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Sevenson's 16-Inch DSC Marlin Dredge with Dredge Tender
(Photo Courtesy of Sevenson)

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The *Journal of Dredging* is published by the Western Dredging Association (WEDA) to provide dissemination of technical and project information on dredging engineering topics. The peer-reviewed papers in this practice-oriented journal will present engineering solutions to dredging and placement problems, which are not normally available from traditional journals. Topics of interest include, but are not limited to, dredging techniques, hydrographic surveys, dredge automation, dredge safety, instrumentation, design aspects of dredging projects, dredged material placement, environmental and beneficial uses, contaminated sediments, litigation, economic aspects, and case studies.

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AN OVERVIEW OF THEORIES DESCRIBING HEAD LOSSES IN SLURRY TRANSPORT: A TRIBUTE TO SOME OF THE EARLY RESEARCHERS

Sape A. Miedema[1]

ABSTRACT

Since the 1950's many researchers have tried to create a physical mathematical model in order to predict the head losses in slurry transport. One can think of the models of Durand and Condolios (1952) and Durand (1953), Worster and Denny (1955), Newitt et al. (1955), Gibert (1960), Fuhrboter (1961), Jufin and Lopatin (1966), Zandi and Govatos (1967) and Zandi (1971), Turian and Yuan (1977), Doron et al. (1987) and Doron and Barnea (1993), Wilson et al. (1992) and Matousek (1997). Some models are based on phenomenological relations and, thus, result in semi-empirical relations; others tried to create models based on physics, like the two and three layer models. It is, however, the question whether slurry transport can be modeled this way at all. Observations in our laboratory show a process that is often non-stationary with respect to time and space. Different physics occur depending on the line speed, particle diameter, concentration, and pipe diameter. These physics are often named flow regimes: fixed bed, shear flow, sliding bed, heterogeneous transport, and (pseudo) homogeneous transport. It is also possible that more regimes occur at the same time, like, a fixed bed in the bottom layer with heterogeneous transport in the top layer. It is the observation of the author that researchers often focus on a detail and sub-optimize their model, which results in a model that can only be applied for the parameters used for their experiments. An analysis of the experiments of some of the early researchers reveals that their equations are often used in a wrong way. Applying these theories in the correct way makes them still very valuable. The theories of Durand and Condolios (1952) and Newitt et al. (1955) will be analyzed and the corrections and issues will be discussed. Based on these theories a new (or modified) regime diagram has been constructed.

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THE DURAND AND CONDOLIOS (1952) AND DURAND (1953) AND GIBERT (1960) MODEL

Durand and Condolios (1952) and Durand (1953) carried out experiments in solids (mostly sand and gravel) with a d_{50} between 0.18 mm and 22.5 mm in pipes with a diameter D_p from 40 to 580 mm and volumetric concentrations C_{vt} from 2% to 22%. Gibert (1960) analyzed the data of Durand and Condolios (1952) and summarized the results. A possible parameter to define the solids effect is the relative excess pressure loss p_{er}:

$$\mathbf{p_{er}} = \left(\frac{\mathbf{i_m} - \mathbf{i_{fl}}}{\mathbf{i_{fl}}} \right) \tag{1}$$

The first step Durand and Condolios (1952) carried out was to define a parameter Φ, which is the relative excess pressure loss per divided by the concentration Cvt and plot the pressure loss data of two sands with the parameter Φ versus the line speed vls. The transport regime of the data points is heterogeneous, which means that there will not be much difference between the spatial and the transport concentration. The parameter Φ was defined as:

$$\mathbf{\Phi} = \left(\frac{\mathbf{p_{er}}}{\mathbf{C_{vt}}} \right) = \left(\frac{\mathbf{i_m} - \mathbf{i_{fl}}}{\mathbf{i_{fl}} \cdot \mathbf{C_{vt}}} \right) \text{ With, in general for the pressure gradient } \mathbf{i}\text{: } \mathbf{i} = \frac{\Delta \mathbf{p}}{\rho_{\mathbf{fl}} \cdot \mathbf{g} \cdot \Delta \mathbf{L}} \tag{2}$$

The volumetric concentration C_{vt} is the transport concentration, which for high line speeds and heterogeneous transport is considered to be almost equal to the spatial volumetric concentration C_{vs}, assuming that the slip between the particles and the carrying fluid can be neglected. At smaller line speeds, close to the limit deposit velocity this is not always true, but not all papers are clear about this. Figure 1 shows the resulting curves of experiments at different volumetric concentrations in the heterogeneous regime, where the data points of different concentrations apparently converge to one curve. The data points of the two types of sand still result in two different curves. Whether the linear relationship between the relative solids excess pressure is exactly linear with the concentration C_{vt} requires more experiments and especially experiments at much higher concentrations in order to see if hindered settling will have an effect, but for low concentrations, the conclusion of Durand and Condolios (1952) seems valid. The second step Durand and Condolios (1952) carried out was to investigate the influence of the pipe diameter D_p. Instead of using the line speed v_{ls} on the horizontal axis, they suggested to use the Froude number of the flow, $\mathbf{Fr_{fl}} = \mathbf{v_{ls}} / \sqrt{\mathbf{g} \cdot \mathbf{D_p}}$. Figure 2 shows how the data points of two sands and four pipe diameters converge to two curves, one for each sand. Within the range of the pipe diameters applied and the range of the particle diameters applied, the assumption of Durand and Condolios (1952) that the parameter Φ is proportional to the square root of the pipe diameter and reversely proportional to the flow Froude number, seems very reasonable. Whether this proportionality is linear or to a certain power close to unity is subject to further investigation.

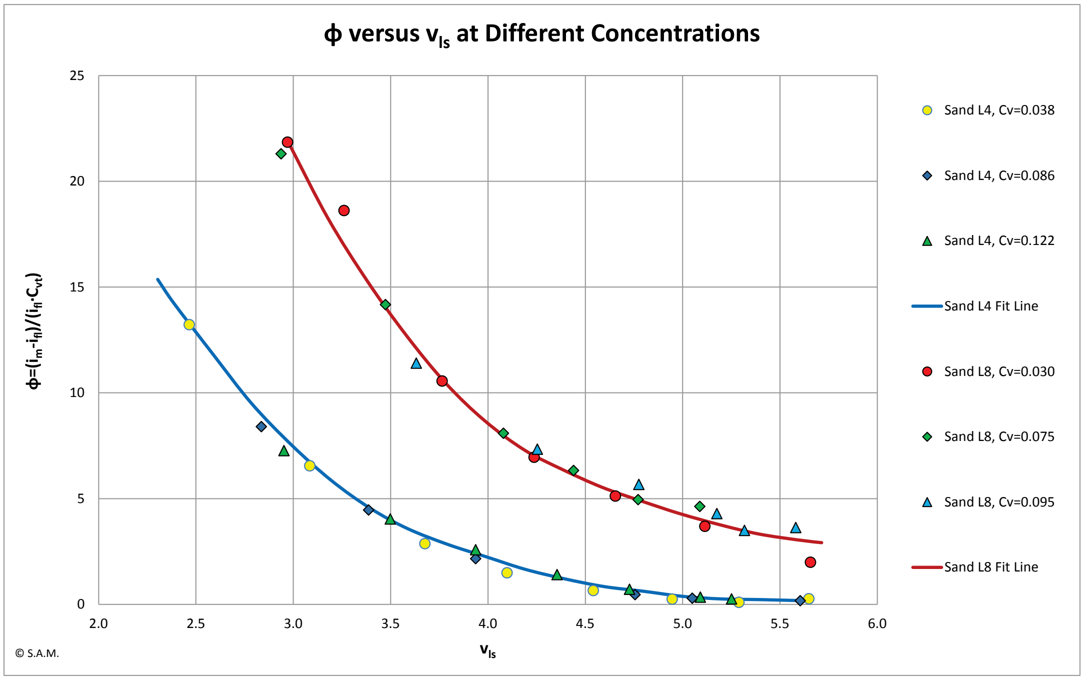

Figure 1. Φ at different concentrations as a function of the line speed in a D_p=0.150 m pipe

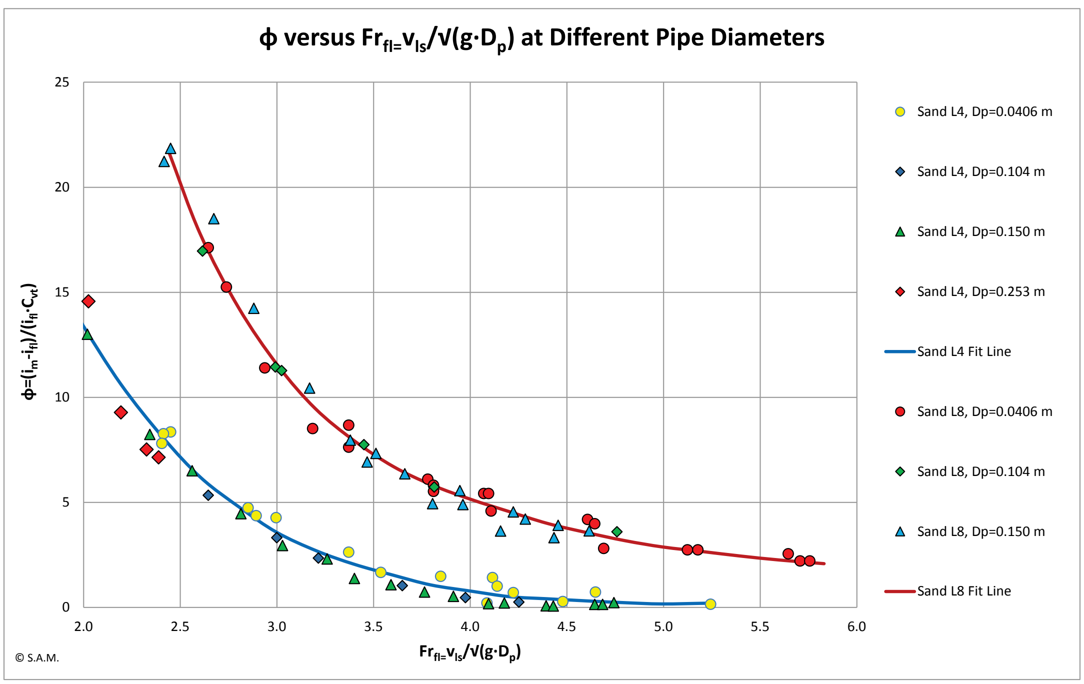

Figure 2. Φ at different pipe diameters as a function of the flow Froude number Fr_{fl}

Now that proportionalities have been found between the parameter Φ on one hand and the concentration C_{vt} and the pipe diameter D_p on the other hand, Durand and Condolios (1952) investigated the influence of the particle diameter. Figure 3 shows the results of experiments in four sands and one gravel ranging from a d_{50}=0.20 mm to a d_{50}=4.2 mm in a D_p=0.150 m pipe. Figure 4 shows the results of seven gravels. Figure 4 shows that in the case of gravels, the relation between Φ and Fr_{fl} does not depend on the particle size, but Figure 3 shows that for smaller particles it does. A parameter that shows such a behavior is the drag coefficient C_D as used to determine the terminal settling velocity v_t of the particles. For small particles, the drag coefficient depends strongly on the particle diameter, but for particles larger than about 1 mm, the drag coefficient is a constant with a value of about 0.445 for spheres and up to about 1-1.5 for angular sand grains. Instead of using the drag coefficient directly, Durand and Condolios (1952) choose to use the particle Froude number, $\mathbf{Fr_p = v_t / \sqrt{g \cdot d}}$. It cannot be emphasized enough that this particle Froude number is different from the reciprocal of the drag coefficient C_D, although it is mixed up in many textbooks, together with some other errors, resulting in the wrong use and interpretation of the Durand and Condolios (1952) results.

