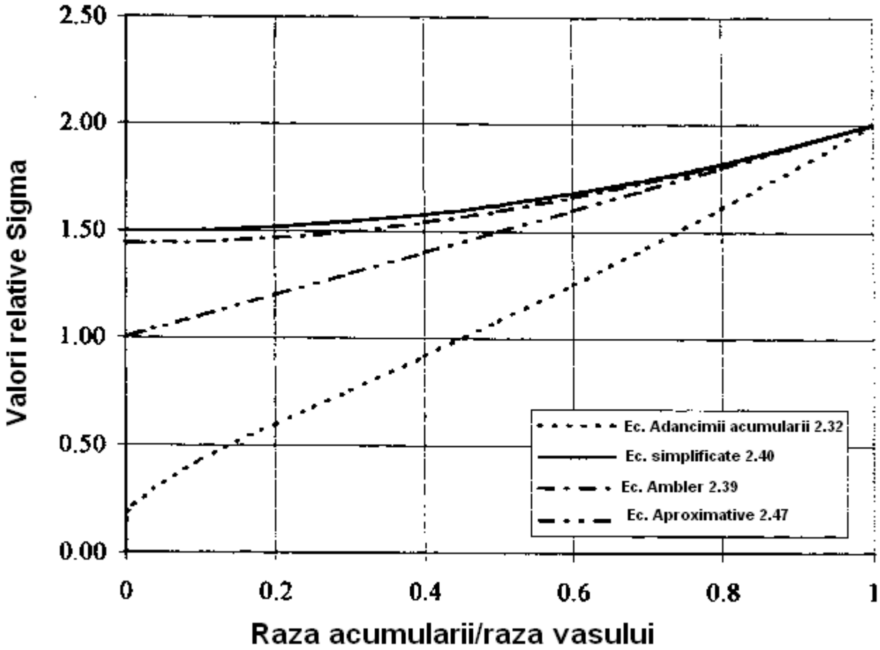

Figure 4.7. Graph comparing the various formulae for Sigma at various pond depths

Considering the assumptions used in the derivation, and the approximations used, one could question whether the use of the simpler equation (4.47) would not suffice, for use with shallower pond machines at least. It is the ratio of Sigma values which is used when scaling from one decanter size to another. Using equation (4.48), the ratio will be little affected as the extra term will increase by approximately the same ratio with geometrically similar machines.

Another expression, in place of Sigma, uses an empirical formula taking a nominal bowl radius, the f bowl radius, *n* (which equals three quarters of r_2).

This expression is termed the "area equivalent", $Ae_{3/4}$ and is determined by:

$$Ae_{\frac{3}{4}} = \frac{2\pi\omega^2}{g}\left(\frac{3}{4}r_2\right)^2\left(L_c + \frac{r_2}{4}\cot\alpha\right) \quad (4.49)$$

where *a* is the beach angle (half included).

More often the abbreviated form is used, which ignores beach volume and uses clarifying length:

$$Ae_{\frac{3}{4}} = \frac{2\pi\omega^2}{g}\left(\frac{3}{4}r_2\right)^2 L \quad (4.50)$$

$Ae_{3/4}$ is then used as a scale-up factor, in place of Sigma. Its use is simply a matter of choice and habit.

All the formulae indicate that a better clarification capacity is achieved at the shallowest pond depth, whereas in practice it is generally the opposite. Therefore, the simple formula is generally considered sufficient for practical purposes. However, when scaling from one machine to another, it is imperative that the same formula is used for both machines. It is also recommended that one should not normally scale between machines of dissimilar geometry.

4.3.2 Using sigma

It is unusual to use any of these formulae to compute the capacity of a single machine. Their most effective and reliable use is in scaling data from one geometrically similar machine to another and assessing relative performances.

Eliminating v_s from equations (4.28) and (4.33):

$$\frac{Q_f}{\Sigma} = \frac{d_c^2(\rho_s - \rho_L)g}{18\eta} \tag{4.51}$$

or

$$\frac{Q_f}{\Sigma} \propto d_c^2 \tag{4.52}$$

Taking logarithms of both sides of equation (4.52):

$$\ln\left(\frac{Q_f}{\Sigma}\right) \propto \ln(d_c) \tag{4.53}$$

It is known from equation (4.27) that the percentage over or under size is a logarithmic probability function of particle diameter. Thus, combining equations (4.29) and (4.53), a logarithmic probability relationship between Q_f/Σ and solids recovery is obtained:

$$\ln\left(\frac{Q_f}{\Sigma}\right) \propto a_1 + a_2 erf^{-1}(2c_w - 1) \tag{4.54}$$

Plotting Q_f/Σ against solids recovery will thus give a good correlation [9]. Plotting on logarithmic probability paper will produce a straight line. The same straight line is obtained for data from different decanters, preferably of the same geometry. However, it must be noted that this only applies to process materials with solids exhibiting a skew Gaussian (logarithmic probability) distribution.

When scaling from one machine to another:

$$\frac{Q_{f2}}{Q_{f1}} = \frac{\Sigma_2}{\Sigma_1} \quad (4.55)$$

where the subscripts 1 and 2 refer to centrifuges 1 and 2, respectively.

When machines of different geometry are used then one needs to take into account the relevant efficiency, £, of each design, when:

$$\frac{Q_{f2}}{Q_{f1}} = \frac{\xi_2}{\xi_1}\frac{\Sigma_2}{\Sigma_1} \quad (4.56)$$

4.3.3 Sigma enhancement

The use of conical discs, or angled vanes, on the conveyor will theoretically enhance the Sigma value, the clarification capacity, of the centrifuge. To estimate the Sigma value of a stack of conical discs, the formula used for disc stack centrifuges may be employed [11]:

$$\Sigma_d = \frac{2\pi n_D}{3} \cdot \frac{\omega^2}{g} \cdot \left(r_3^3 - r_1^3\right) \cdot \cot\theta \quad (4.57)$$

where $\sum_D$ is the Sigma value for the disc stack; n_D is the number of discs; r_3 is the outside radius of the discs; and *9* is the half included angle of the discs.

