November 8th - 11th 2016 Fuzhou Empark Exhibition Grand Hotel, Fuzhou China

Paper #11

SEDIMENTATION IN PIPELINES & COUNTERING BLOCKING

Tony Cooke

BM&T Pty Ltd, (Consultants to the Fibre Cement Industry),
169 Elms Road, Lade Vale, NSW 2581, Australia

tonycookeinoz@gmail.com

ABSTRACT

Blocking of pipelines and reduction in the availability of water returned from water clarifications systems is frequently a problem for manufacturers of fibre cement using the Hatschek process.

This paper examines the theory behind settling slurries. Using theory established for settling slurries used in slurry pipelines conveying a wide range of solids, it determines for typical particles found in fibre cement water systems, the range of critical velocities in the usual sizes of pipelines to maintain suspension of those particles. From these velocities, an estimate is made of the pressure drops per unit length of pipeline.

It then considers the changes likely due to sedimentation leading to a reduction in the effective diameter of the pipelines and the change in roughness of the pipe surface due to the build-up of fibre cement material.

The paper then considers design and other measures that need to be taken to ensure the continuous running of such systems without blocking or significant reductions in the availability of water from the water clarification systems. Such means include avoidance of long horizontal pipelines, using smooth open bends, avoidance of static regions in the outlet of cone tanks and similar.

The paper concludes that each specific design and implementation of each design is important to avoid blocking and sedimentation within the equipment. The designer must therefore take great care in the design and give specific instruction to the implementation of the design in critical areas.

KEYWORDS:
Settling slurries, critical velocity, blocking, pipeline design

Introduction

It is common in all fibre cement plants for distribution pipes to block. In new plants it is also very common for difficulties to be experienced in recovering sufficient water from the recuperation system to run the plant particularly as the plant comes to its design capacity. For example, it is usually easy to run the plant with 1, 2 or 3 vats but it is frequently difficult to bring the 4th and subsequent vats on line because of lack of sufficient dilution water. This paper examines the reasons why this may be so and suggests solutions to these problems. It is important to note however, that because of local conditions and local implementation of even proven and repeated designs, every plant is different and it is often necessary to undertake specific investigations to resolve local problems.

This paper first establishes the flow conditions for maintaining solids in suspension under regimes that may be expected to exist when a plant first comes on line and more generally after a maintenance stoppage when the components of the plant may be considered to be relatively clean. We then determine the flow rates and likely driving pressures in a typical delivery system and determine if these are adequate for the circumstances. We then consider the situation in stoppages where the pipelines may partially block or are progressively roughened due to the deposition of cementitious films inside the pipes. We also consider situations that may be induced by valves, bends etc. Finally, we make some suggestions for solutions to blocking problems and for future work to experimentally evaluate the suggestions made here.

Head Loss in Transport of Solids in Slurry Pipelines

Transport of solids in slurry form has been common for many years[1] and the seminal work in the area was published by Newitt, Richardson, Abbot and Turtle in 1955 (reference 1). Newitt et al., investigated the flow of slurries composed of various materials, concentrations and sizes at different flow rates through a 1” (25mm) diameter pipe.

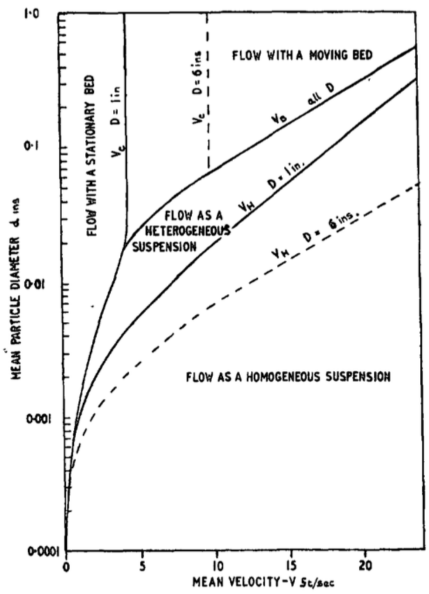

Figure 1: Flow characteristics of slurries

They observed and considered two types of flow – homogenous flow where the particles are of such a size that they do not settle out in turbulent conditions and heterogeneous flow where coarser particles may settle out forming a concentration gradient from the top to bottom of the pipe.

In homogenous flow all the particles are transported fully suspended but in heterogeneous flow all the particles may be suspended at sufficiently high velocity of the fluid although they may travel at a lesser speed than the liquid. At lower velocities particles tend to collect at the bottom of the pipe forming a stationary bed. At slightly higher velocities the bed itself may be pulled along the bottom of the pipe at a much slower velocity than the flow of the liquid above it or the bed remains stationary and some particles may be transported by saltation i.e.

[1] As quoted in Reference 1, theoretical and practical investigations started with the work of Hazen and Hardy in 1906.

essentially bouncing along the top of the stationary or quasi stationary bed. This is illustrated in Figure 1 taken from Reference 1.

It will be noted in Figure 1, that the flow regimes are determined by the velocity of flow, the particle diameter and the size of pipe through which they flow. Newitt et al also noted that heterogeneous flow can be expected with particles larger than 30μm.

Newitt et al. were concerned to determine the friction factors associated with the flow of slurries so that the power requirements for slurry pipelines could be estimated and pipelines could be designed and satisfactorily operated. They therefore derived theoretical equations to determine friction factors and correlated these with their observations. Most of their studies concerned uniform size roughly cuboid particle slurries but they also included flow of a sand slurry composed of two different particle sizes. They observed that some particles in some slurries were prone to break down due to abrasion and therefore slurry properties may change with time during pumping. Newitt et al. did not consider the situation of cement containing slurries that are prone to adhere quite strongly to the surface of the pipes through which they flow. Nor did they consider slurries of particles that also contain flexible fibres.