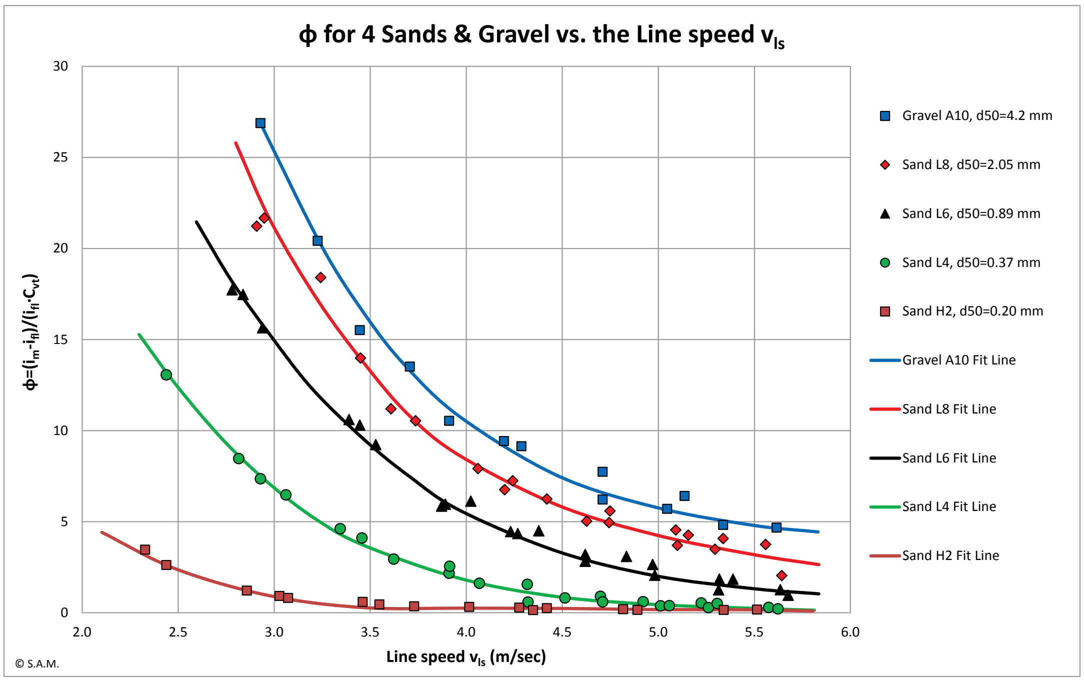

Figure 3. Φ as a function of v_{ls} for four sands and one gravel

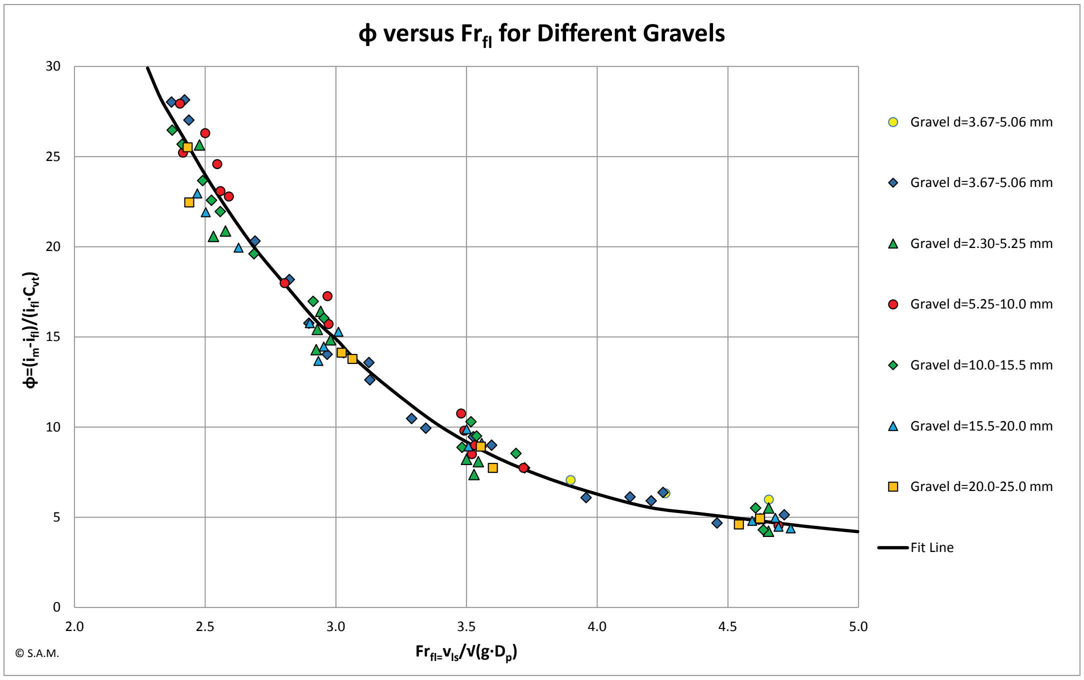

Figure 4. Φ as a function of Fr_{fl} in seven gravels

In order to make the five curves in Figure 3 converge into one curve, Durand and Condolios (1952) extended the parameter on the horizontal axis with the particle Froude number to:

$$\psi=\left(\frac{v_{ls}}{\sqrt{g\cdot D_p}}\right)^2\cdot\left(\frac{v_t}{\sqrt{g\cdot d_{50}}}\right)^{-1}\quad\text{with: } Fr_{fl}=\left(\frac{v_{ls}}{\sqrt{g\cdot D_p}}\right),\ Fr_p=\frac{1}{\sqrt{C_x}}=\left(\frac{v_t}{\sqrt{g\cdot d_{50}}}\right),\ C_x=\frac{g\cdot d_{50}}{v_t^2}=Fr_p^{-2}\qquad(3)$$

Which are the Froude number of the flow Fr_{fl} and the Froude number of the particle Fr_p, which also looks like a sort of reciprocal drag coefficient (but it's not), which will be explained later. The variable ψ can now also be written in terms of the two Froude numbers as defined above. Durand and Condolios (1952) plotted all their data against this new parameter ψ as is shown in Figure 5 on linear scales. The assumption of using the particle Froude number Fr_p seemed to be successful. The curves for different sands, pipe diameters, and concentrations converged into one curve with the equation:

$$\psi=Fr_{fl}^2\cdot Fr_p^{-1}\quad\text{and}\quad\Phi=K\cdot\psi^{-3/2}\qquad\text{with : } K=176\qquad(4)$$

The number of 176 is deduced from the original graph of Durand and Condolios (1952), source Bain and Bonnington (1970). The next step was to investigate the influence of the relative submerged density of the particles. Gibert (1960) reported on a set of experiments on plastic, sand, and Corundum with three different relative densities. Figure 6 shows the results of these experiments. By extending the fluid flow Froude number Fr_{fl}, with the relative submerged density R_{sd} the data points converge to one curve for each particle diameter. The fluid flow Froude number Fr_{fl} is modified according to:

$$\mathbf{Fr_{fl,m}} = \frac{\mathbf{v_{ls}}}{\sqrt{\mathbf{g \cdot D_p \cdot R_{sd}}}} \tag{5}$$

Applying this modified Froude number, nothing changes in the equation for sand with a relative submerged density of about R_{sd}=1.65, using in a new constant of 83 instead of 176 for K. The equation for the particle Froude number does not change according to Gibert (1960) because of the relative submerged density R_{sd} and remains $\mathbf{Fr_p = v_t / \sqrt{g \cdot d}}$. The final equation of Durand and Condolios (1952) and later Gibert (1960) now becomes:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + \Phi \cdot C_{vt}\right) \text{ with: } \Phi = \frac{i_m - i_{fl}}{i_{fl} \cdot C_{vt}} = K \cdot \psi^{-3/2} = K \cdot \left(\frac{v_{ls}^2}{g \cdot D_p \cdot R_{sd}} \cdot \sqrt{C_x}\right)^{-3/2} \text{ with: } K \approx 83} \tag{6}$$

Durand and Condolios (1952) and Gibert (1960) do not claim that equation (6) is rigorously exact and believe that a more accurate, although more complex, means of correlating their data is possible. They claim only that their equation brings all their results together quite well, especially if one considers that 310 test points cover a broad range of pipe diameters (D_p=40 to 580 mm), particle diameters (d_{50}=0.2 to 25 mm) and concentrations (C_{vt}=2% to 22.5%). In normal sands, there is not only one grain diameter, but a grain size distribution has to be considered. The Froude number for a grain size distribution can be determined by integrating the Froude number as a function of the probability according to:

$$\mathbf{Fr_p} = \frac{\mathbf{v_t}}{\sqrt{\mathbf{g \cdot d}}} = \frac{\mathbf{1}}{\sqrt{\mathbf{C_x}}} = \frac{\mathbf{1}}{\int_0^1 \frac{\sqrt{\mathbf{g \cdot d}}}{\mathbf{v_t}} \mathbf{dp}} = \frac{\mathbf{1}}{\sum_{i=1}^{n} \left(\sqrt{\mathbf{C_x}}\right)_i \cdot \mathbf{\Delta p_i}} \tag{7}$$

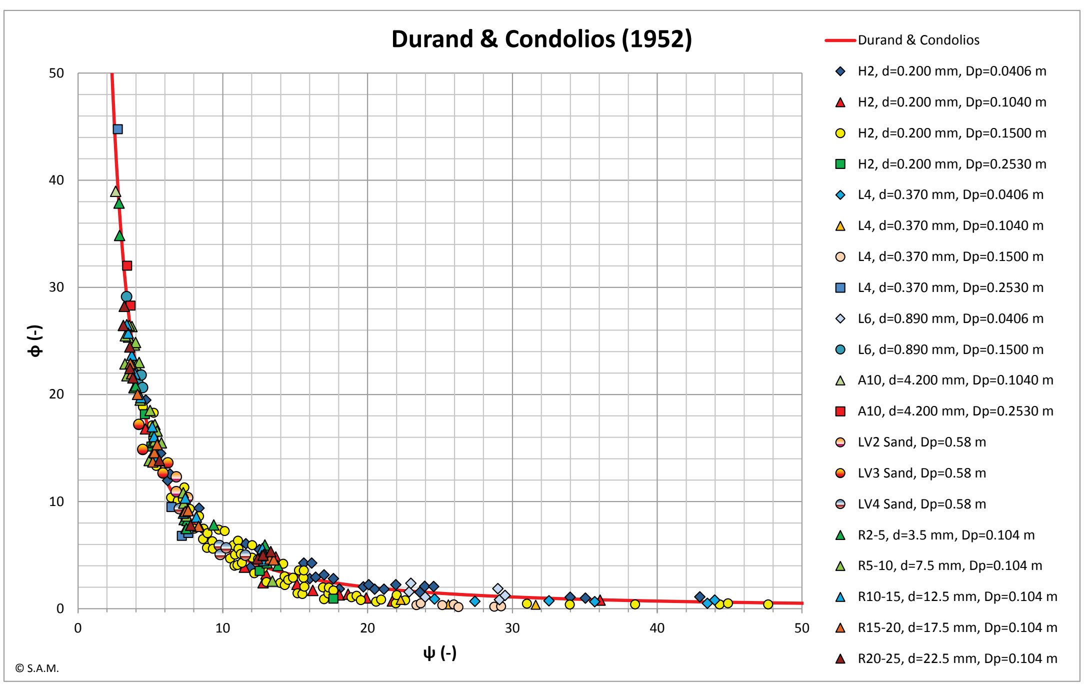

Figure 5. The relationship of the Durand and Condolios (1952) model

It is also possible to split the particle size distribution into n fraction and determine the weighted average particle Froude number. Gibert (1960) published a graph with values for the particle Froude number that match the findings of Durand and Condolios (1952). Figure 7 shows these published values. If one uses the values of Gibert (1960), the whole discussion about whether the C_D or the C_x value should be used can be omitted.