The total Sigma value for the centrifuge is obtained by adding $\sum_D$ to the Sigma value calculated for the conveyor section between the feed zone and the discs.

There is little published on the effect of longitudinal angled vanes, but the equation is derived in a similar fashion to that used for the disc stack centrifuge:

$$\Sigma_V = \frac{n_V L_V \omega^2}{2g \cdot \cot\psi}\left(r_3^2 - r_1^2\right) + \frac{\pi L_V \omega^2}{g} \cdot r_3^2 \quad (4.58)$$

where $\sum_V$ is the Sigma value for the vanes; L_v is the length of the vanes; n_v is the number of vanes; and ψ is the angle between the vane and a radius.

If the vanes do not extend the full length between the feed zone and the centrate discharge, then the Sigma of the plain section needs to be added to $\sum_V$ to obtain the total Sigma value for the centrifuge.

Caution is needed in using these extended Sigma values, particularly for the angled vanes. This is because flow through the vanes or discs can channel, to take the easiest path. This will reduce the effectiveness of the devices. Good designs, therefore, will endeavor to ensure even distribution of the flow across the vane and disc openings. Even then as liquor flows from the outer edge of vanes or discs, towards the centrifuge axis for discharge at the weir lips, considerable changes in kinetic energy occur. This can cause very complex flow patterns, turbulence and Coriolis effects.

4.3.4 Flocculent requirement

Chapter 5 is devoted to flocculants. However, it is appropriate at this juncture to mention the need for polymers in some process applications, particularly effluent applications which are a large market for the decanter. In these applications, without a polymer flocculent, it would not be possible to employ a decanter. It is clear from equation (4.28), Stokes' law, that the settling velocity of a particle, v_s, is proportional to the square of its diameter. Thus doubling the particle diameter will increase v_s by a factor of four. This results in greater separation efficiency. The objective of flocculants is to change the electrochemical forces on the surface of the particles, so as to bind them together such that they act as one large particle. Once flocculated, these particles must be handled carefully so as not to break them up mechanically. This is especially true when processing them in a decanter.

In most applications, the amount of polymer used is just sufficient to flocculate a sample of the feed. The amount necessary, as assessed in the laboratory, is generally the amount used in practice on the centrifuge, plus or minus a small fraction. However, recently there has been considerable development in decanters and their use in obtaining extra-dry cakes from compressible sludge, particularly effluents. In these instances, the consumption of polymer has increased considerably.

The amount of flocculent needed increases as the extent of dryness required in the cake increases, and it increases exponentially. The amount of flocculent required also increases with the feed rate to the centrifuge. In practice, on a "dry solids" application, the polymer used will be two to three times that which would be used on a standard application with the same process material.

There has not been a theoretical formula proposed to quantify polymer demand. However, the available data suggest a format similar to equation (4.59):

$$P_D = k_1 + k_2 \cdot e^{(x_s - k_3)} \qquad (4.59)$$

where k_1, k_2, and k_3, are constants.

Practical data can be very erratic, as it is easy to overdose when striving for extra dryness. When assessing the minimum polymer requirement, it is necessary carefully to adjust all operating parameters, to ensure performance is at the limit, without contingency levels added.

4.4 Classification

Classification, the fractionation or separation of particles by size, could be considered as merely inefficient clarification. The cut, or desired classification, is adjusted by altering the centrifuge's efficiency. This is most easily done by altering the feed rate or bowl speed. However, adjustment of pond depth or differential may, in certain circumstances, be used.

In a thick suspension hindered settling occurs, when there is a tendency for the larger particles, which should settle, to get held up by the dense concentration of the smaller particles. In these circumstances higher differentials could be used, to agitate the suspension and so release the heavier particles. The disadvantage of this is that the cake or "heavy fraction" tends to be wetter, as a result of the higher differential, and thus entrains larger quantities of the smaller particles. To correct this, a shallow pond is selected to allow release of liquid containing the smaller particles on the dry beach.

In some classification applications, the required cut point is very sharp and the rheology of both separated phases is such that they remain quite fluid. In this type of application the pond used would be relatively deep, and separation would be akin to a liquid/liquid separation, using a hydraulic balance under some form of baffle.

Very occasionally there will be found a classification application where it is required to separate two distinctly different particles, such as in the refining of minerals. In these cases the two different substances to be separated may have markedly different densities. This is particularly acceptable and quite advantageous when the denser material comprises the larger-sized particles. However, if this is not so, one must consider a combination of density and particle size for the cut point of each of the two substances, in relation to Stokes' law. One could visualize the situation of a large, low-density particle settling faster than a high-density, small particle. Thus for such a process to be feasible:

$$d_{ch}^2(\rho_{sh} - \rho_1) > d_{cl}^2(\rho_{sl} - \rho_l) \qquad (4.60)$$

where d_{ch} is the required cut point size of the heavy fraction; $d_c\backslash$ is the required cut point size of the light fraction; ρ_{sh} is the density of the heavy solids; and ρ_{sl} is the density of the light solids.

Each of the two solid constituents will have their own size distribution from which a cut point size can be chosen to give the desired purity of product and yield.

Poor efficiencies can occur in some classification applications, due to natural agglomeration of particles. In these applications, the use of dispersants is quite common. Dispersants have the opposite effect to flocculants, and can be equally powerful.

4.5 Three-Phase Separation

Decanter three-phase operation involves the separation of two immiscible liquids from a solid. The two immiscible liquids are generally oil and water. This could be a waste oil application or the separation of a vegetable oil, such as palm or olive oil.

To put the decanter into operation two weir heights or equivalent have to be set relative to the solids discharge level, as illustrated in Figure 4.8.

Firstly the weir height, governed by radius r_{lf} is set to fix the extent of the dry beach required before solids are discharged at radius r_s.

The radius r_h has then to be set, to create a hydraulic balance between the two liquid phases, to maintain the equilibrium line at r_e, where required.