In the first portion of their paper Newitt et al considered flow of an homogeneous suspension and derived a relationship for the head loss of such suspensions compared to the head loss of water in the same size pipe. They then went on to consider flow of heterogeneous suspensions by considering the turbulent energy required to prevent a typical average particle from settling from the centre to the bottom of the pipe. Finally, they considered the more specific flow of suspensions with moving beds and derived expressions for the head loss of these particles also.

Newitt et al found the following dimensionless expressions for the head loss[2]

Homogeneous Suspensions
$$\frac{h-h_w}{C_v h_w} = (s-1) \qquad 1(a)$$

Heterogeneous Suspensions
$$\frac{h-h_w}{C_v h_w} = K_1(s-1)\frac{w_m g D}{V^3} \qquad 1(b)$$

Flow with a moving bed
$$\frac{h-h_w}{C_v h_w} = K_3(s-1)\frac{gD}{V^2} \qquad 1(c)$$

where h = head loss of suspension
h_w = head loss of water at the same flow rate as the suspension
C_v = volume concentration of solids in the suspension
s = ρs/ρw ρs = density of suspended solids
ρw = density of water
wm = terminal velocity of the settling particles
D = pipe diameter
V = flow velocity
K_1, K_3 = constants of proportionality appropriate to the flow type
g = acceleration due to gravity

[2] The notation used for these expressions has been changed from Newitt et al notation to be consistent with the rest of the paper.

These expressions were tested with slurries of various materials including coal, Perspex, sand and Manganese Dioxide in a system comprising a 1" internal diameter. pipe. Good correlation was found with flow conditions and the dimensionless expressions above.

Prevention of Sedimentation

While the work of Newitt et al. is interesting and useful, it does not address the issue of prevention of sedimentation of slurries. Negro et al. (reference 2) measured the chord lengths of particles in fibre cement slurries using the FBRM instrument and found them to be 40-50μm. This measure is considered to be representative of the free particle diameter and since as was noted by Newitt et al, sedimentation will be readily found with particle sizes of 30 μm, such suspensions will be expected to readily settle out. Indeed, this is the common experience in fibre cement plants. We therefore now address the issue of prevention of sedimentation or alternatively maintenance of suspension of these particles in the process water and from Newitt et al this is clearly achieved by keeping the slurries in pipelines moving at more than a critical velocity.

Kokpinar and Gogus (reference 3) investigated this matter and determined the following theoretical equations and correlated them with their own data and compared them with similar equations derived by other researchers. They first determined that the critical flow velocity (V_c) would be a function of the following parameters

$$V_c = f(\rho_s, d_s, \mu_f, \rho_f, C_v, D, w_m, g) \qquad 2(a)$$

where ρ_s = density of the solid
d_s = diameter of the solids
μ_f = viscosity of the fluid
ρ_f = density of the fluid
C_v = volume fraction of the solid in the fluid
D = diameter of the pipe carrying the suspension,
w_m = Settling velocity of the solids
g = acceleration due to gravity

Non-dimensional grouping of the parameters results in the following expression

$$\frac{V_c}{\sqrt{gD}} = f\left[(s-1), C_v, \frac{\rho_f w_m d_s}{\mu_f}, \frac{d_s}{D}\right] \qquad 2(b)$$

Where $\frac{V_c}{\sqrt{gD}}$ is the Froude number F based on the critical flow velocity, s = specific gravity of the solid particles, $\frac{\rho_f w_m d_s}{\mu_f}$ is the Reynolds number R based on the particle settling velocity and $\frac{d_s}{D}$ is the ratio of solid particle diameter to pipe diameter. In this particular case d_s can be replaced by the average particle size of the solids and in our case we will assume that this is equivalent to the mean chord value of the suspended particles as measured by Negro et al.

Kokpinar and Gogus considered slurry flow data from 6 investigators including their own, who used a variety of materials such as coal, sand, blue plastic, black plastic, fine tuff, coarse tuff, anthracite, polystyrene and PVC. They arrived at the following relationship after correlating all of the data.

$$\frac{V_c}{\sqrt{gD}} = 0.055.\left(\frac{d_s}{D}\right)^{-0.60}.C_v{}^{0.27}.(s-1)^{0.07}.\left(\frac{\rho_f w_m d_s}{\mu_f}\right)^{0.3} \qquad 2(c)$$

We can rearrange this equation to determine $V_c$directly as follows.

$$V_c = 0.055.\sqrt{gD}.\left(\frac{d_s}{D}\right)^{-0.60}.C_v{}^{0.27}.(s-1)^{0.07}.\left(\frac{\rho_f w_m d_s}{\mu_f}\right)^{0.3} \qquad 2(d)$$

Determination of V_c for Fibre Cement Slurries

We need to make some assumptions about the various slurries found in fibre cement plants and as the pipe diameter is significant we need to make further assumptions about likely pipe sizes that will be used. As will be seen later we also need to estimate the flow rates of various slurries in the different parts of Hatschek Machine and its auxiliary equipment. The following discussion is based on these assumptions.

The Hatschek Machine is assumed to operate with the following parameters.

1. Each vat in operation takes in about 25kg/min of dry materials.
2. Feed slurry contains 250g/litre of dry materials.
3. Each vat is fed with approximately 900 litres/min (~50 m^3/hour) of slurry containing 50g/litre of dry solids.
4. Each vat discharges to the water recuperation system about 850 litres/min (~50 m^3/hour) of slurry containing 24g/litre of solids.

Steel pipes used around the Hatschek machine are of nominal diameters 80,100, 125, 150, or 200 mm.

From equation 2 above we need data on solids and slurry density as well as slurry viscosity. We also need to take into account typical operating conditions within the production equipment as normally the temperature is 30°C or thereabouts.

We assume that autoclaved formulations are being used and the average solids density is calculated to be 2.64 g/ml as shown in Table 1.

Table 1: Calculation of Fibre Cement Average Density

	SG	% wt	Vol
Cement	3.18	35.20	11.07
Sand	2.65	52.80	19.92
ATH	2.54	4.00	1.57
Cellulose	1.50	8.00	5.33
		100.00	37.90
Average Density			2.64 g/cc3

As we have already indicated we assume that d_s = 50 μm. This does not take into account the presence of fibres that may significantly change the effective particle size but we will ignore this for this paper.