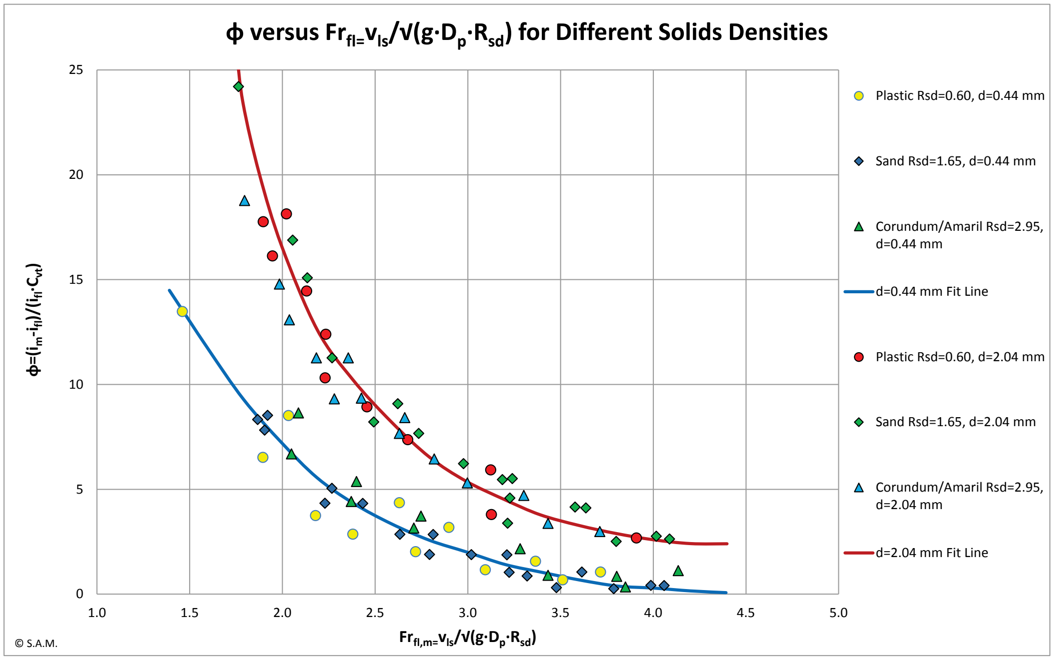

Figure 6. The influence of the relative submerged density R_{sd}

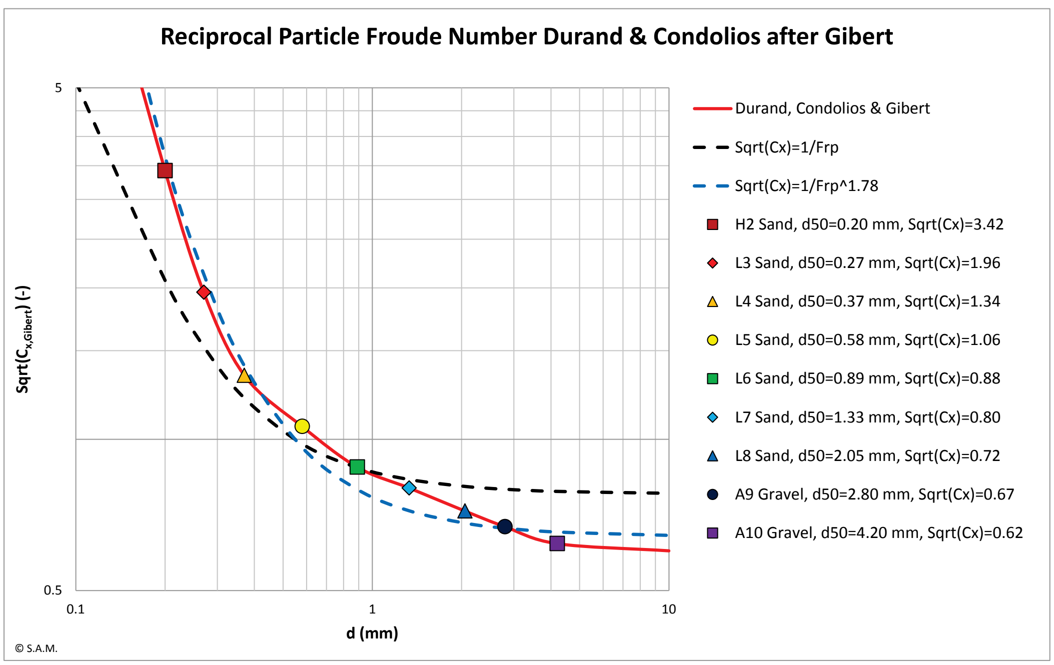

Figure 7. Modified reciprocal particle Froude number, determined experimentally for various sorts of sand and gravel by Durand and Condolios (1952) and Gibert (1960)

Analyzing this table however, shows that a very good approximation of the table values can be achieved by using the particle Froude number to the power 1.78 instead of the power 1, assuming that the terminal settling velocity v_t is determined correctly for the solids considered (Stokes, Budryck, Rittinger or Zanke).

$$\sqrt{C_{x,Gibert}} = \sqrt{C_x}^{1.78} = C_x^{0.89} \tag{8}$$

Figure 7 shows the original data points, the theoretical reciprocal particle Froude number using the Zanke (1977) equation for the terminal settling velocity of sand particles and the curve using a power of 1.78. Only for large particles, there is a small difference between the original data and the theoretical curve applying the power of 1.78.

The Limit Deposit Velocity

When the flow decreases, there will be a moment where sedimentation of the grains starts to occur. The corresponding line speed is called the limit deposit velocity. Often other terms are used like the critical velocity, critical deposition velocity, deposit velocity, deposition velocity, settling velocity, minimum velocity, or suspending velocity. Here we will use the term limit deposit velocity. Although in literature researchers do not agree on the formulation of the limit deposit velocity, the value of the limit deposit velocity is often derived by differentiating equation (6) with respect to the line speed v_{ls} and taking the value of v_{ls} where the derivative equals zero. This gives:

$$v_{ls,cr} = \sqrt{g \cdot D_p \cdot R_{sd} \cdot \left(\frac{K \cdot C_{vt}}{2}\right)^{2/3} / \sqrt{C_x}} \tag{9}$$

At line speeds less than the limit, deposit velocity sedimentation occurs and part of the cross-section of the pipe is filled with sand, resulting in a higher flow velocity above the sediment. Durand and Condolios (1952) assume equilibrium between sedimentation and scour, resulting in a Froude number equal to the Froude number at the limit deposit velocity.

$$Fr_{cr} = \frac{v_{ls,cr}}{\sqrt{g \cdot D_p \cdot R_{sd}}} = \sqrt{\left(\frac{K \cdot C_{vt}}{2}\right)^{2/3} / \sqrt{C_x}} \tag{10}$$

By using the hydraulic diameter concept, at lines speeds less than the limit deposit velocity, the resistance can be determined applying equation (3) using the hydraulic diameter instead of the pipe diameter. At low flows, resulting in small hydraulic diameters, the pressure gradient may be so large that a sliding bed may occur, limiting the pressure gradient. But, Durand and Condolios (1952) did not perform experiments in that flow region. Equation (9) can be written in the form of the Durand limit deposit velocity based on the minimum pressure loss, according to:

$$v_{ls,cr} = \left(\frac{K \cdot C_{vt}}{2}\right)^{1/3} \cdot \sqrt{\frac{1}{2 \cdot \sqrt{C_x}}} \cdot \sqrt{2 \cdot g \cdot D_H \cdot R_{sd}} = F_L \cdot \sqrt{2 \cdot g \cdot D_H \cdot R_{sd}} \tag{11}$$

With minor adjustments for hindered settling and the ratio between the particle size d and the hydraulic diameter $D_{p,H}$ according to Wasp et al. (1970), the following equation is derived by Miedema (1995). The coefficient β is determined with the Richardson and Zaki (1954) equation. Equation (10) can be modified to match the original Durand and Condolios (1952) graph according to equation (12):

$$F_L = 1.9 \cdot \left(\frac{K \cdot C_{vt}}{2}\right)^{1/3} \cdot \sqrt{\frac{\left(1 - C_{vt}\right)^{\beta}}{2 \cdot \sqrt{C_x}}} \cdot \left(\frac{1000 \cdot d}{D_H}\right)^{1/6} \cdot e^{-\frac{d}{0.0006}} + 0.6 + 1.3 \cdot \left(1 - e^{-\frac{d}{0.0006}}\right) \tag{12}$$