The pressure at any radius, r, in a rotating centrifuge is given by P_r where:

$$P_r = \frac{\rho\omega^2}{2g}\left(r^2 - r_1^2\right) \qquad (4.61)$$

Thus, in the three-phase centrifuge, the pressure at the equilibrium line is P_e where:

$$P_e = \frac{\rho_l\omega^2}{2g}\left(r_e^2 - r_1^2\right) = \frac{\rho_h\omega^2}{2g}\left(r_e^2 - r_h^2\right) \qquad (4.62)$$

where ρ_l is the density of the light phase; and ρ_h is the density of the heavy phase.

The choice of e-line position depends upon a number of factors. The volume of each phase in the bowl could be chosen in proportion to the volumes of each in the feed. Then approximately the value of Q/∑ for each phase would be the same. However, if the separation of one phase from another is relatively more difficult than vice versa, then extra volume could be given to one phase in the bowl to improve its clarification efficiency.

Alternatively, the purity of one phase may be more important than the other, and then bias would be given to the more important phase.

Nevertheless, care has to be taken in setting the e-line in order not to allow breakthrough of one phase into the other.

When the flow rate of one or more of the phases is high, cresting over the weirs can move the e-line considerably, and adjustment of the weir heights will become necessary. Back-pressure from a centripetal pump or skimmer pipe will require recalculation of weir settings.

Working with three-phase separation will require revision of the formulae used for performance evaluation. Not only will there be interest in solids recovery and clarity (absence of solids), for both liquid phases, there will be an avid interest in what has happened to the oil. How free of water is the oil? What is the recovery of the oil? How much oil is left in the cake and the water phase?

In some of the three-phase applications, water is added to dislodge the oil from the solids. With water addition there are two input streams, feed and water, and three outlet streams, light liquid phase (oil), heavy liquid phase (water) and cake (solids). Each stream is analyzed for the three elements, oil, water and solids. Four of the five streams are monitored for flow rate. The cake rate would be difficult to measure accurately.

To analyze performance, a mass balance across the centrifuge is performed for the three separate elements and the total mass, after which the cake rate is eliminated. Formulae are then developed for pertinent recoveries and purities.

It is not necessary to develop these here as the pertinent formulae will depend on the application, and in any case the development is similar to that already shown for two-phase separation (see Section 4.2).

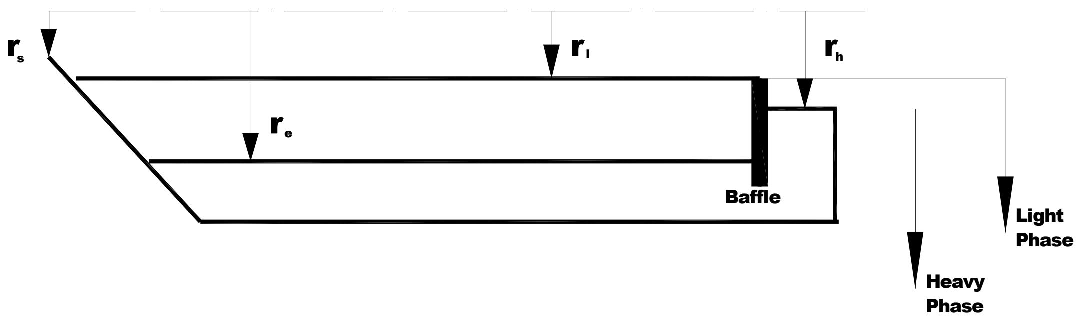

Figure 4.8. Hydraulic balance in three-phase separation.

4.6 Fluid Dynamics

By design the decanter handles very high throughputs relative to the small space it occupies. Moreover, the flow is not simply in one end, and straight through and out the other end. Flow can be under, over and around baffles; it can be a helical path around the conveyor flights, or axially through holes in the conveyor, or a combination of both.

The axial velocity of the feed into the decanter has to be converted to a rotational velocity in a very short time. This can cause considerable turbulence, and help is required outside the feed zone to keep the bowl contents up to speed, if not to get it fully to speed. The rotational speed of liquor at the pond surface can slip below that at the bowl wall.

To maintain flow down the bowl and over the weirs, a hydraulic head builds up with a crest.

In this section some of these phenomena will be examined more closely.

4.6.1 Reynolds number

The degree of turbulence in pipes and channels is characterized by the value of the Reynolds number. For a pipe:

$$R_e = \frac{\rho u D_p}{\eta} \qquad (4.63)$$

where D_p is the pipe diameter; and *u* is the velocity in the pipe. In the decanter with axial flow:

$$u = \frac{Q_l}{\pi\left(r_2^2 - r_1^2\right)} \qquad (4.64)$$

In a channel, the pipe diameter, D_p, is substituted by a hydraulic mean diameter, d_m:

$$d_m = \frac{4A}{p} \qquad (4.65)$$

where A is the cross sectional area of the channel; and p is the wetted perimeter of the channel.

Thus, for a non-circular pipe or channel:

$$Re = \frac{\rho u d_m}{\eta} \qquad (4.66)$$

For an annulus:

$$d_m = 2(r_2 - r_1) \qquad (4.67)$$

However the annulus of the pond in a decanter does not have its inner surface " wetted", and thus the hydraulic mean diameter becomes:

$$d_m = \frac{4\pi(r_2^2 - r_1^2)}{2\pi r_2} = 2\left(r_2 - \frac{r_1^2}{r_2}\right) \qquad (4.68)$$

If the values for d_m from equation (4.68) and velocity from equation (4.64) are introduced into the Reynolds number in equation (4.66), this would imply axial flow. For helical flow:

$$p = P + 2(r_2 - r_1) \qquad (4.70)$$

and:

$$A = P(r_2 - r_1) \qquad (4.71)$$

Thus:

$$d_m = \frac{4P(r_2 - r_1)}{P + 2(r_2 - r_1)} \qquad (4.72)$$

and:

$$u = \frac{Q_l}{\{P(r_2 - r_1)\}} \qquad (4.73)$$

Equations (4.72) and (4.73) can be substituted into equation (4.66) to find the Reynolds number for helical flow.