We have observed that the average settling rate of slurries from the Hatschek Machine cone tank is 300mm/min or 0.005 m/sec.

We assume that the temperature of the process water will be 30°C and at that temperature water has a viscosity of 0.008 Pa.secs. As we have no access to direct measurements of the likely viscosities of the slurries mentioned above we make estimates using Krieger-Dougherty Equation (reference 4)[3] who estimate the viscosity of a slurry using equation 3.

[3] Equation 3 has been expanded to determine the exponent of the bracketed term.

$$\mu_s = \mu_w \left[1 - \frac{C_v}{C_{vm}}\right]^{-3.5} \qquad (3)$$

where μ_s = slurry viscosity
μ_w = viscosity of water
C_v is the slurry volume concentration and
C_{vm} is the maximum slurry volume concentration = 0.65

While this equation is useful for estimating the slurry viscosities of spheroidal solids, because of the presence of fibres in fibre cement slurries, it is probably not very accurate. This is particularly the case for the feed slurry and to a lesser extent the diluted slurry fed to the vat from the homogeniser. It will be more accurate for the dilution water taken from the cone tanks because this contains only very small concentrations of fibre.

Critical Velocity Results

However, given these assumptions above we obtain the following results shown at the right in Table 2.

The table shows results for slurries typical of the solids concentrations of the discharge from the cone tank (24g/litre), feed slurry to the vats (50 g/litre) and for feed slurry to the homogeniser (250g/litre).

The pipe diameters are typical of those used around Hatschek Machines. The critical velocity values are also converted to Critical Volume flow rates in m^3/hour to facilitate the analysis of the results.

Table 2		Solids Concentration g/litre		
	Solids wt	24	50	250
	Solids vol fraction	0.009091	0.018939	0.094697
	Slurry Density pf	1.014909	1.031061	1.155303
	Slurry Viscosity uf	0.008418	0.008902	0.014148
Pipe Φ mm				
Nominal ID	Actual ID	Critical Velocity m/sec		
80	7.8	0.83	1.00	1.39
100	10.2	1.12	1.35	1.88
125	12.8	1.44	1.73	2.41
150	15.4	1.76	2.12	2.95
200	20.3	2.38	2.87	3.99
		Critical Volume Flow Rate m3/hour		
80	7.8	14	17	24
100	10.2	33	40	56
125	12.8	67	80	112
150	15.4	118	142	198
200	20.3	277	333	463

The critical flow rates are also shown graphically in Figure 2.

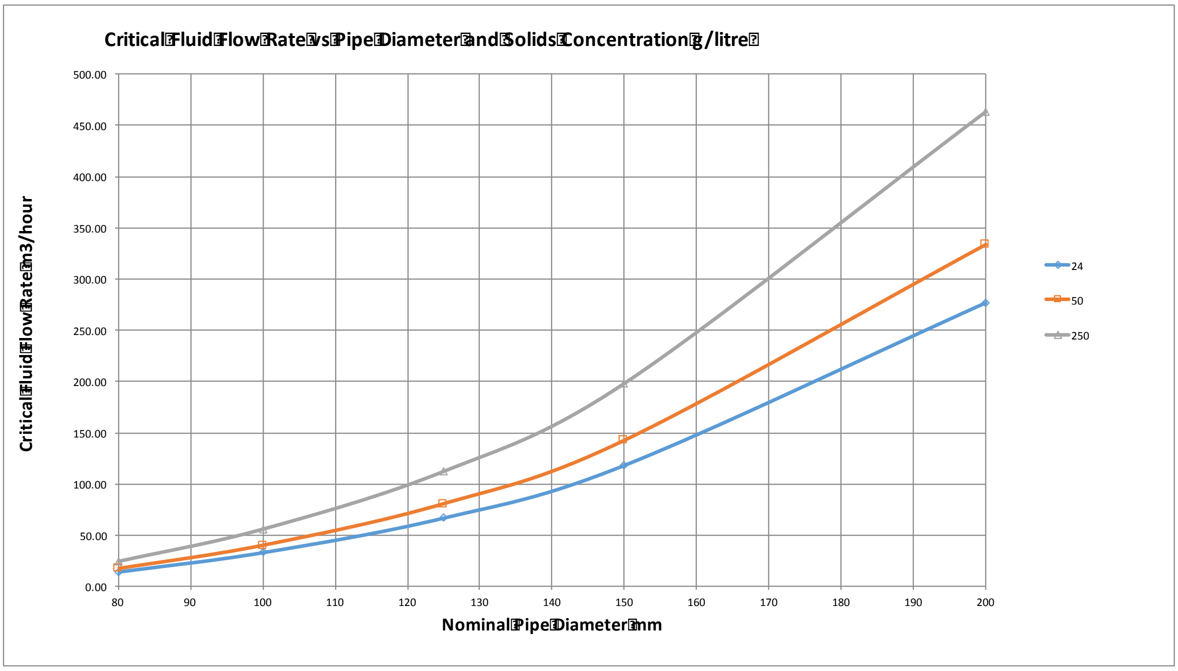

Figure 2: Critical Fluid Flow Rate to avoid sedimentation of Slurries with varying solids contents in Pipes of differing Nominal Diameters.

It can be seen from Table 1 and Figure 2 that the critical velocity for maintenance suspension of solids increases as both the pipe diameter and the concentration of the slurry increases. The relationship for critical velocity is linear in pipe diameter and in solids concentration. When velocity is translated to volume flow rates then the relationship is proportional to the second power which follows because the area of the pipe is proportional to the square of the diameter. The relationship makes sense in terms of the energy required that must be supplied by turbulence to prevent sedimentation.

Required Pipe Size for Maintaining Suspension

We now compare the operation of the Hatschek system under constant and varying production conditions to determine what size pipes are needed to provide sufficient velocity to prevent significant sedimentation in the slurries they contain. Let us start with normal operating conditions in a 4-vat Hatschek machine.