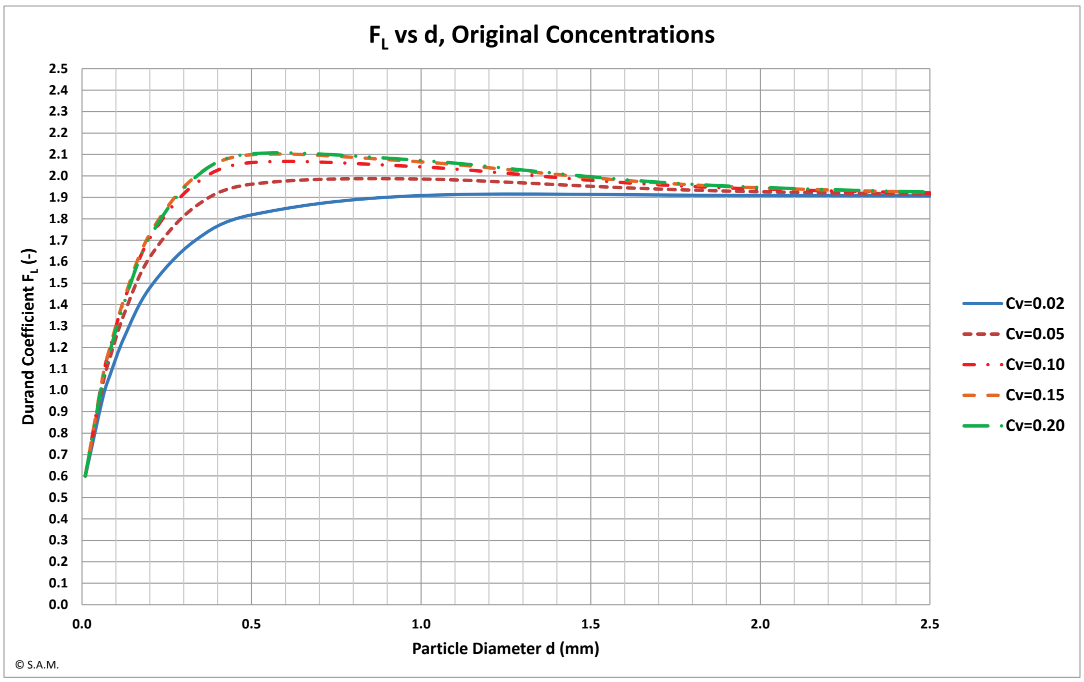

Figure 8. Durand F_L approximation according to equation (12) (D_p=0.5m) for the original concentrations

Equation (12) gives a good approximation of the original F_L graph published by Durand and Condolios (1952). The original graph is shown in Figure 8. Durand and Condolios (1952) and Gibert (1960) used concentrations up to 15% in their graph. Often it is referred to that for higher concentrations the curve of 15% should be used. It should be noted here that Durand and Condolios (1952) did their experiments in medium pipe diameters. The pressure gradients in larger pipes are often not high enough to result in a sliding bed. So, it is assumed that the limit deposit velocity of Durand and Condolios (1952) is the velocity below which particles are at rest on the bottom of the pipe, forming a stationary bed and not a sliding bed. Figure 9 shows the limit deposit velocity as a function of the pipe diameter D_p and the particle diameter d, according to equation (12), matching the findings of Durand and Condolios (1952) and Gibert (1960).

Gibert (1960) analyzed the measurements of Durand and Condolios (1952) and created Figure 10, showing the Froude number F_L at the limit deposit velocity as a function of the volumetric transport concentration C_{vt} for five sands and gravel for pipe diameters D_p of 0.04 m and 0.15 m. He concluded that on average sand H2 has a coefficient F_L=1.7 for higher concentrations, while the other sands and gravel have an F_L=2.1. Tests in a D_p=0.7 m pipeline at concentrations of 15%-20% have resulted in F_L=2.1-2.3. The tendencies found in Figure 10 confirm the findings of Durand and Condolios (1952), but the asymptotic value of about 1.9 in Figure 8 is a bit low, considering that deposition should be avoided.

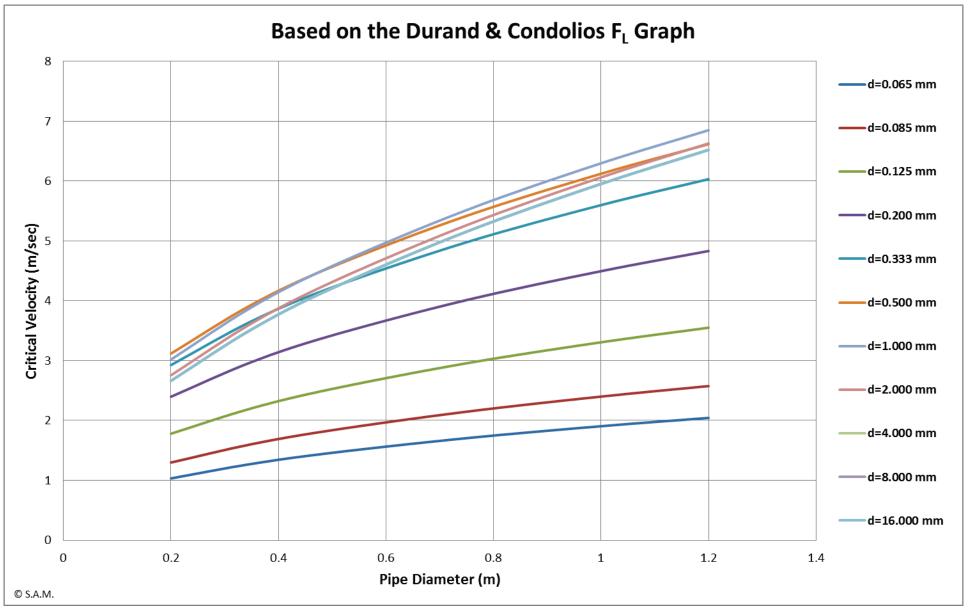

Figure 9. Limit deposit velocity according to equation (12) at a concentration of 0.1

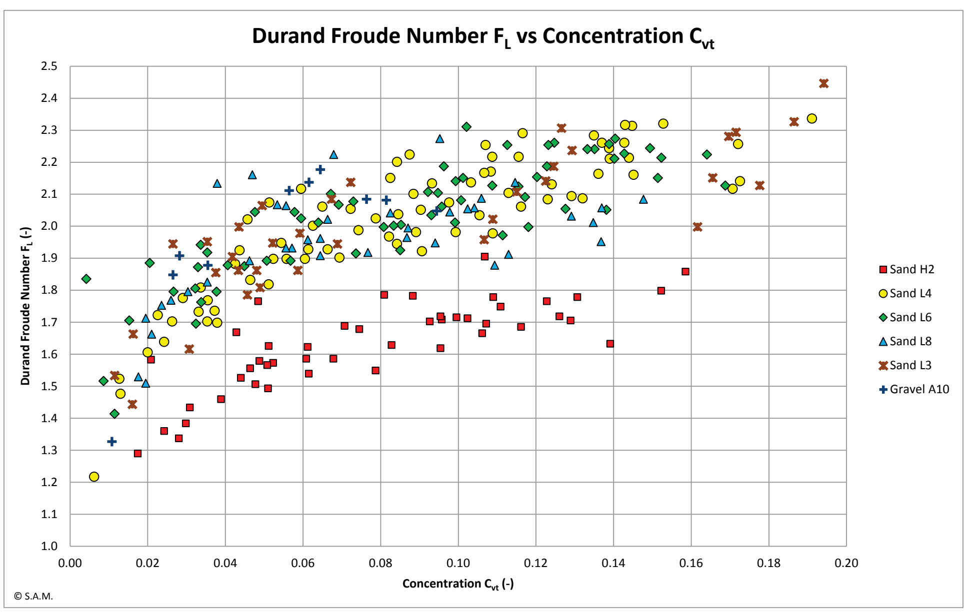

Figure 10. The limit deposit velocity Froude number F_L as a function of the transport concentration C_{vt} for 5 different sands and gravel

ISSUES REGARDING THE DURAND AND CONDOLIOS (1952) AND GIBERT (1960) MODEL

There are six issues to be discussed:

1. The drag coefficient of Durand and Condolios (1952) versus the real drag coefficient.
2. The drag coefficient of Gibert (1960).
3. The relative submerged density R_{sd} as part of the equation.
4. The F_L value as published by many authors.
5. The friction coefficient λ.
6. The solids effect term in the pressure gradient equation.

The Drag Coefficient of Durand and Condolios vs the Real Drag Coefficient

It should be noted that Durand and Condolios (1952), Gibert (1960), and Worster and Denny (1955) use the particle Froude number in their equations and not the particle drag coefficient. The virtual drag coefficient, as used by Durand and Condolios (1952), Gibert (1960), and Worster and Denny (1955) is:

$$\mathbf{C_x = \frac{g \cdot d}{v_t^2} = Fr_p^{-2} \quad with: \quad Fr_p = \frac{v_t}{\sqrt{g \cdot d}}} \tag{13}$$

The drag coefficient C_D as used in the equation for the terminal settling velocity is:

$$\mathbf{v_t = \sqrt{\frac{4 \cdot g \cdot (\rho_s - \rho_{fl}) \cdot d \cdot \xi}{3 \cdot \rho_{fl} \cdot C_D}} = \sqrt{\frac{4 \cdot g \cdot R_{sd} \cdot d \cdot \xi}{3 \cdot C_D}} \quad \Rightarrow \quad C_D = \frac{4 \cdot R_{sd} \cdot \xi}{3} \cdot \frac{g \cdot d}{v_t^2}} \tag{14}$$

So, the relation between the drag coefficient C_D and the virtual drag coefficient according to Durand and Condolios (1952) C_x is:

$$\mathbf{C_D = \frac{4 \cdot R_{sd} \cdot \xi}{3} \cdot C_x \quad or \quad \frac{1}{\sqrt{C_x}} = \frac{v_t}{\sqrt{g \cdot d}} = \sqrt{\frac{4 \cdot R_{sd} \cdot \xi}{3}} \cdot \frac{1}{\sqrt{C_D}}} \tag{15}$$

For irregular shaped sand particles with a shape factor of ξ=0.5-0.7 and a relative submerged density of R_{sd}=1.65 this results in a drag coefficient almost equal to the Durand and Condolios, Gibert (1960), and Worster and Denny (1955) coefficient. The term $\mathbf{\frac{4 \cdot R_{sd} \cdot \xi}{3} = 2.2 \cdot \xi}$ is almost unity for a shape factor of ξ=0.5, and because it is used by its square root, the error is just 5%. However for spheres with ξ=1.0 this factor is 2.2 which cannot be neglected. For solids with another relative submerged density however, there may be a much bigger difference. Zandi and Govatos (1967) and many others also use the Durand and Condolios (1952), Gibert (1960), and Worster and Denny (1955) coefficient, although they name it C_D. It is often not clear whether authors used the C_D value or just named it C_D using the C_x value. Because the error depends on both the shape factor ψ and the relative submerged density R_{sd}, the original particle Froude number Fr_p should be used, because the relation of Durand and Condolios (1952), matching their experiments is based on this particle Froude number.

The Drag Coefficient of Gibert (1960)

Gibert (1960) published a table with numerical values for the virtual drag coefficient or particle Froude number. If one uses the values of Gibert (1960), the whole discussion about whether the C_D or the C_x value should be used can be omitted. Analyzing this table however, shows that a very good approximation of the table values can be achieved by using the particle Froude number to the power 1.7 instead of the power 1, assuming that the terminal settling velocity v_t is determined correctly for the solids considered (Stokes, Budryck, Rittinger or Zanke).

$$\sqrt{\mathbf{C_{x,Gibert}}} = \sqrt{\mathbf{C_x}}^{\mathbf{1.78}} = \mathbf{C_x^{0.89}} \tag{16}$$

Together with the correction of Gibert (1960) regarding the constant of 85/180, the equation now becomes:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + \Phi \cdot C_{vt}\right) \quad With: \quad \Phi = 85 \cdot \left(\frac{v_{ls}^2 \cdot C_x^{0.89}}{g \cdot D_p \cdot R_{sd}}\right)^{-3/2}} \tag{17}$$

The Relative Submerged Density as Part of the Equation

In a number of books and publications, the relative submerged density R_{sd} is added to both the flow Froude number Fr_{fl} and the particle Froude number Fr_p. This is to enable the use of the equations for different types of solids. Gibert (1960) clearly states that the relative submerged density should only be added to the flow Froude number Fr_{fl} and not to the particle Froude number Fr_p, because it is already part of the terminal settling velocity v_t. Equation (17) is, thus, the final equation of the Durand and Condolios (1952) and Gibert (1960) model.