Once the value of the Reynolds number is known, for whatever type of flow is used, the level of turbulence can be assessed. With a Reynolds number below 2000 the flow would be laminar. It will be found that the flow in many, if not most, of all practical cases is in the turbulent regime.

In all decanters, with solids moving radially out and liquid moving radially in, acceleration and deceleration occur, respectively. Without any mechanical device to do this, viscous drag of the pond is the only means by which these actions can be accomplished. If the viscosity is low, considerable turbulence can occur, affecting cresting, interface location, stability, and sedimentation and re-entrainment of settled solids.

4.6.2 Moving layer

In a pond of an operating decanter centrifuge, there often tend to be two, distinct liquid layers. The upper or surface layer, the moving layer, moves rapidly and turbulently towards the discharge weirs. Under this moving layer, the pond is quiescent, allowing solids to settle under a laminar flow regime, and then to compact. This is a simplistic picture, as the shape of the conveyor and its movement adds to the complexity.

It is sometimes useful to estimate the depth of the moving layer to know when it is liable to disturb and re-entrain sedimented solids.

It will be appreciated that the thickness of the moving layer will depend upon whether the flow is axial or helical. Research has shown that for axial flow:

$$h_m \propto \frac{\sqrt{Q_l}}{\sqrt[3]{g_1}} \qquad (4.74)$$

where h_m is the thickness of the moving layer.

It will be seen that the formula is independent of path length, the clarifying length. It has also been found that moving layer thickness closely follows cresting height (see Section 4.21.3). Thus, the shape, size and number of weir plates used can affect the moving layer thickness. The moving layer thicknesses found in helical flow are greater than those calculated using equation (4. 12 6).

4.6.3 Cresting

The level difference between centrate and cake discharges can be quite critical when optimizing process performance. When the level difference is small, the degree and consistency of the centrate cresting can play an important part in the process performance optimizing.

The crest height, the pond surface level above the actual ^weir height, is a function of the centrate rate, the total weir width and centrifuge g-level, as well as physical constants of the liquor such as viscosity and density. Crest height is h_c, given by:

$$h_c = \frac{1}{\sqrt[3]{2r_1}}\left(\frac{Q_l}{c_0\omega B}\right)^{\frac{2}{3}} \quad (4.75)$$

where c_0 is a constant and generally approximately 0.415; and B is the total length of weirs.

This equation is derived from the Francis formula.

Experimental data show that, due to the interrupted nature of the discharge weir, the calculated value of crest height needs to be increased by 35% for axial flow and 90% for helical flow. With a 360° internal weir, B would be the full circumference, and thus would cause the least cresting.

4.6.4 Feed zone acceleration

Feed zones are designed to accept the maximum possible feed rate, and bring it up to bowl speed with the minimum of splashing and rejection. Bringing the feed up to the angular velocity of the bowl is not necessarily enough. As the process material flows out of the feed zone to the pond, it has a constant linear velocity fixed by its angular velocity at its point of exit from the feed zone. To maintain its angular velocity extra linear velocity is required as the radius increases.

The power required to bring the feed material up to bowl speed at the pond surface is P_P, where:

$$P_p = Q_f \rho_f \omega^2 r_1^2 \quad (4.76)$$

Power available in the feed stream at the pond surface is P_A, where:

$$P_A = \frac{1}{2} Q_f \rho_f \omega^2 r_1^2 \quad (4.77)$$

The power lost on entry is thus the difference between equations (4.72) and (4.73), and this is dissipated in heat and turbulence on entry. Thus to minimize turbulence and power loss, it is necessary to design the decanter with the pond surface as close as practicably possible to the centre line. Nevertheless, other process considerations may require the taking of a different view.

4.7 <u>Power Consumption</u>

The total power input required by a decanter centrifuge comprises a number of separate power components:

$$P_T = P_P + P_{WF} + P_S + P_B \quad (4.78)$$

where P_T is the total power required by the decanter; P_P is the power required to accelerate the process material to the bowl speed at the discharge radius; P_{WF} is the power to overcome windage and friction; P_s is the power required for conveying; and PB is the power for braking. From equation (4.76):

$$P_p = Q_f \rho_f \omega^2 r_d^2 \quad (4.79)$$

where r_d is the process discharge radius.

Naturally, if cake and centrate are discharged at different radii, then these two power components have to be calculated separately. The windage and friction component is given by:

$$P_{WF} = k_7 + k_8\omega + k_9\omega^2 \quad (4.80)$$

where k_7, k_8 and Jc_9 are constants. P_{WF} can be calculated with difficulty but is more generally derived practically in the factory by measuring the power absorbed for different bowl speeds. The conveying component is given by:

$$P_S = NT \quad (4.81)$$

where N is the conveyor differential; and T is the conveyor torque. Similarly, the braking component is:

$$P_B = S_P T_P \quad (4.82)$$

where S_P and T_P are the pinion speed and torque, respectively.

In some types of back-drive the braking power can be regenerated, so that the total power used is reduced.

4.7.1 Main motor sizing

The power of the main motor will be based on the calculation of P_T from equation (4.78), while its physical size will be influenced by its starting requirements. Motor manufacturers rate their motors on the basis of the maximum power delivered at the motor shaft, P_M- This has to be greater than P_T to cater for frictional losses in the drive belts and fluid coupling, if used. Thus, the motor power is P_M:

$$P_M \vartheta_D \vartheta_B = P_T \quad (4.83)$$

where ϑ_D is the fluid drive efficiency; and CB is the efficiency of the drive belts.

The power used, however, will be greater than P_M, due to losses in the motor itself and losses in some control gear when used, such as an inverter.

Power is lost within a motor due to a number of factors, which include:

* iron losses in the magnetizing material, producing heat in the motor rotor and stator;

* friction in the rotor bearings;

* energy needed to drive a cooling fan, internally attached to the motor shaft;

* windage losses; and

* copper losses (the power lost due to the resistance of the windings, sometimes referred to as the ϑ_M, losses).

These five factors combine to give a motor efficiency, CM, of less than unity.