1. Steady Operating Conditions

a) Feed from Stock Chest to Homogeniser - With the assumptions above we find that the feed rate of new material is 100 kg/min and that we will need to deliver to the homogeniser 24 m^3/hour of slurry with a concentration of 250g/litre. We also have to deliver sufficient material for the trim because the trim is normally redissolved and sent back to the stock chest. Allowing for say 20% trim the total requirement of slurry to the homogeniser will be 29 m^3/hour. Reviewing Table 1 for a solids concentration of 250g/litre, we see that a 80 mm nominal diameter pipe will just maintain suspension of solids. In the event of a stoppage or during start-up so we must conclude that there is good chance that this line will block.

b) Feed from the Homogeniser to the Vats – Each vat will take from the homogeniser approximately 50 m^3/hour of slurry with a solids content of 50g/litre. Typically, one 80mm pipe is used to deliver slurry to each vat and it will be seen from table 1 that the critical flow rate for this pipe is 17 m^3/hour. So this line is not likely to block. If two 80mm lines are used for each vat, as in some designs, then blockages will also be unlikely and this is the author's experience.

c) Process Dilution Water from Cone Tank to Homogeniser – As stated above we will need to deliver to the homogeniser about 850 litres per minute of process water from the cone tank for each vat. For a 4 vat machine in operation therefore we need to deliver approximately 205 m^3/hour of process water. From Table 1 we see that a 200 mm pipe is liable to block however a 150 mm pipe should be adequate to provide the necessary volume of fluid.

2. Start-Up Conditions

During start-up, conditions are very different and it is normal after thoroughly wetting the felt to start with one vat at low solids content, progressively increase the solids content and as this happens start the second vat. As these stabilise the third and fourth or greater number of vats are progressively started. At start-up, there will normally be low solids content in all of the flows with the exception of the fresh feed to the machine and the requirements for slurry delivery will also be low. It is clear that in some circumstances the flow rates will be much less than those required to maintain suspension of the solids in the slurries and therefore sedimentation within the pipelines is likely at this time and blocking may well commence just after or during start-up.

3. Unstable Conditions and Stoppages

Clearly there is a great chance of blockages occurring during stoppages than during normal steady production. Pipelines are usually full before stoppages so stoppage of flowwill lead to deposition of solids within the pipes. As we have demonstrated above it is necessary to maintain minimum flow rates to avoid and it is general practice to try to maintain such flows.

4. Closing Comments on Maintaining Suspension of Solids

We have assumed in this analysis that prevention of sedimentation in the pipelines can be modelled by equivalent horizontal pipelines containing slurries of spheroidal particles. While the majority of particles in fibre cement slurries are mainly spheroidal, the fibres are not and they affect both the settling velocities of the reminder of particles and the viscosity of their slurries. So this needs further investigation.

Our assumption of horizontal pipelines is not realistic. Pipelines in fibre cement factories are seldom horizontal and in most practical installations may slope up or down or both, they usually include bends and valves and may even change diameters depending on circumstances. We now wish to extend the analysis to model some simplified systems and to consider the effects of changes within the system during normal operating cycles. As we have said previously it is common to find that the performance of the plant changes during the operating cycle and despite the cone tank being full and overflowing, it may not be possible to withdraw sufficient water from it to operate the plant. We will investigate the causes of this phenomenon and blockages in the next section.

Estimation of Friction Factors and Head Loss for Flows within the Hatschek System

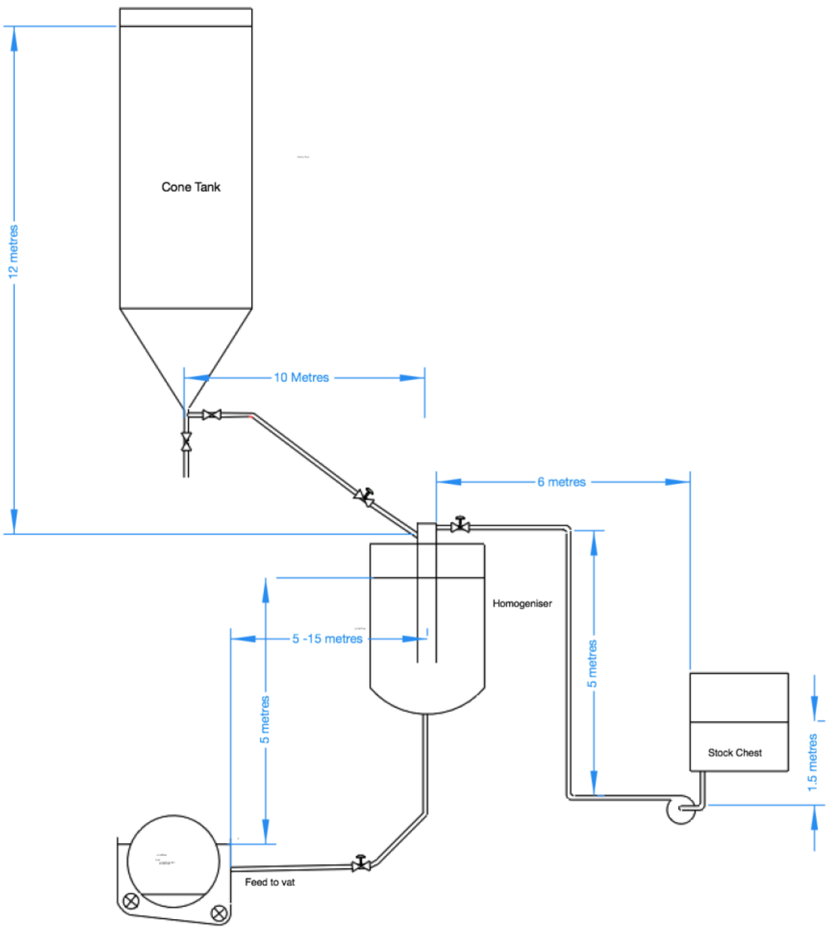

Figure 3: Partial Water System for a Hatschek Machine (hypothetical)

Figure 3 illustrates portions of the water system of a hypothetical Hatschek Machine that we will use to illustrate the analysis of flow. The illustration does not include all of the components of the water circuit or the sand processing, cellulose system and mixer where blockages may also occur. It includes only those components most likely to be susceptible to blocking with fibre cement slurries.