The F_L Value as Published by Many Authors

The issue of the critical deposit velocity Froude number F_L is of great importance. In their original publication, Durand and Condolios (1952) published four graphs showing the F_L value as a function of the volumetric transport concentration C_{vt} for the sands H2, L4, L6 and L8 (4 different particle diameters), these graphs are summarized in Figure 11. The critical deposit velocity Froude number is defined as:

$$\mathbf{F_L = Fr_{cr} = \frac{v_{ls,cr}}{\sqrt{g \cdot D_p}}} \tag{18}$$

Based on Figure 11, the relation between F_L and the particle diameter d can be derived at different concentrations. With some curve fitting and extrapolation, the graph as published by Durand and Condolios (1952) was constructed. The graphs, as shown here, are directly constructed from the original data points, without curve fitting and/or extrapolation. The graph, as published, has an asymptotic value for large particle diameters of about 1.9. Figure 8 shows a reconstruction of the original Durand and Condolios (1952) graph. Durand (1953) published his findings in the English language, while the original paper was in the French language. He modified the F_L coefficient, by including the relative submerged density and a factor 2, but he divided the vertical axis only by $\sqrt{2}$, resulting in an asymptotic value of 1.34. Because (probably) most authors of books and publications read the English paper from 1953, the incorrect graph was copied and can be found in almost every textbook about slurry transport. The vertical axis should have been divided by $\sqrt{2 \cdot R_{sd}}$ resulting in an asymptotic value of about 1.05, a difference of about 28%. Using this incorrect graph results in an overestimation of F_L and, thus, the limit deposit velocity by about 28%.

$$\mathbf{F_L = \frac{Fr_{cr}}{\sqrt{2 \cdot R_{sd}}} = \frac{v_{ls,cr}}{\sqrt{2 \cdot g \cdot D_p \cdot R_{sd}}}} \tag{19}$$

Later Condolios and Chapus (1963A) and (1963B) published a graph for non-uniform particle size distributions, where the original Durand and Condolios (1952) graph is considered to be for uniform particle size distributions. Figure 12 shows the graph of Condolios and Chapus. Figure 12 also gives a comparison of the F_L value for uniform and non-uniform particle size distributions. The trends of uniform and non-uniform particle size distributions are the same, but non-uniform particle size distributions have, in general a higher F_L value, at the same d_{50}. This results in smaller limit deposit velocities, which makes sense, because a uniform particle size distribution also contains larger particles, resulting in a lower limit deposit velocity.

The Friction Coefficient Λ

Many researchers use the following equation for the contribution of the solids to the pressure losses:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + \Phi \cdot C_{vt}\right)} \tag{20}$$

Some other researchers disconnected the solids effect from the hydraulic gradient i_{fl}, implying that the solids effect is independent from the hydraulic gradient, according to:

$$\mathbf{i_m = i_{fl} + \Phi \cdot C_{vt}} \tag{21}$$

Of course, the formulation of Φ is different in both equations. Because the hydraulic gradient i_{fl} depends strongly on the value of the friction coefficient λ, the formulation of equation (20) also depends strongly on the friction coefficient λ, while the formulation of equation (21) does not.

Because the friction coefficient λ may vary from about 0.01 for large smooth pipes (D_p=1 m) to about 0.03 for small smooth pipes (D_p=0.0254 m), a difference of a factor 3 may occur between both equations when extrapolating from a very small pipe in a laboratory to a large pipe in reality. Because most experiments are carried out in small to medium pipe diameters (D_p=0.0254 m to D_p=0.254 m), this should be taken into consideration. Thus, it is important to know whether the solids effect depends on the friction coefficient λ or not. If it does, a formulation like equation (20) should be used; if it does not, a formulation like equation (21) should be used. The Durand and Condolios (1952) equation in this form will look like:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + 85 \cdot \left(\frac{v_{ls}^2 \cdot C_x^{0.89}}{g \cdot D_p \cdot R_{sd}}\right)^{-3/2} \cdot C_{vt}\right) = i_{fl} + 85 \cdot \frac{\lambda \cdot R_{sd}}{2} \cdot \left(\frac{v_{ls}^2}{g \cdot D_p \cdot R_{sd}}\right)^{2/2} \cdot \left(\frac{g \cdot D_p \cdot R_{sd}}{v_{ls}^2}\right)^{3/2} \cdot \left(\frac{1}{C_x^{0.89}}\right)^{3/2} \cdot C_{vt}} \tag{22}$$

With λ=0.02 for small pipes and R_{sd}=1.65 for sand this gives:

$$\mathbf{i_m = i_{fl} + 1.4 \cdot \left(\frac{g \cdot D_p \cdot R_{sd}}{v_{ls}^2}\right)^{1/2} \cdot \left(\frac{1}{C_x^{0.89}}\right)^{3/2} \cdot C_{vt}} \tag{23}$$

This last equation will give about the same results as the original equation in small pipes; but for large pipes, the results will be different. The value of λ will be smaller like 0.01, resulting in a smaller solids effect.

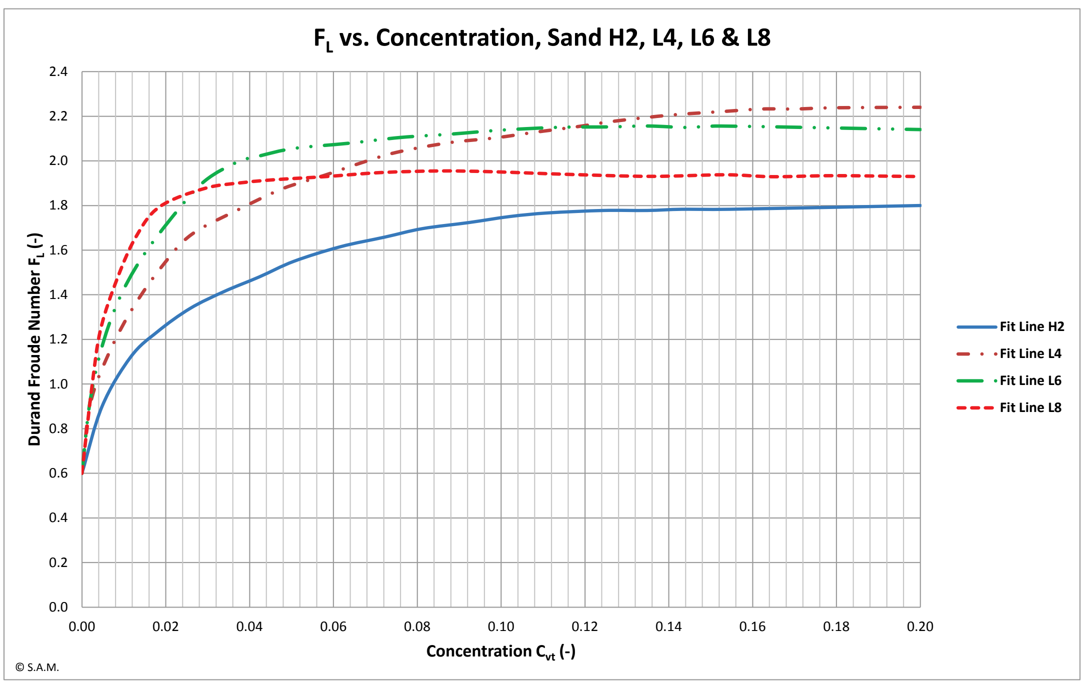

Figure 11. The trend lines of the F_L value as a function of the concentration

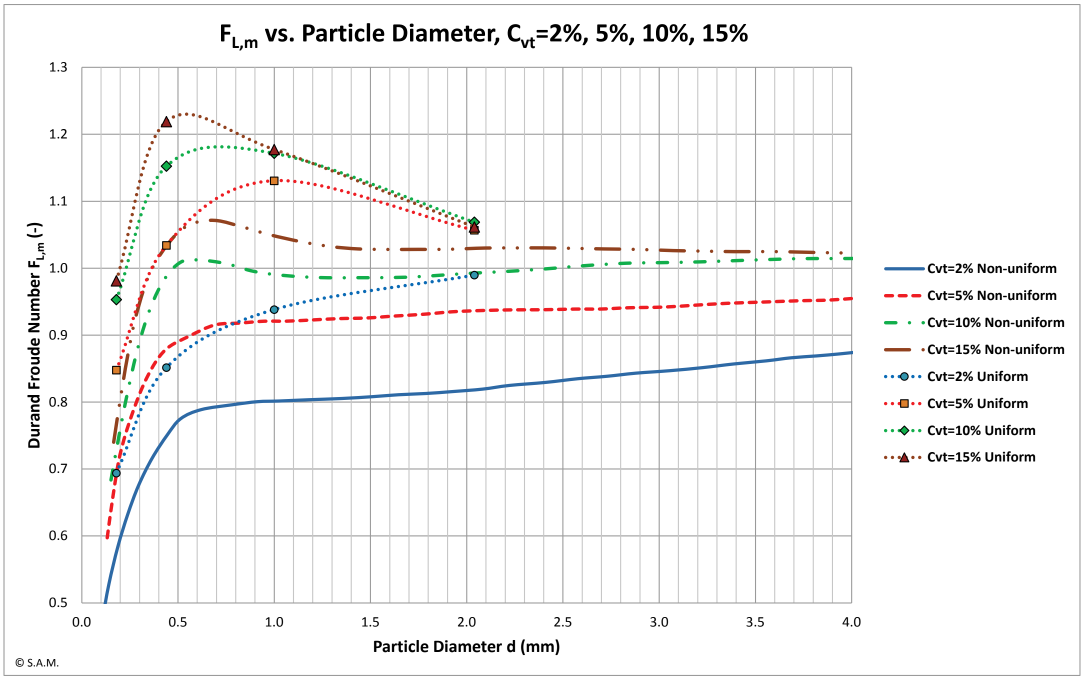

Figure 12. Comparing uniform and non-uniform distributions

The Solids Effect Term in the Pressure Gradient Equation

In both equations (20) and (21), the solids effect is incorporated as one term Φ. This term often consists of a one-term equation, often based on Froude numbers. Now the question is whether the solids effect can be described physically by a one-term equation. It is very well possible that the solids effect depends on a number of different physical phenomena, each with its own term in the equation. Using just one term may force the curve fit equations, as used by most researchers, into a low correlation equation, just because a one-term equation does not describe the processes involved accurately. Using Froude numbers forces the fit equation in a fixed ratio between a number of parameters involved. The flow Froude number forces a fixed ratio between the line speed and the pipe diameter, while the particle Froude number forces a fixed ratio between the terminal settling velocity and the particle diameter. Using an equation for the solids effect with more than one term, without fixing certain ratios, would probably give a better correlation with the experimental data.