Extra power is also necessarily supplied to the motor, when the power factor is less than unity. The power factor will never be unity, and is a measure of how much the current lags or leads the applied voltage. It is measured as the cosine of the phase angle between current and voltage. When an induction motor is connected to an AC electrical supply, whether the motor does useful work or not, a current is drawn to excite the motor. This current, instantaneously on start-up, lags 90° out of phase with the voltage, and is reactive current, or so-called idle or wattles current. The power factor increases as the motor accelerates.

When the motor is put to work, it will take in addition to its excitation current, a current according to the amount of work to be done. The power factor will increase and will be maximum when the motor works at its full power rating. Thus the power taken from the mains supply will be P_c where:

$$F_P P_C \vartheta_M = P_M \qquad (4.84)$$

where F_P is the power factor.

To combat the anomaly of a low power factor, the installation of a capacitance bank, ideally directly across the motor windings, causes the motor current to reach its maximum value closer to when the voltage does in the alternating cycle. Therefore, a suitably designed capacitor added to an induction motor will reduce the lag of current, by any desired amount. Generally, in industry, because the cost of small capacitors is high, it is more economical and expedient to install large banks of capacitors at the supply source, and automatically switch in and out various sets of capacitors as the demand fluctuates. Moreover, a leading current, which is possible if the capacitor is too large, increases wattless current as much as a lagging current.

Motor manufacturers supply motors in standard increments of power. Thus, after power demand for the decanter is calculated, the next larger size is specified. Motor manufacturers can supply tables of efficiency and power factors for ranges of loading. Also available are performance curves for their motors, giving output torque against rotational speed. The selected size of motor, for economic reasons, needs to be as near as possible to the power demanded by the centrifuge.

Details of the installation need to be considered in the motor specification. These factors would include the ambient temperature, whether the installation is indoors or out, and whether any hazards exist, such as flammable materials in use, and whether the motor will need to be hosed down.

The installed electrical services need to be assessed to ensure that they are adequate for the method of starting contemplated. It is important that supply cables be adequately sized to minimize voltage drops before reaching the main motor.

The power supplied to the motor reduces proportionally to the square of any voltage drop. Nevertheless current will increase to compensate for the drop in voltage, increasing the heating and losses.

Moreover regulations restrict voltage drops to a total of 4%.

If reduced voltage starting is used, it is important that the reduced starting torque is never less than the sum of the friction and windage torque. Since the torque available to accelerate the bowl is equal to the difference between the motor torque and friction and windage torque, the motor may not reach full speed in a reasonable time, unless care is taken.

4.7.2 Main motor acceleration

Most decanter rotating assemblies have high inertias, which can require several minutes' acceleration, or run-up time, t_a- If the run-up time is too short, drive belts will slip and wear out prematurely, or even break. If the run-up time is too long, then the motor could overheat and burn out. The run-up time is:

$$t_a = \frac{\omega_M(J_M - J_P)}{(T_a - T_l)} \quad (4.85)$$

where J_M is the inertia of the motor; J_P is the inertia of the decanter at the motor; u;$_M$ is the motor speed; T_a is the motor torque; and T_l is the reactive torque of the decanter.

The decanter inertia is given by:

$$J_P = \frac{\omega^2}{\omega_M^2} J_D \tag{4.86}$$

where J_D is the inertia of the rotating assembly

Both T_a and T_l vary with speed, and not linearly. Examples of motor and decanter torque/speed curves are shown in Figure 4.9. To use equation (4.85), T_a and T_l are averaged over the speed range from zero to full motor speed.

Given the inertia at the motor shaft, the equations in this section are used to determine whether the torque of the chosen motor is sufficient to accelerate the decanter bowl smoothly to speed, without slippage of the belts. The thermal limits and the torque limit of the number of drive belts used and their cross-sections have to be checked, with the pre-set diameter of the smallest pulley taken into account. Causing the belts to slip will end in their failure, while producing copious amounts of dust in the belt guard, which could be an explosion hazard.

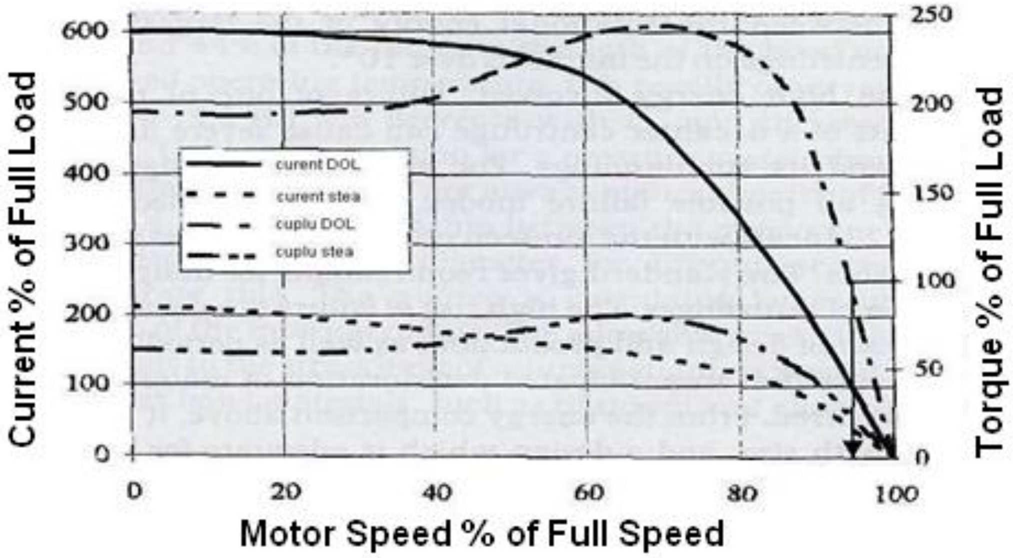

Figure 4.9. Motor and decanter torque / speed curves

4.8 Mechanical Design

The design of a good and reliable decanter centrifuge requires a thorough knowledge of most mechanical engineering disciplines such as machine dynamics, strength of materials, bearing design, and gearbox design. Some of the more important fundamental aspects of the mechanical design of decanter centrifuges will now be discussed.