The central component of the system is the homogeniser where fresh feed slurry is diluted with process water from the cone tank before being delivered to the vats of the Hatschek machine. Both feeds to the homogeniser are introduced into the mixing tube in the centre and their flow rates are controlled by control valves. Similarly, the discharge of the homogeniser to each Hatschek Machine vat is also controlled via a control valve. The feed of fresh slurry from the stock chest to the homogeniser is usually effected by a pump but it could be by gravity. Process water is usually delivered from the cone tank by gravity but under some circumstances it could also be delivered by a pump. The same applies to the delivery of diluted feed from the homogeniser to the vats. The actual arrangements in any plant will depend on the designer of the plant and on the physical restrictions of its installation.

Using the assumptions of pipe lengths and devices in Figure 3, we now determine the friction factors and the head loss appropriate to each pipeline and flow rate using commercially available software (see reference 5). We then consider the effects of deposition of solids and changes in the roughness of the internals of the pipes (and other components) on the ability of the system to deliver sufficient slurry to the machine.

1. Cone Tank to Homogeniser

Assumptions: Pipe diameter = 150mm nominal

Roughness = 0.2mm

Pipe length = 13m (2m vertical on discharge +3m horizontal + 7.5m angled to Homogeniser + 0.5m within the mixing tube).

90º Bend and one 45º Bend

One Isolation Valve

1 Control Valve

Pressure Head = 12.5 metres

Starting point flow rate = 50-200m^3/hour = 0.013-0.055m^3/sec.

Putting the first of these values into the spreadsheet gives the results in Figure 4.

1. User Inputs and Preliminary Calculations

Inputs (enter values in the blue boxes)			**Calculations** (done by spreadsheet)		
Pipe Diameter, D_{in} =	154	mm	Pipe Diameter, **D** =	0.1540	m
Pipe Roughness, ε =	5	mm	Cross-Sect. Area, **A** =	0.0186	m^2
Pipe Length, **L** =	13	m	Ave. Velocity, **V** =	0.75	m/s
Pipe Flow Rate, **Q** =	0.01388889	m^3/s	Reynolds number, **Re** =	13,846	
Fluid Density, ρ =	1015	kg/m^3	Friction Factor, **f** =	0.0612	
Fluid Viscosity, μ =	0.008418	$N\text{-}s/m^2$			
Sum of minor loss coefficients, **ΣK** =	4.5				

(set ΣK = 0 to calculate straight pipe losses)

Since Re > 4000, this flow is turbulent and the value of f was obtained from an iterative calculation in cells C91:C100.

2. Calculate the frictional head loss and pressure drop, h_L and ΔP_f

Equations: $h_L = [f(L/D) + \Sigma K](V^2/2g)$ and $\Delta P_f = \rho g h_L$

Frict. Head Loss, h_L =	0.27	m
Frictional Pressure Drop, ΔP_f =	2,726	Pa
Frictional Pressure Drop, ΔP_f =	2.7	kPa

Equipment	Loss Coefficient
Diaphragm Valve, Open	2.3
Diaphragm Valve, Half Open	4.3
Diaphragm Valve, 1/4 Open	21
Water meter	7

Fitting	Loss Coefficient, K
Globe valve, fully open	10.0
Angle valve, fully open	5.0
Swing check valve, fully open	2.5
Gate valve, fully open	0.2
Short-radius elbow	0.9
Medium-radius elbow	0.8
Long-radius elbow	0.6
45 degree elbow	0.4
Close return bend	2.2
Standard tee, flow through run	0.6
Standard tee, flow through branch	1.8
Square entrance	0.5
Exit	1.0

Figure 4: Output from spreadsheet and listing of Loss coefficients

The spreadsheet separately estimates the friction factor firstly by the Churchill Equation[4] and uses the result as the first guess for an iterative calculation using the Colebrook equation[5]. The minor loss coefficients are used to determine the effects of bends, valves etc. In this case we ignore the cone tank and assume that all of the resistance is due to the pipe line, the bends, the isolation valve and the control valve that is assumed to be a diaphragm valve. As will be seen from the results above, the driving pressure head is much greater than the Frictional head loss from the pipe and the bends. So the flow can only be obtained by using the diaphragm control

[4] Churchill Equation for Moody Friction Factor: $f = \left\{-2Log\left[\frac{0.27e}{D}\right] + \left(\frac{7}{Re}\right)^{0.9}\right\}^{-2}$

[5] Colebrook Equation for Moody Friction Factor: $f = \left\{(-2)Log\left[\left(\frac{e}{3.7D}\right) + \left(\frac{2.51}{(Ref^{0.5})}\right)^{0.9}\right]\right\}^{-2}$

valve and we can use the head loss equation to calculate what the head loss coefficient must be for the diaphragm valve. Equation 4 is used to calculate the head loss and we can rearrange it to determine the sum of the head loss coefficients ΣK.

$$h_l = \left[f\left(\frac{L}{D}\right) + \Sigma K\right]\frac{V^2}{2g} \quad \text{or rearranging} \quad \Sigma K = \frac{2gh_l}{V^2} - f\left(\frac{L}{D}\right) \qquad (4)$$

Substituting for the actual driving head pressure of 13m in Equation 4 we find in this instance that total ΣK is 449.1 and the head loss due to the control valve must be 449.1 less 0.9 for the discharge elbow, 0.2 for the isolation valve and 0.4 for the long radius bend giving K = 447.6 for the valve. This implies that the valve must be almost completely closed and is consistent with Kabwe (Reference 6) where he shows that a flow through Saunders valve would be about 6% open to produce this flow rate. We can therefore be confident that this analysis gives a reasonable estimate for the flow conditions. Performing this analysis for the other values of the flow rate corresponding to 1, 2, 3 or 4 vats in operation and using Kabwe's chart to estimate the opening degree of the control valve we obtain the results in Table 3.