THE NEWITT ET AL. (1955) MODEL

Newitt et al. (1955) carried out experiments in a 1-inch pipe with sands of 0.0965 mm, 0.203 mm, and 0.762 mm, and gravel of 4.5 mm. They also carried out experiments with gravel of 3.2 to 6.4 mm, coal of 3.2 to 4.8 mm (R_{sd}=0.4) and MnO2 of 1.6 to 3.2 mm (R_{sd}=3.1). Newitt et al. (1955) distinguished a heterogeneous regime and a sliding bed regime.

The Heterogeneous Regime

For the heterogeneous regime, they assumed that the energy loss due to the solids is based on keeping the particles floating. In other words, due to gravity, the particles will move down continuously and the energy required moving them up, the potential energy, results in an excess pressure loss. Based on the conservation of potential energy of the particles the following equation is derived:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + K_1 \cdot \left(g \cdot D_p \cdot R_{sd} \right) \cdot v_t \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}} \right)^3 \right) \quad with: \quad K_1 = 1100} \qquad (24)$$

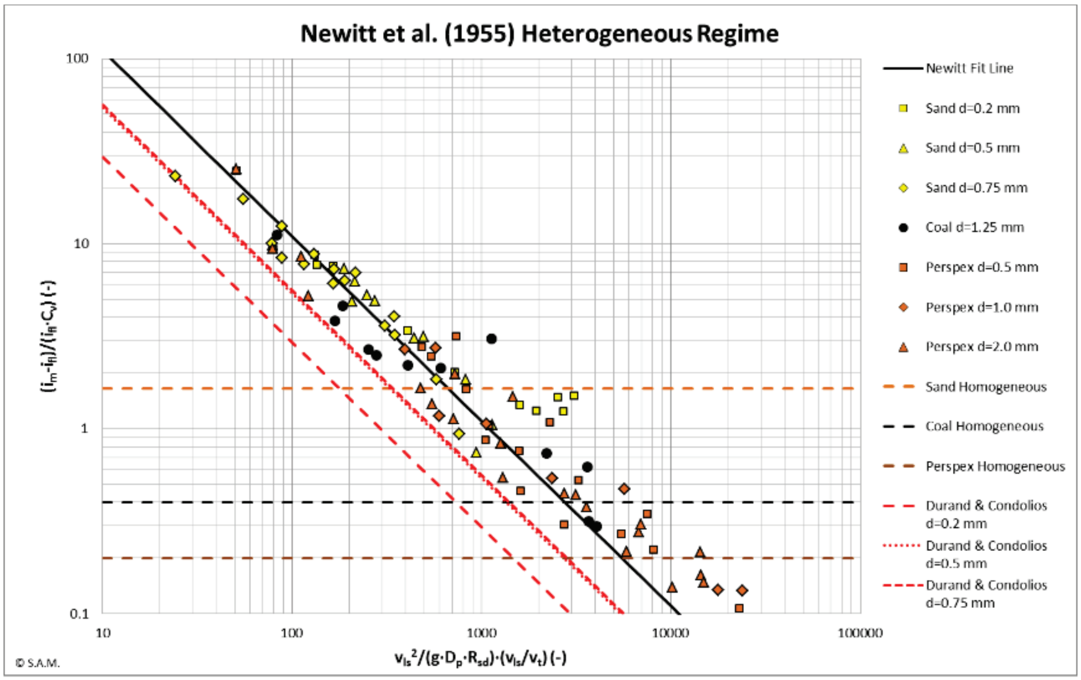

Figure 13. Correlation for particles travelling as a heterogeneous suspension

The coefficient K_1=1100 does not follow from the derivation, but from a best fit of the data points as is shown in Figure 13. A coefficient based on the potential energy derivation would have a value of around 200. Newitt at al. state that the process of keeping the particles in suspension is not very efficient, resulting in a much larger coefficient. A factor 5 to 6 larger would imply an efficiency of 16% to 20%, which is very low. Newitt et al. did not take into consideration the loss of kinetic energy due to the collisions during heterogeneous transport. This would give a second term for the excess pressure losses. The data points follow the Newitt et al. curve reasonably in Figure 13, although a power of the line speed of less than -3 would give a better fit. For sand, 3 Durand and Condolios (1952) curves are drawn. It is clear that the data points of Newitt et al. are all above the Durand and Condolios (1952) curves. It is very well possible that the sheet flow has occurred, due to the high-pressure gradients in such a small pipe (1 inch). For all three materials, some data points are below the equivalent fluid lines, meaning that the pressure gradient is in between the water line and the equivalent fluid line. Now Newitt et al. (1955) carried out their experiments using a 1-inch pipe. In a 1-inch pipe, normally, higher friction coefficients are encountered compared to large pipes as applied in dredging. In a D_p=1 inch pipe a friction coefficient of λ=0.02 is common, while in a D_p=1 m pipe a λ=0.01 would be expected. The difference is a factor 2. Because the Newitt et al. model is based on supplying enough potential energy to keep the particles in suspension, the solids effect should not depend on the viscous fluid friction. In a large diameter pipe with much less fluid friction, the solids effect should be the same as in a small diameter pipe. In order to achieve this, equation (24) will be written in a more general form. Equation (24) in a more general form:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + \left(\frac{\lambda \cdot K_1}{2}\right) \cdot \frac{2 \cdot \left(g \cdot D_p \cdot R_{sd}\right) \cdot v_t}{\lambda} \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}}\right)^3\right) \quad With: \ \lambda{=}0.02 \ \ and \ \ K_1 = 1100 \ \ this\ gives:}$$

$$\mathbf{i_m = i_{fl} \cdot \left(1 + 11 \cdot \frac{2 \cdot \left(g \cdot D_p \cdot R_{sd}\right) \cdot v_t}{\lambda} \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}}\right)^3\right) = \Delta p_f \cdot 1 + \Delta p_f \cdot 11 \cdot \frac{2 \cdot \left(g \cdot D_p \cdot R_{sd}\right) \cdot v_t}{\lambda} \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}}\right)^3} \tag{25}$$

$$\mathbf{i_m = i_{fl} + 11 \cdot \frac{R_{sd} \cdot v_t}{v_{ls}} \cdot C_{vt}}$$

It should be noted that, although the derivation of Newitt et al. model is based on the settling velocity of the particles, hindered settling is not considered. It should also be noted that very small pipe diameters give very high-pressure gradients, often leading to a sliding bed regime or heterogeneous and (pseudo) homogeneous transport. The limit deposit velocity in such a case is based on the transition between a sliding bed and heterogeneous transport. At much larger pipe diameter, with much smaller pressure gradients, the limit deposit velocity is based on the transition between a stationary bed and heterogeneous transport.

The Sliding Bed Regime

For the sliding bed regime Newitt et al. assumed that the weight of all the solids is transferred to the pipe bottom, resulting in a friction force, which is equal to the weight of the solids $\rho_{fl} \cdot g \cdot \mathbf{R_{sd}} \cdot \mathbf{C_{vt}}$ times a friction coefficient μ. They carried out experiments with gravel of 3.2-6.4 mm, coal of 3.2-4.8 mm (R_{sd}=0.4) and MnO2 of 1.6-3.2 mm (R_{sd}=3.1). Figure 14 shows the results of these experiments with a new coordinate on the vertical axis $\left(\mathbf{i_m - i_{fl}}\right)/\left(\mathbf{R_{sd} \cdot C_{vt}}\right)$. The advantage of this parameter is that for a sliding bed it gives the friction coefficient μ directly. Below the limit deposit velocity for a fixed bed, Newitt et al. found:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + K_2 \cdot \left(g \cdot D_p \cdot R_{sd}\right) \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}}\right)^2\right) \quad with: \quad K_2 = 66} \tag{26}$$

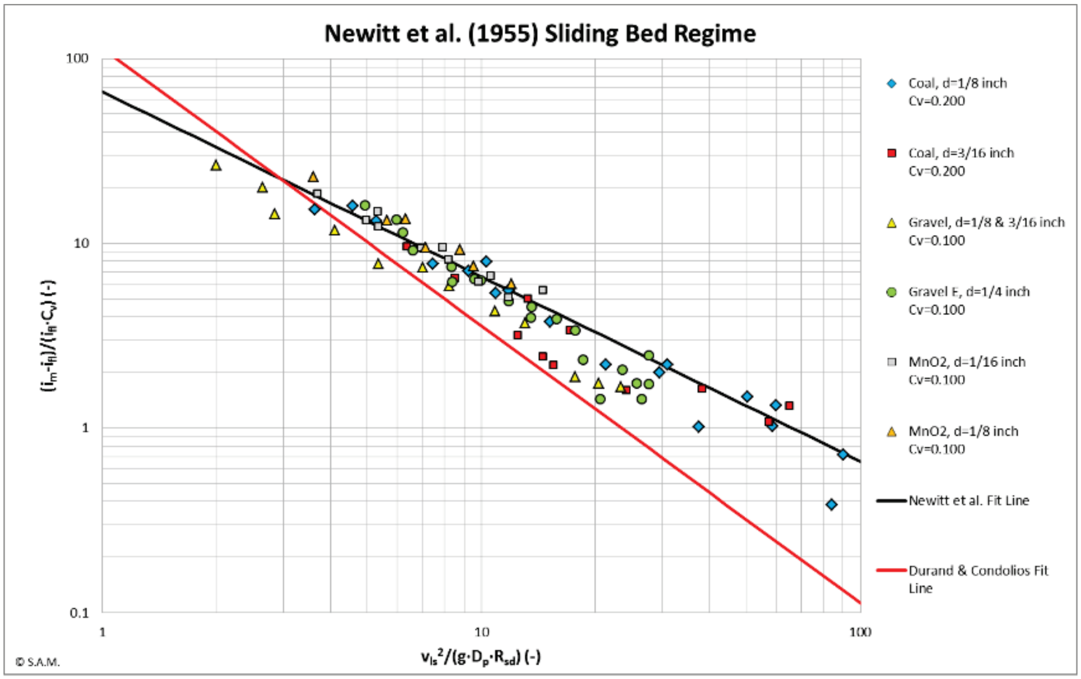

Figure 14. Correlation for large particles travelling in saltation or as a sliding bed

Newitt et al. considered a sliding bed with a friction coefficient of μ=0.8, but an analysis of the data points shows a decreasing tendency with increasing line speed. This matches the constant volumetric transport concentration model, which seems to be applied by Newitt et al. Friction coefficients of 0.35-0.7 have to be used to make the data points match the theory. The different materials have different friction coefficients. A better average of the friction coefficient would be μ=0.66, matching a friction factor λ=0.02 and K_2=66. Because Newitt et al. considered the solids effect to be the result of sliding friction, this solids effect should not depend on the viscous friction, although equation (26) implies this. In a D_p=0.0254 m (1 inch) pipe a friction coefficient of λ=0.02 is common, while in a D_p=1 m pipe a λ=0.01 would be expected. The difference is a factor 2. In a large diameter pipe with much less fluid friction, the solids effect should be the same as in a small diameter pipe. In order to achieve this, equation (26) will be written in a more general form:

$$\mathbf{i_m = i_{fl} \cdot \left(1 + \left(\frac{\lambda \cdot K_2}{2}\right) \cdot \frac{2 \cdot \left(g \cdot D_p \cdot R_{sd}\right)}{\lambda} \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}}\right)^2\right)} \quad \textbf{With: } \lambda\textbf{=0.02 and } \mathbf{K_2 = 66} \textbf{ this gives:} \tag{27}$$

$$\mathbf{i_m = i_{fl} \cdot \left(1 + 0.66 \cdot \frac{2 \cdot \left(g \cdot D_p \cdot R_{sd}\right)}{\lambda} \cdot C_{vt} \cdot \left(\frac{1}{v_{ls}}\right)^2\right) = i_{fl} + 0.66 \cdot R_{sd} \cdot C_{vt}} \tag{28}$$

Note that the second term between the brackets leads to a constant pressure loss independent of the line speed. The friction coefficient of 0.66 of course depends on the type of solids transported. In the original graph of Newitt et al. the Durand and Condolios (1952) curve is incorrect, having the wrong slope (power). Most data points are below the Newitt et al.

approximation for a sliding bed. A line with a steeper slope, so a higher power of the ($1/v_{ls}$) term would give a better fit. This matches the constant volumetric transport concentration behavior.