The need for a careful mechanical design can be illustrated by examining the energy accumulated in a decanter centrifuge in operation. The rotational energy in a medium size decanter with rotational inertia of 50 kg m^2, rotating at 3600 rpm, will be 3.55 MJ. This energy corresponds to the kinetic energy of a vehicle weighing 9.2 tons travelling at 100 km/h. Furthermore, it can be shown that the rotational energy of a decanter centrifuge will increase with the fourth power of the diameter, when the centrifugal force at the bowl wall, g_c, and the length/diameter ratio, A, are kept constant. The ratio of the diameters of the largest to the smallest industrial bowl is over 10. Thus, the ratio between the rotational energy of the largest and the smallest decanter centrifuge on the market is over 10^4.

With the high energy involved, failure of one of the major rotating components of a decanter centrifuge can cause severe damage, both to the decanter and its surroundings. For all decanter designs a risk analysis, evaluating all possible failure modes, must be carried out. A European standard deals with the foreseen risks for centrifuges in normal operating environments. The standard gives requirements for design, verification, and installation of centrifuges. The high risk of failure of a decanter requires a high quality, both of design and production, as well as periodic inspection during use, to ensure that unanticipated deterioration of materials of construction has not occurred. From the energy comparison above, it is seen that the risk increases with size, and a design which is adequate for a small laboratory-scale decanter, may be extremely dangerous on a large industrial scale.

4.8.1 Maximum bowl speed

The bowl shell of a decanter centrifuge will be subjected to both the pressure loads from the material inside the decanter and the centrifugal load acting on the material of the bowl shell. The cylindrical part of the bowl shell will be, for normal decanter designs, the part of the bowl subjected to the highest stress levels. The maximum pressure on the bowl shell is calculated using equation (4.61):

$$P_{2m} = \frac{\rho_M \omega^2}{2}(r_2^2 - r_1^2) \qquad (4.87)$$

where P_{2m} is the maximum pressure at the bowl wall; and P_M is the maximum bulk density of process material ever likely in the decanter.

Defining t_w as the wall thickness of the decanter bowl shell and *p^* as the density of the bowl material, the average tangential stress in the bowl shell can be expressed, for a straight cylinder, as:

$$\sigma_t = \frac{r_2}{t_w} P_2 + \omega^2 \rho_b (r_2 + \frac{t_w}{2})^2 \qquad (4.88)$$

where P_2 is the actual pressure at the bowl wall; and cr_t is the mean tangential stress in the bowl wall.

The formula is equivalent to the well-known pressure vessel formulae for thin cylindrical shells. To ensure safety against failure, the tangential stress must be below a certain allowable stress. According to the European Engineering Directive, the tangential stress shall be kept below 66% of the yield strength and 44% of the ultimate strength of the bowl material, at the maximum allowed operating temperature. It is readily observed that the first term of equation (4.88) will decrease with t_w, and the second term will increase with t_w. Unlike the situation for a pressure vessel, simply increasing the thickness of the bowl shell will not always reduce the risk of failure. On the graph of Figure 4.26, the relationship between the maximum obtainable g-force, and the decanter internal diameter, for different process densities is shown to illustrate this. The g-force is calculated by assuming a bowl thickness of 10% of the internal radius. The allowable stress is set to 240 MPa, which corresponds to the stress limit of a duplex stainless steel at 100°C.

Of course, other bowl materials, such as titanium and alluminium, will give different values.

The pressure inside the bowl will also create an axial force, acting on the end hubs. The maximum axial force, F_x, is found by taking the mean pressure in the pond as half the pressure calculated from equation (4.88), and multiplying by the cross-sectional area of the pond. Thus:

$$F_x = \frac{\pi}{4} \omega^2 \rho_M (r_2^2 - r_1^2)^2 \qquad (4.89)$$

On a large decanter, the axial force on the end hubs will be greater than 10^6 N. The end hubs, and axial fixings, on on the rotor must therefore be designed to withstand this force with a sufficient safety margin. Components of the decanter may be subjected to several other design-dependent static loads, which must be considered by the designer. One example is the axial load, acting on the conveyor, caused by conveying the solids.