Table 3: Estimated valve opening to obtain flow rates in clean pipe

e roughness	0.2 mm			
Flow m^3/hr	50	100	150	200
Flow m^3/sec	0.01388889	0.02777778	0.04166667	0.05555556
v m/sec	0.75	1.49	2.24	2.98
h_l m	0.2	0.8	1.7	3.0
ΣK	449.1	105.6	42.2	20.0
K valve	447.8	104.3	40.9	18.7
Est. % Open	6%	10%	27%	38%

Table 4: Estimated valve opening to obtain flow rates in dirty pipe

e roughness	5 mm			
Flow m^3/hr	50	100	150	200
Flow m^3/sec	0.01388889	0.02777778	0.04166667	0.05555556
v m/sec	0.75	1.49	2.44	2.98
h_l m	0.3	1.1	1.7	4.3
ΣK	402.5	89.5	35.1	11.2
K valve	401.2	88.2	33.8	9.9
Est. % Open	6%	10%	29%	70%

So it is quite feasible to deliver the particular flow rates with the assumed arrangement of pipes and valves.

Blockages are known to occur in this situation that are partly due to the gradual buildup of cement materials roughening the interior of the pipe system and partly to reduction of the overall pipe diameter.

We now consider roughening of the interior of the pipe to a nominal 5mm and we obtain the results in Table 4. Clearly there is little difference in the situation where there is a large buildup of the interior of pipe until we reach the highest flow rates when it becomes necessary to open the control valve from 38% to 70% open.

Considering Table 1, we see that until flow rates of 150 m^3/hour are reached the flow rates are lower than the critical velocity and we can expect that the flow will include moving beds or saltation effects. We now go on to estimate head loss changes due to the additional friction of the sliding bed. Newitt et al have shown that Equation 1(c) applies to both of these situations with K_3 being determined at 66 giving the following equation 5.

$$h = h_l + 66 C_v h_l (s-1)\frac{gD}{V^2} \qquad (5)$$

where h = the head loss of the flow with a sliding or saltating bed and
h_l = the head loss calculated for fully heterogeneous flow as calculated above.

Table 5: comparision of head loss estimates

Flow m^3/hr	50	100	150	200
h_l m (smooth)	0.2	0.8	1.7	3.0
h_l m (rough)	0.3	1.1	1.7	4.3
h m (sliding bed)	1.0	1.8	2.1	5.1

A comparison of the estimates of head loss under the various assumptions are shown in Table 5. It is clear that there is very little difference in the estimates and that the presence of the sliding bed of particles has only a moderately large effect when the desired flow rates are largest. It is also clear that the flow rates are within the range of control by the control valve.

We have not considered the situation where blocking may occur. We now consider what will happen if we also include a reduction in effective diameter of the pipe as well as roughness of the interior. To do this we calculate the hydraulic diameter by assuming that the diameter of the pipe is equivalent to the diameter it would have if its cross-sectional area were reduced. It is easily shown that this is equivalent to the original diameter of the pipe multiplied by the square root of the open fraction of the pipe.

Table 6: Head loss (m) estimates with reduction in pipe diameter due to blocking - assuming rough interior

Flow m^3/hr	50	100	150	200
100% open	0.3	1.1	1.7	4.3
80% open	0.4	1.8	4.0	7.0
60% open	0.9	3.6	8.0	14.2
40% open	2.5	9.7	21.9	38.8

Thus

$$D_{blocked} = \sqrt{F_{open}}\, D_{unblocked}$$

Substituting for $D_{blocked}$ at various levels of blocking we obtain the estimates for the head loss in Table 5. It will be seen that when the blockage exceeds 40%, the head loss exceeds the maximum head loss due to gravity (12.5m) that we have assumed in the hypothetical cone tank/homogeniser system when flow rates of 200 m^3/hour are required. Thus it will not be possible to run the system with 4 vats if blockage of 40% occurs. When blockage of 60% occurs then it will not be possible to run more than 2 vats and the system will most likely become unstable even with 20% blockage. We have shown the infeasible situation with bright yellow highlighting and the unstable situation with dull yellow highlighting.

2. Homogeniser to Vat

Assumptions: Pipe diameter = 80mm nominal

Roughness = 0.2mm

Pipe length = 9 - 18m (4m vertical +5-14m horizontal depending on the vat to which the homogeniser is connected).

2 off 45° Bends

1 Control Valve

Pressure Head = 5 metres

Flow rate = 50 m^3/hour = 0.013m^3/sec.

Firstly, we find that using this size of pipe, the flow velocity will be 2.91m/sec that greatly exceeds the critical flow velocity of slurries of this composition.

Secondly after applying the assumptions we also find the results in Table 7. It is clear that all vats can have slurry delivered from the homogenizer in a controlled manner as the driving pressure head exceeds the head loss by a reasonable margin. There will also be little chance of blocking of the pipelines because the flow velocity is quite high. This as previously noted accords with the author's experience.

Table 7: Head loss Estimate Homogeniser to Vat

Vat Number	1	2	3	4
Length of Pipe m	9	12	15	18
Head Loss m	2.82	3.31	3.81	4.3

3. Stock Chest to Homogeniser

Assumptions: Pipe diameter = 100 mm nominal

Roughness = 0.2mm

Pipe length = 12m (1m vertical + 3m horizontal + 5m vertical +3m horizontal)

2 off 90º Bends

1 Control Valve

Pressure Head = delivered by pump

Flow rate = 7.5 - 30 m^3/hour = 0.0021 - 0.0083m^3/sec.