The Limit Deposit Velocity

The limit deposit velocity is often defined as the velocity below which the first particles start to settle and a bed will be formed at the bottom of the pipe. Often this limit deposit velocity is a bit smaller than the minimum velocity, which is at a pressure of three times the water resistance, based on the derivative of the head loss equation for heterogeneous flow. In Hydraulic Engineering, it is assumed that particles stay in suspension when the so-called friction velocity equals the settling velocity of the particles, giving:

$$\mathbf{u_* \geq v_t} \quad \textbf{At the minimum resistance velocity this gives:}\ \mathbf{u_*^2 = 3 \cdot \frac{\lambda}{8} \cdot v_{ls,\,cr}^2} \tag{29}$$

$$\textbf{With}: \lambda = 0.01 \quad \Rightarrow \quad v_{ls,\,cr} = \sqrt{\frac{8}{3 \cdot \lambda}} \cdot v_t \approx 16.33 \cdot v_t \tag{30}$$

The limit deposit velocity found matches the findings of Newitt et al. (1955). Including the effect of hindered settling, this would result in a decreasing limit deposit velocity with an increasing concentration, according to:

$$v_{ls,\,cr} = \sqrt{\frac{8}{3 \cdot \lambda}} \cdot v_t \cdot \left(1 - C_{vt}\right)^{\beta} \approx 16.33 \cdot v_t \cdot \left(1 - C_{vt}\right)^{\beta} \tag{31}$$

Newitt et al. used the following simple equation for the limit deposit velocity:

$$\frac{i_m - i_{fl}}{i_{fl} \cdot C_{vt}} = 1100 \cdot \frac{g \cdot R_{sd} \cdot D_p}{v_{ls,cr}^2} \cdot \frac{v_t}{v_{ls,cr}} = 66 \cdot \frac{g \cdot R_{sd} \cdot D_p}{v_{ls,cr}^2} \Rightarrow v_{ls,cr} = 16.67 \cdot v_t \tag{32}$$

Newitt et al. (1955) assume that the transition between a sliding bed/saltation on one hand and a stationary bed on the other hand follow the well-known Durand and Condolios (1952) equation:

$$v_{ls,cr} = F_L \cdot \sqrt{2 \cdot g \cdot D_p \cdot R_{sd}} \tag{33}$$

The factor F_L can be found in the graph published by Durand and Condolios (1952). Newitt et al. used the graph of Durand (1953) with the factor F_L=1.34 for large particles.

The Transition Heterogeneous vs (Pseudo) Homogeneous Transport

Newitt et al. (1955) found that the excess pressure gradient for (pseudo) homogeneous transport is not exactly the water resistance with the mixture density substituted for the water (fluid) density, but about 60% of the extra resistance, giving:

$$\frac{i_m - i_{fl}}{i_{fl} \cdot C_{vt}} = 0.6 \cdot R_{sd} = 1100 \cdot \frac{g \cdot R_{sd} \cdot D_p}{v_{ls,cr}^2} \cdot \frac{v_t}{v_{ls,cr}} \quad \Rightarrow \quad v_{ls,h=h} = \sqrt[3]{1833 \cdot g \cdot D_p \cdot v_t} \tag{34}$$

REGIME DIAGRAMS

Based on the different transition velocities of Newitt et al. and the equation for the terminal settling velocity of Zanke (1977), the regime diagram of Newitt et al. has been reconstructed. Now there are three issues regarding the equations of Newitt et al. The first issue is the issue of the error with the F_L graph of Durand and Condolios (1952). The value for large particles should not be 1.34, but about 1.05 and corrected by a factor of 1.1 according to Gibert (1960). The second issue is that it is the question whether for homogeneous transport 60% of the solids weight should be applied, or the full 100%. Here 100% is applied. The third issue is the construction of the regime graph. The curves found by applying the equations, do not exactly match the curves of Newitt et al., but then in 1955 computers were not yet available. It should be noted that these regime diagrams do not incorporate the influence of the volumetric concentration.

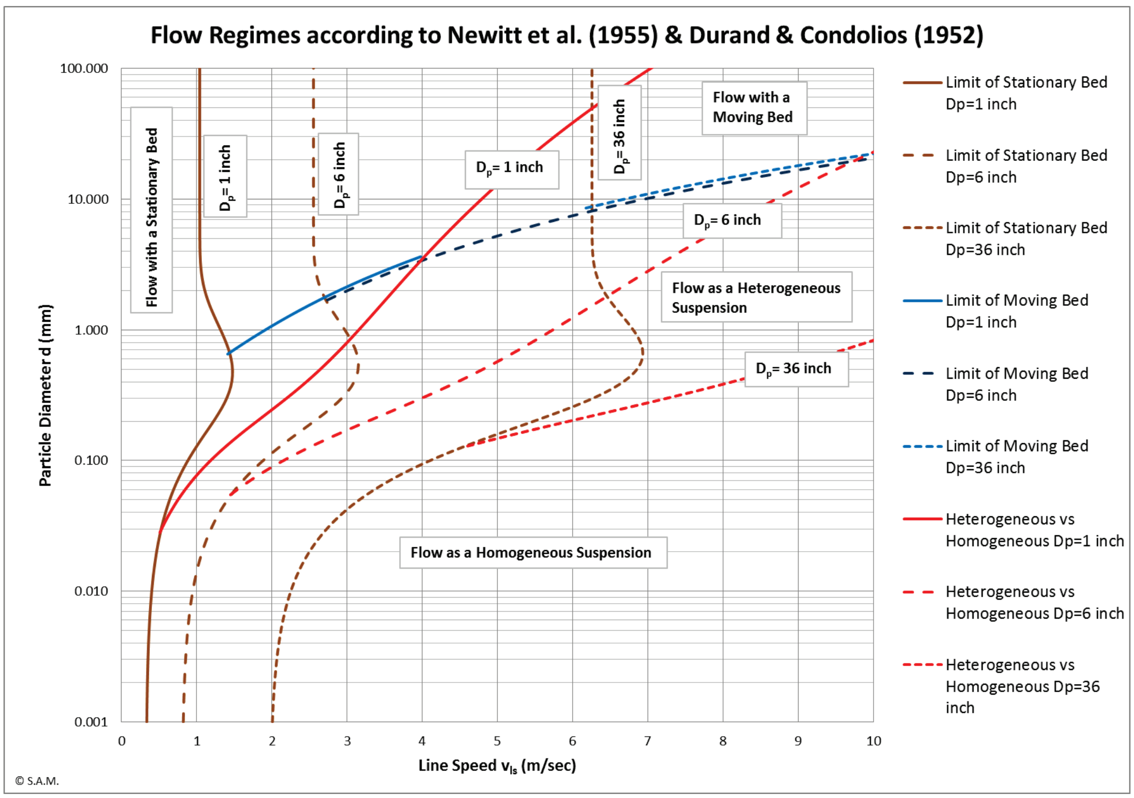

Figure 15. Flow regimes according to Durand and Condolios (1952) and Newitt et al. (1955), modified (Captions for the 36 inch pipe diameter, C_v=0.15)

The regime diagrams of Newitt et al. however give a good impression of the different regimes and the transitions between the different regimes.

- For very small particles, there will be a transition from a stationary bed to homogeneous flow directly.
- For small particles, there will be a transition from a stationary bed to heterogeneous flow to homogeneous flow. For medium sized particles, there will be a transition from a stationary bed to a moving bed to heterogeneous flow to homogeneous flow.
- For very large particles, there will be a transition from a stationary bed to a moving bed to homogeneous flow directly.

Of course, this depends on the pipe diameter and the concentration. Especially the pipe diameter is playing a very big role in the location of the different transition lines. Figure 15 shows the regime diagram based on the equations derived here.

DISCUSSION AND CONCLUSIONS

About 60 years after Durand and Condolios (1952) and Newitt et al. (1955) carried out their research, their results are still valid and important. In spite of criticism of Zandi and Govatos (1967), Babcock (1970), Wilson et al. (1992) and others, their equations are still widely used. There are some issues identified, leading to a wrong interpretation of the equations. The main issues are; the wrong use of the particle Froude number $\sqrt{C_x}$ vs. the drag coefficient C_D, the wrong use of the relative submerged density R_d in the particle Froude number $\sqrt{C_x}$, the wrong power of the particle Froude number $\sqrt{C_x}$ and the use of the wrong graph for the limit deposit velocity coefficient F_L in the Durand and Condolios (1952) equations. For the limit deposit velocity coefficient F_L, the correction factor of about 1.1 according to Gibert (1960) should be applied. For Newitt et al. (1955) it should be considered that the sliding bed equation is based on an average of some specific materials in a very small pipe (1 inch). Other materials and pipes may lead to sliding friction coefficients in the range of μ=0.35-0.7. Combining both theories results in a flow regime chart as is shown here. In this chart, the Newitt et al. equations are used for the transition moving bed-heterogeneous transport and heterogeneous-homogeneous transport. The Durand and Condolios (1952) approach is used for the stationary bed curve. Both Durand and Condolios (1952) and Newitt et al. (1955) consider the excess pressure losses for heterogeneous transport to be reversely proportional to the line speed. Zandi and Govatos (1967) found a power of -1.93 and Wilson et al. (1992) a power of -1.7 for uniform particle size distributions and smaller powers up to -0.25 for non-uniform distributions. This leads to relative excess pressure powers of 3, 3.93, and 3.7. Now heterogeneous transport is dominated by the energy losses due to collisions of particles. If we assume that the occurrence of collisions is dominated by the settling velocity of the particles, then the number of collisions per unit of time is almost independent of time and, thus, of the line speed. This implies that the number of collisions per unit of pipeline length is reversely proportional to the line speed, resulting in a power of -1 of the line speed in the excess pressure losses or -3 in the relative excess pressure losses. If one considers the momentum of the particles in the direction of the line speed, higher powers can be explained. The final conclusion is that the 60-year-old theories can still be applied if one takes the effort to use them properly.