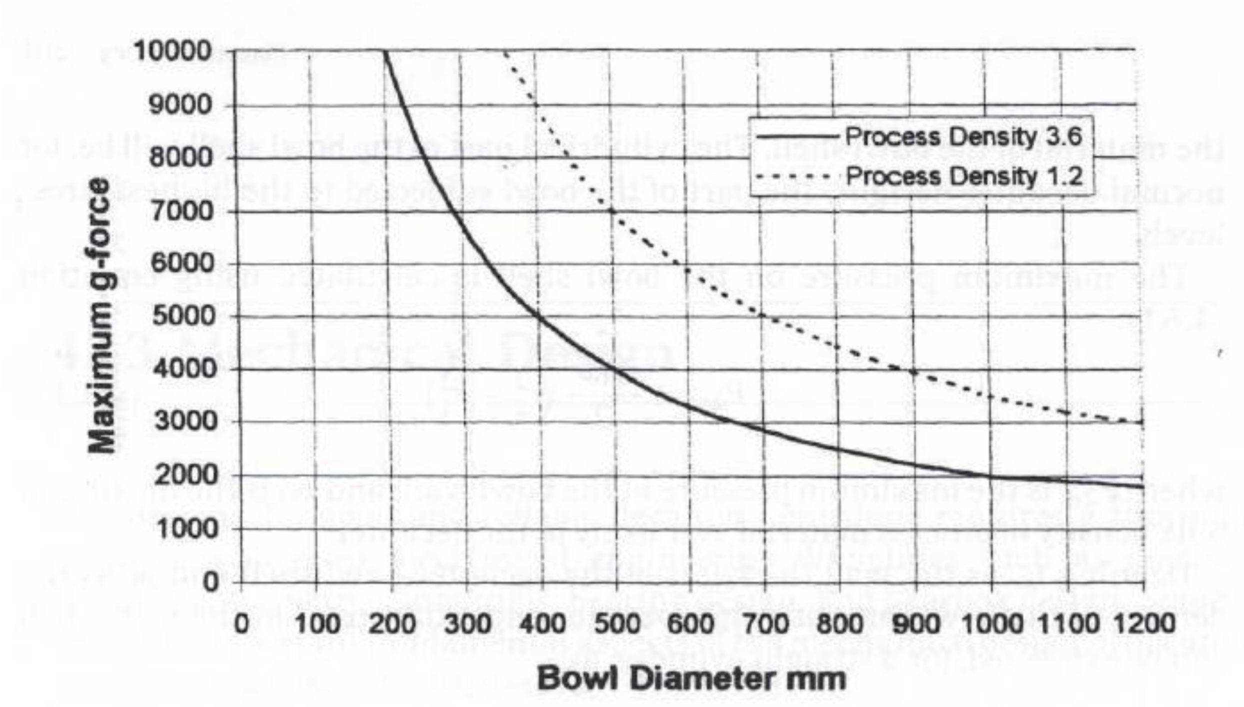

Figure 4.26 Example of the relationship between bowl radius, max g-force and cake density for one material and one relative bowl shell thickness

The decanter will also be subjected to cyclic loads, which can cause mechanical fatigue damage on both the rotating assembly and on the stationary parts. Among the cyclic loads which must be considered by the designer are the bending forces on the shafts, caused by the weight of the rotor, the loads from belt drives, unbalance forces from the rotor, and cyclic loads from frequent starts and stops, or intermittent loading with process material. On complicated hub geometries, often it will be necessary to make a finite element calculation of the stresses to make a proper fatigue evaluation. The notch sensitivity and ductility of the material must be considered.

Quality assurance procedures during manufacturing, such as X-raying of critical welds and die penetrant testing of castings, must be maintained.

4.8.2 Critical speeds

The natural frequencies and critical speeds of a decanter will depend on its actual configuration. A conventional decanter centrifuge consists of a frame holding the double rotor — conveyor and bowl — in rigid bearings. The main motor can be attached to the rotor frame either by a rigid connection or flexibly through vibration isolators. Further the motor can be attached to a sub-frame or to another part of the supporting structure. These factors are more fully described in Section 2.1.

At speeds below operating speed the main frame and the rotor can be considered as one rigid body. If the decanter frame is mounted on soft vibration isolators the decanter assembly will have six natural frequencies and associated vibration modes below the operating speed. The natural frequencies are determined by the spring stiffness of the vibration isolators and the mass and inertia of the total system. When the main motor also is supported on the decanter frame by vibration isolators, it will have six additional natural frequencies below the operating speed.

The important critical speed for a decanter is the lowest speed at which there is significant flexible deformation of the rotor. This speed is called the first rotor critical speed. Decanters will always have a certain unbalance, both due to the handling of solids from the process and due to wear on the rotor. Operating the decanter close to, or just above, the flexible critical speed of the rotor will result in high vibration levels and very high stresses in the rotor components. The critical value of the rotor speed will therefore be an upper limit for the operating speed, and the decanter must be operated below this speed with a safe margin.

The first rotor critical speed will mainly be a function of conveyor geometry, bowl geometry, gearbox weight, main bearing stiffness and conveyor bearing stiffness. The first rotor critical speed will decrease with the length of the decanter. The critical speed of a decanter can be calculated by using a finite element method and verified by measurements. It is normal practice to test decanters at a speed 15-20% above operating speed, to verify the design integrity, and such an over-speed test can also reveal if the operating speed is close to a critical speed. These factors influencing the first rotor critical speed have been more fully covered by Madsen.

4.8.3 Liquid instability problems

Often very large vertical and horizontal vibrations are seen in some speed intervals on decanters when they are started and stopped with liquid inside the bowl. The vibration frequencies in the instability intervals correspond to rigid body natural frequencies of the decanter, but the vibrations are not caused by unbalances. Rather, they are due to interaction between the liquid inside the bowl and the decanter.

The vibrations which occur in some instability speed ranges are sub-synchronous, i.e. the vibration frequency is a fraction (normally about 0.7) of the actual operating speed. If, for example, the decanter is vibrating at an operating speed of 1000 rpm, the vibration frequency will be around 700 rpm.

The vibrations are usually harmless, but as very large forces could be acting on the foundations of the decanter, the manufacturer must supply information on the magnitude of these forces, and the foundation must be designed to withstand these forces. By having a constant flow of water to the decanter during starting and stopping, the instability vibrations can be suppressed.

The complicated dynamic phenomenon, which is related to all rotating cylinders with an internal annulus of liquid, has been dealt with in a number of publications.

4.8.4 Length/diameter ratio

In general the bowl strength, the first rotor critical speed, and the maximum permissible speed of the main bearings control the maximum speed at which a' decanter can be operated.

It is argued that a long, slender decanter centrifuge will give advantages with respect to overall economy, power consumption and process performance. For a centrifuge with the *L/D* ratio above 3, the critical speed will often be the main factor controlling the maximum obtainable speed and it can therefore be desirable to increase the critical speed in order to obtain a high *L/D* ratio without sacrificing the maximum operating speed.

When a decanter bowl, for calculation purposes, is approximated to a beam, its natural frequency is inversely proportional to the square of its length. In that g-force is proportional to the square of bowl speed, and it is necessary to keep resonance frequency above bowl speed, the maximum bowl speed is proportional to its length to the fourth power. To obtain g-forces in the range 2000-3000, generally required for commercial decanters, the maximum length-to-diameter ratio, for the most frequently used designs, has to be restricted to a little over 4.0.

In order to increase the critical speed of the rotor a number of different modifications can be made to the rotor system. By supporting one or both main bearings in a flexible pillow block the first critical speed of the rotor can be turned into a low speed rigid-body motion for the rotor. It can then be operated super-critically with respect to this critical speed. Other modifications are the floating conveyor and the separately supported gearbox.

These sorts of modifications have been utilized by Alfa Laval in producing a decanter with an *L/D* of over 5 which can operate with up to 10 000 g.

How these modifications extend the possible *L/D* ratio and clarification capacity was graphed by Madsen, for 250 mm diameter bowls, and reproduced in Figure 4.27.