To estimate the head losses for this situation, we first calculate the frictional head losses due to flow and then add to this the head loss of 3.5m due to the difference in height of the stock chest and the homogenizer entry. The flow velocities except for the greatest flows are less than the critical velocity of 1.37m/sec so that heterogeneous flow with a sliding or saltating bed will result. Thus we go on to calculate the effect of the sliding bed on the frictional losses using Equation 5 and we obtain the estimates in Table 8.

Table 8: Head loss Estimate Homogeniser to Vat

Flow m^3/hr	7.5	15	22.5	30
Velocity m/sec	0.44	0.87	1.31	1.75
h_l m (smooth)	3.57	3.89	4.34	4.95
h m (sliding bed)	3.62	3.96	4.41	N/A

If we further assume that due to the low velocity in the pipeline that there is likely to be blocking, we can estimate the approximate head losses with various degrees of blockage and different required flow rates. Table 9 shows the estimated head losses.

Table 9: Head loss (m) estimates for fresh feed to homogeniser with reduction in pipe diameter due to blocking

Flow m^3/hr	7.5	15	22.5	30
100% open	3.57	3.89	4.34	4.95
80% open	3.62	4.15	4.90	6.38
60% open	3.71	4.76	6.19	9.08
40% open	4.38	6.71	10.40	17.85

We have assumed that the delivery of feed from the stock chest to the homogenizer will be by a pump. It is usual to use a centrifugal pump for this duty and in this instance one would most probably use a 4/3 pump, i.e. 4" or 100mm inlet and 3" or 75 mm outlet. Fitted with an appropriately sized motor, such pumps are easily able to deliver the required amount of slurry.

Blockages and their Avoidance

The above analysis indicates that blockages should be avoidable for the most part with the possible exception of the fresh feed to the homogenizer where the flow velocities are frequently below the critical velocity to avoid settlement of the solids. Of course we are assuming constant production therefore constant movement of slurries through the pipelines and this is not a normal circumstance. Slurry demand even in a constantly operating plant will vary due to recirculation of trim, changes to reject levels and many other factors. We have also assumed somewhat idealized structures within the machinery and we have had to make a number of further assumptions on matters such as viscosity of the slurries and generalisations that may or may not be accurate for particular plants. We have also not considered the effects of stoppages that may be out of the control of the plant operators such as failure of equipment and the like.

It is the author's experience that stoppages often result in blockages that may extend the stoppages. Blockages are commonly found in specific parts of the plant water system and may be associated with specific actions (or lack of them) by the operators. Figure 5 shows our idealized water system and highlights those parts of the system where in the author's experience, blockages are common.

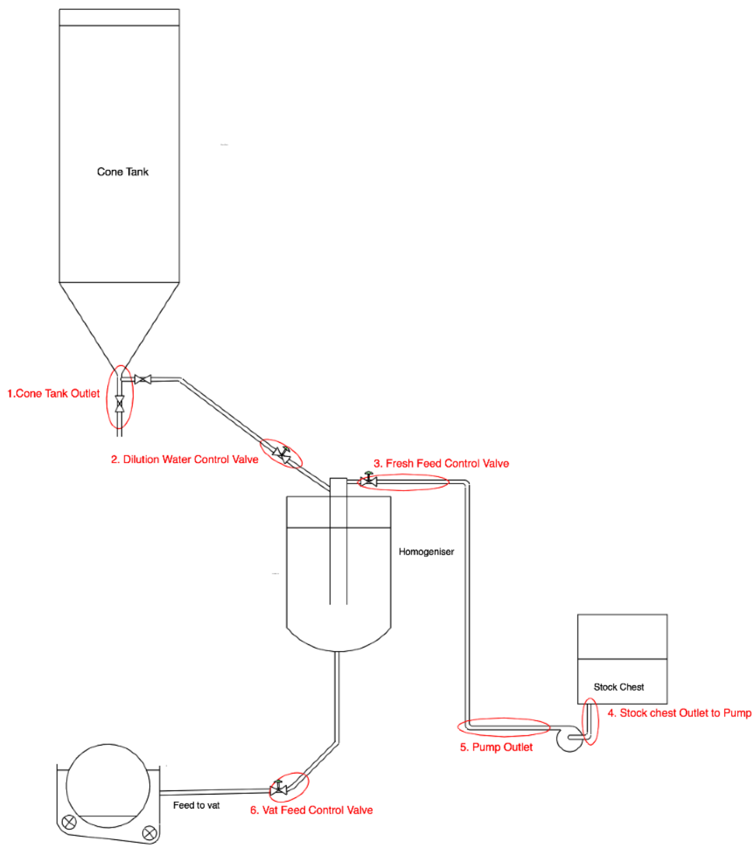

Figure 5: Partial Water System showing blockage points

1. The cone tank outlet can easily be blocked because in a stoppage the cone tank contains a significant amount of solids (mainly cement and mineral fillers) that have been flocculated and will rapidly settle out. A typical cone tank will contain between 60 and 80 m^3 of slurry of which around 80% will have 24g/litre of solids. The remainder will be turned back from the central entry tube and will contain only about 0.5g/litre of solids. So there is a total of about 50 to 60 m^3 of slurry containing between 1.2 and 1.4 tonnes of solids. A cone tank is about 9m high and the free settling rate of solids is about 0.3m/min. So if the cone tank outlet was closed then the last solids would have settled to the outlet in about 27 minutes assuming free settlement of the solids could occur. Although this is a big assumption and settling into the bottom of the cone would be quickly hindered, experience tells us that the solids would rapidly form a bridge across the lower part of the cone tank that would prevent water from exiting the tank when the plant restarts. It is therefore important to keep the cone tank water moving through the system

to avoid blockages at its exit and it would seem that a stoppage here could cause a blockage in less than ½ hour if the water is allowed to remain stationary.

This also raises another issue. The discharge velocity of the normally operating cone tank will increase as the diameter of the cone decreases towards the exit pipe. At some point in the cone the flow velocity will be lower than the critical velocity to keep the solids in suspension and a ring of solids will tend to build up from this point. For this reason, many cone tanks are fitted with rotating scrapers to break up the rings. These scrapers are usually implemented with a rotating beam that drags a heavy chain at each end extending almost to the bottom of the cone. Where scrapers are not fitted, it is not unknown to have a cone tank fill with solids with the exception of a hole running to the bottom. This of course is a prelude to having to stop the plant.