NOMENCLATURE

C_D	Particle drag coefficient	-
C_v	Volumetric concentration	-
C_{vs}	Volumetric spatial concentration	-
C_{vt}	Volumetric transport/delivered concentration	-
C_x	Inverse particle Froude number squared according to Durand and Condolios Fr_p^{-2}	-
d	Particle diameter	m
d_{50}	Particle diameter at which 50% by weight is smaller	m
D_p	Pipe diameter	m
D_H	Hydraulic diameter	m
F_L, $F_{L,m}$	Durand and Condolios limit deposit velocity coefficient	-
Fr_{cr}	Flow Froude number at the limit deposit velocity/critical velocity	-
Fr_{fl}	Flow Froude number	-
Fr_p	Particle Froude number $1/\sqrt{C_x}$	-
g	Gravitational constant	$9.81 \cdot m/sec^2$
i	Pressure gradient	m.w.c./m
i_m	Pressure gradient mixture	m.w.c./m
i_w,i_{fl}	Pressure gradient water/fluid	m.w.c./m
K	Durand and Condolios constant (176-181) or (81-85)	-
K_1	Newitt coefficient for heterogeneous transport (1100)	-
K_2	Newitt coefficient for sliding/moving bed (66)	-
L, ΔL	Length of the pipeline	-
p_{er}	Relative excess pressure	-
Δp	Head loss over a pipeline length ΔL	kPa
Δp_m	Head loss of mixture over a pipeline length ΔL	kPa
Δp_f, Δp_w	Head loss of fluid/water over a pipeline length ΔL	kPa
R_{sd}	Relative submerged density	-
S_{rs}	Slip ratio squared/Stratification ratio solids/Friction coefficient (see also μ)	-
u_*	Friction velocity	m/sec
v_{ls}	Line speed	m/sec
$v_{ls,ld}$	Limit deposit velocity (often called critical velocity)	m/sec
$v_{ls,cr}$	Critical velocity (often the limit deposit velocity)	m/sec

$v_{ls,h\text{-}h}$	Transition velocity heterogeneous vs. homogeneous according to Newitt	m/sec
v_t	Terminal settling velocity of particles	m/sec
β	Power of Richardson and Zaki equation	-
ρ_{fl}	Fluid density	ton/m^3
ρ_w	Density of water	ton/m^3
λ	Moody friction factor	-
μ	Friction coefficient for sliding bed	-
Φ	Durand ordinate, relative excess pressure	-
ψ	Durand abscissa, equations may differ due to historical development, later the relative submerged density has been added, sometimes the particle Froude number is omitted	-
ξ	Particle shape coefficient, usually near 0.7	-
ν_w,ν_{fl}	Kinematic viscosity of water/fluid	m^2/sec

REFERENCES

Babcock, H. A. (1970). *"The sliding bed flow regime."* Hydrotransport 1, pp. H1-1 - H1-16. Bedford, England: BHRA.

Bain, A. G., & Bonnington, S. T. (1970). *"The hydraulic transport of solids by pipeline."* Pergamon Press.

Berg, C. H. (1998). *"Pipelines as Transportation Systems."* Kinderdijk, the Netherlands: European Mining Course Proceedings, IHC-MTI.

Condolios, E., & Chapus, E. E. (1963A). *"Transporting Solid Materials in Pipelines."* Journal of Chemical Engineering, pp. 93-98.

Condolios, E., & Chapus, E. E. (1963B). *"Designing Solids Handling Pipelines Part II."* Journal of Chemical Engineering, pp. 131-138.

Doron, P., & Barnea, D. (1993). *"A three layer model for solid liquid flow in horizontal pipes."* International Journal of Multiphase Flow, Vol. 19, No.6., pp. 1029-1043.

Doron, P., Granica, D., & Barnea, D. (1987). *"Slurry flow in horizontal pipes, experimental and modeling."* International Journal of Multiphase Flow, Vol. 13, No. 4., pp. 535-547.

Durand, R. (1953). *"Basic Relationships of the Transportation of Solids in Pipes - Experimental Research."* Proceedings of the International Association of Hydraulic Research. Minneapolis, MN.

Durand, R., & Condolios, E. (1952). *"Etude experimentale du refoulement des materieaux en conduites en particulier des produits de dragage et des schlamms."* Deuxiemes Journees de l'Hydraulique, pp. 27-55.

Fuhrboter, A. (1961). *"Über die Förderung von Sand-Wasser-Gemischen in Rohrleitungen."* Mitteilungen des Franzius-Instituts, H. 19.

Gibert, R. (1960). *"Transport hydraulique et refoulement des mixtures en conduites."* Annales des Ponts et Chausees, 130(3), pp. 307-374, 130(4), pp. 437-494.

Jufin, A. P., & Lopatin, N. A. (1966). *"O projekte TUiN na gidrotransport zernistych materialov po stalnym truboprovodam."* Gidrotechniceskoe Strojitelstvo 9, pp. 49-52.

Matousek, V. (1997). *"Flow Mechanism of Sand/Water Mixtures in Pipelines, PhD Thesis."* Delft, Netherlands: Delft University of Technology.

Miedema, S. A. (1995). *"Dynamic Pump/Pipeline Behavior Windows."* Software. Delft, The Netherlands: SAM-Consult.

Newitt, D. M., Richardson, M. C., Abbott, M., & Turtle, R. B. (1955). *"Hydraulic conveying of solids in horizontal pipes."* Transactions of the Institution of Chemical Engineers Vo.l 33., 93-110.

Raudviki, A. J. (1990). *"Loose Boundary Hydraulics."* University of Auckland: Pergamon Press.

Richardson, J. F., & Zaki, W. N. (1954). *"Sedimentation & Fluidization: Part I"*. Transactions of the Institution of Chemical Engineering 32, 35-53.

Turian, R. M., & Yuan, T. F. (1977). *"Flow of slurries in pipelines."* AIChE Journal, Vol. 23., 232-243.

Wasp, E. J., Kenny, J. P., Aude, T. C., Seiter, R. H., & Jacques, R. B. (1970). *"Deposition velocities transition velocities and spatial distribution of solids in slurry pipelines."* Hydro Transport 1, paper H42. (pp. 53-76). Coventry: BHRA Fluid Engineering.

Wilson, K. C., Addie, G. R., & Clift, R. (1992). *"Slurry Transport using Centrifugal Pumps."* New York: Elsevier Applied Sciences.

Wilson, K. C., Addie, G. R., Clift, R., & Sellgren, A. (1997). *"Slurry Transport using Centrifugal Pumps."* Glasgow, UK.: Chapman & Hall, Blackie Academic & Professional.

Worster, R. C., & Denny, D. F. (1955). *"Hydraulic transport of solid materials in pipelines."* Institution of Mechanical Engineers (London), 563-586.

Zandi, I. (1971). *"Hydraulic transport of bulky materials."* In I. Zandi, *"Advances in Solid–Liquid Flow in Pipes and its Applications."* (pp. 1-38). Oxford: Pergamon Pres.

Zandi, I., & Govatos, G. (1967). *"Heterogeneous flow of solids in pipelines."* Proc. ACSE, J. Hydraul. Div.,, 93(HY3)., 145-159.

Zanke, U. C. (1977). *"Berechnung der Sinkgeschwindigkeiten von Sedimenten."* Hannover, Germany: Mitteilungen Des Francius Instituts for Wasserbau, Heft 46, seite 243, Technical University Hannover.

NOTES FOR CONTRIBUTORS

General

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Keywords

Please provide five keywords that are not already contained in the title, <u>on a separate sheet of paper</u>.

MANUSCRIPT PREPARATION

Order of Contents

Title, author(s), affiliations, addresses, countries
Abstract (not to exceed 300 words).
Introduction, main body, and following text, conclusions, nomenclature (if necessary), and references. Five keywords that are not already contained in the title (<u>on a separate sheet of paper</u>).

Refer to a previous issue of the journal for general guidelines on format.

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The text should be submitted on unlined white 8½ x 11 inch paper with **single line spacing**, and top and side margins of 1 inch. Use full justification. **The image area or block of text will then be 6.5 x 9.0 inch.** The bottom margin should be 1½ inch. <u>Page numbers should be marked in pencil and placed at the bottom center of each page.</u> **Do not leave additional margins. Do not use company letterhead paper.**

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Equations

All symbols must be defined in the nomenclature section that follows the conclusions. The SI system of units should be used. If units other than SI units are included, they should be given in parenthesis after the relevant SI unit. Equations should be successively numbered (in parenthesis) flush with the right-hand margin (see example below).

$$y = a + b + cx^2 \qquad (1)$$

References

References in the text should be given as follows: Smith (1988), (Smith, 1988) or (Jones et al., 1986). References should be listed alphabetically in the References section at the end of the paper. Give the names and initials of all authors, followed by the title of the article and publication, the publisher and the year of publication. References to conference papers or proceedings should include the name of the organizers. References to articles published in journals should also include the name of the journal, the number of the issue and page numbers (see example below). References to publications in a foreign language should give all details in the original language followed by a translation of the title.

Hunt, J.B. (1995). *"Environmental Dredging"*. Smith & Son, Inc., New York, NY.

Donegan, T.M., and Dinicola, W.J. (1986). *"Turbidity Associated With Dredging Operations." Technical* Report, XYZ Consultants, Inc., Baltimore, MD., 60 p.

Jones, F., Doe, A., Hart, E.J.E., and Next, J.P.J. (1986). *"The Design of Dredged Material Disposal Sites."* Proceedings XIVth World Dredging Congress, CEDA, Amsterdam, The Netherlands, pp. 350-368.

White, F.K. and J.M. Jones (1991). *"The Analysis of Flow Fields Around Dragheads."* Journal of Waterway, Port, Coastal and Ocean Engineering, ASCE, Vol. 121, No. 5, pp. 1-16.

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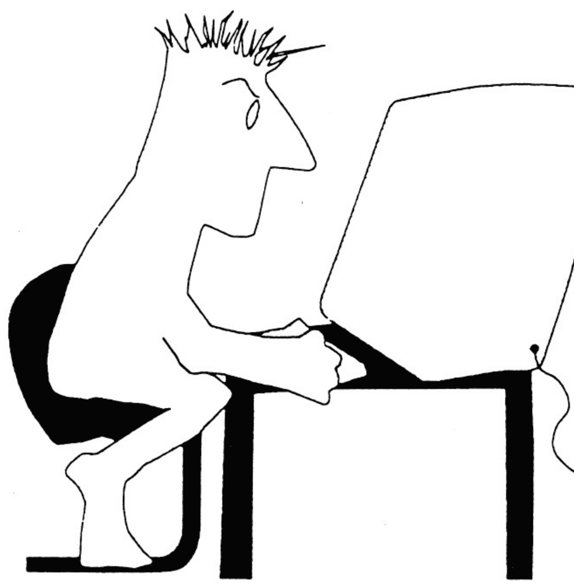

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