4.8.5 Bearing life

One of the most frequent reasons for breakdown of decanters is failure of one of the main bearings. The operating conditions of decanters are often very arduous, and there can be a high load on the bearings. The failure of a main bearing on a properly designed decanter will not lead to a dangerous situation, but it can cause damage to other parts of the decanter, and expensive downtime.

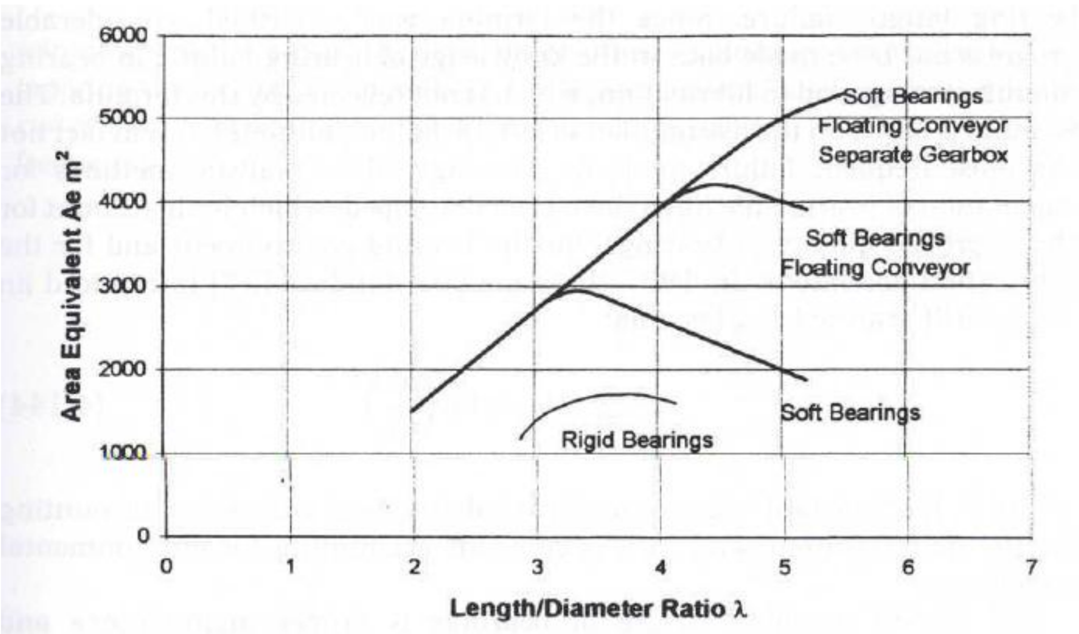

Figure 4.27. Maximum area equivalents & L/Dfor various 250 mm diameter bowl

The bearing life is defined as the number of revolutions or number of hours at constant speed a bearing will operate before it fails. According to the international standard [27], and based on the assumption that the bearing will fail by fatigue, the expected life of a bearing is calculated by the simple formula:

$$L_{10} = \left(\frac{C}{C_E}\right)^w \qquad (4.90)$$

where L_{10} is the expected life measured in 10^6 revolutions; C is the dynamic load capacity of the bearing, a characteristic figure for the bearing, determined by the manufacturer in accordance with the ISO standard; C_E is the equivalent dynamic load, calculated from the dynamic and static loads; and *w* is a number depending upon the bearing type (e.g. for ball bearings w = 3 and for roller bearings w = 10/3).

Both C and C_E are expressed in a unit of force. The L_{10} life is also sometimes referred to as the B_{10} life.

For a machine rotating at a constant speed, *n,* in revolutions per minute, the expected life can be expressed in expected hours of operation, L_{10h}:

$$L_{10h} = \frac{10^6}{60n}(\frac{C}{C_E})^w \quad (4.91)$$

where L_{10h} is the expected life in hours; and *n* is the number of revolutions per minute

This simple formula was developed around 1950 and was based on data for bearing fatigue failure. Since the formula was published, considerable progress has been made both in the knowledge of bearing failure, in bearing manufacturing, and in lubrication, which is not reflected by this formula. The formula is based on the assumption of fatigue failure, although it is in fact not the most frequent failure mode for bearings. More realistic methods for calculation of bearing life have since been developed, which both account for the improved quality of bearings, for the bearing environment and for the ' lubrication conditions. In 1977 the same ISO standard introduced an adjusted life-rating L_{10ah} formula:

$$L_{10ah} = b_1 b_2 b_3(\frac{C}{C_E})^w \quad (4.92)$$

where b_1 is a constant accounting for reliability; b_2 is a constant accounting for the material used; and b_3 is a constant accounting for environmental conditions.

The key to avoiding failure of bearings is proper maintenance and lubrication. By monitoring and analyzing vibrations measured with sensors directly on the bearing housings of a rotating machine, bearing faults can often be detected before they lead to failure. Several systems for detection of bearing faults, by continuous vibration monitoring, are available, and some decanter manufacturers offer their own specialized systems. For critical installations, and installations with several decanters, such monitoring systems can be a good investment, to avoid inconvenient bearing failures, damage to the machine, and unnecessary downtime.

4.8.6 Gearbox life

The decanter manufacturer will often quote the expected life of the gearbox. This will be based on the fatigue life of the gear teeth, which is proportional to the ninth power of the torque encountered. Thus, one has to be extremely careful not to overload the gearbox above its torque rating. An 8% increase of torque over its rating will halve the expected life of the gearbox.

4.8.7 Feed tube

Each component of the decanter has its own natural frequency, even the stationary components, which could resonate sympathetically if this frequency is close to the bowl speed. The feed tube is a good example, being a long, thin tube. Apart from the inverse square relationship with length, resonance frequency is also proportional to the fourth power of diameter in its simplest form. Unless care is taken the feed tube can be caused to resonate like a tuning fork.

The design engineer thus endeavors to maximize diameter and minimize the length of the feed tube. Other techniques employed include tapering the feed tube and making it of lighter materials. Of course the double concentric tube used, when flocculent is added, helps to increase the natural frequency.