Figure 5 is a schematic of the normal situation at the discharge from the first cone tank with a discharge port closed by a gate valve in a vertical pipe at the bottom of the tank to allow emptying. The connection to the homogenizer is by a second pipe at right angles to the discharge ppipe and is fitted with an isolation valve so that the pipe and the homogenizer can be removed or repaired without emptying the cone tank. If the vertical pipe holding the gate valve extends very far from the discharge to the homogenizer, then solids will accumulate in the vertical pipe to the level of the discharge to the homogenizer and it is common to have a blockage that may prevent emptying of the cone tank. The detail of this section of the tank is very important for successful operation of the system and it is not usual for designers to specify exactly how the connections must be made in terms of the separations of the valves, bends etc.

2. The control valve in the line from the cone tank to the homogeniser can also be a source of blockages. The valve selection can have a big influence on this and a valve that causes a big change in direction of the flow is more likely to block than a straight through valve.

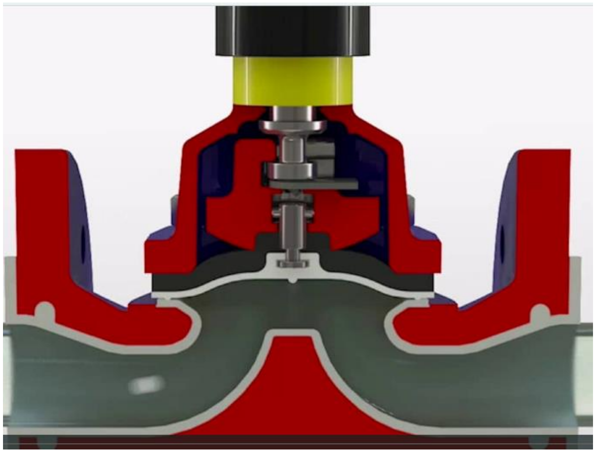

Figure 6(a) Traditional Saunders Valve

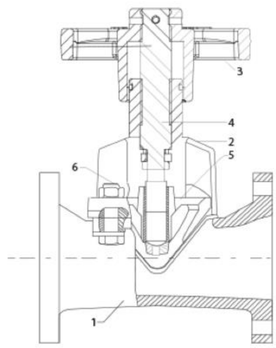

Figure 6(b) Straight Through Saunders Valve

Saunders valve are often chosen to control these flows and there are two types of this valve design that can be used. Manual versions of these valves are shown in Figure 6. These valves are usually fitted with actuators when they are used as control valves. I expect that straight through styles of valves are less likely to cause problems.

3. The control valve for the fresh feed input to the homogeniser is also likely to exhibit the same problems as the control valve of the cone tank water. In addition, the control valve as drawn is set in a horizontal portion of the pipe and as the flow rates in this section of the pipe

are frequently less than the critical velocity to maintain suspension of the solids, there will be a sliding or saltating bed of materials in this section of the pipe. The valve in Figure 5(a) will pose a significant barrier to the movement of this layer and is therefore likely to result in blockages. The straight through type of valve does not have this problem to the same extent and is preferable in this situation.

4. The discharge of the stock chest to the feed delivery pump is often problematical and it is particularly important that the geometry of the connection is smooth. If the pump is stopped then sedimentation of the solids within the pump can cause blockage of the pump if the pump is not restarted in a relatively short time.

5. The discharge of the feed delivery pump is shown in this case to be horizontal and to enter a vertical section through a small radius bend. This is also a potential blockage region for the same reasons as described in paragraphs 3 and 4 immediately above.

6. As stated in the analysis section there is little chance of blockage in the vat feed area. Apart from the favourable delivery of vat feed from the homogeniser, when a stoppage occurs, the vats are often dumped into the backwater channel and this allows the input pipes to drain. Having said that the most likely problems will occur around the control valves and their selection is quite important.

Final Remarks and Conclusions

Analytical Methods – It has been necessary in this paper to make a series of assumptions as to the properties of the slurries used in fibre cement manufacture viz., viscosity of the suspensions and the size of the particles within them. This is particularly problematical because of the fibres present that can be entangled and have long range effects on other particles. The particles are also chemically reactive and stick readily to each other and to the vessels containing them. This is further complicated by the presence of flocculants in parts of the systems that affect the adhesion of the particles to each other (thus increasing the particle sizes) and the effects of their adhesion to the interior of pipes and vessels.

The analysis used has also been taken from investigations and reports where non-reactive slurries of high volumetric concentration appropriate for transport of finely divided solids in water. So it is not certain that the analysis can be properly used in this case. However, the general principles are convincing and the values determined in this analysis are reasonable.

Analysis and Design of Real Systems – It is clear that the analysis of real systems requires very careful consideration of the parameters of the slurries within different parts of the plant. It is necessary to know the changes that take place within the equipment and to take these into account in the analysis.

It is also important to carefully consider the selection of equipment and layout. It is common practice because of different circumstances in different plants that designers only specify the particular valves and other components to be used but do not specify in detail how they are to be connected. Small details are very important to avoid blocking and because it is not easy to determine exactly the optimum implementation of designs, it is often the case that apparently good designs will not perform well because of poor installation.

The operation of systems also is important for reduction of blocking and most blocking problems occur during stoppages due to allowing systems to remain stagnant. So operating experience and skill are also important.

Research in Fibre Cement Slurry Systems – The author was stimulated to undertake this research after a visit to a client where behaviour of slurry systems, mixing and design of homogenisers and delivery of slurry to vats were matters under consideration. It soon became clear that although there has been considerable work on slurry transport systems, there has been nothing published on fibre cement, if indeed there has been anything done systematically. It is the author’s suggestion that there is a good opportunity for original research in this area.

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