THEORY AND PRACTICE OF FILTRATION IN THE ALUMINA INDUSTRY

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Abstract

Continuous rotary filters like disc, drum or pan filters are of great importance in the industrial practice of solid/liquid separation and especially in alumina refineries they are standard technology for a variety of filtration duties. In the daily practice however the knowledge of the theory of cake forming filtration, which is the precondition for understanding the operation behavior of these filters, is rather basic. Essential interrelations of the physical conditions of the filtration process, product characteristics and filter design are insufficiently understood and consequently the operation behavior of running filters is often misinterpreted. Operation issues like incomplete cake discharge on disc or drum filters, increased Na2O-content in the filter cake of pan or drum filters or lack of capacity are often referred to OEMs in terms of "your filter is not performing". What is the reason for incomplete cake discharge on disc or drum filters? Why do results of cake wash fall off in quality on pan or drum filters? A correct analysis and adequate answers to such operational problems are not possible without knowledge the basic functions and interrelations described by the filtration theory.

1. Introduction

Rotary filters are already being used since decades. Further development and innovations concerning the filter design as well as improvements concerning performance and operation have been made within the last years by introducing the BOKELA rotary filters (see fig. 1). The most important feature of these highly efficient filters is the precisely calculated and cleverly designed hydraulic system comprising all filtrate and gas conveying filter parts such as the filter segments or filter cells, the filtrate manifold pipes, the control head and the filtrate receiver. The quality of the hydraulic system is decisive upon the quality of the filtration result and the operating performance.

Figure 1: BOKELA rotary filters: Boozer vacuum disc filter (left), drum filter (mid) and pan filter (right)

The highest percentage of rotary filters is made up by the category of vacuum filters and so the maximum pressure difference available ranges from 0.8 to 0.9 bar, whereas for the pressure rotary filters a pressure difference between 2 and 6 bar can be realized. The typical sizes for drum filters are from 0.2 m^2 up to 100 m^2, whereas the disc filters are built with a filtering surface from 3 m^2 up to more than 200 m^2. Rotary pressure filters are smaller due to higher surface specific performance and they have filtering surfaces of up to 50 m^2 for drum filters and up to 168 m^2 for disc filters.

Even if the separation characteristics of solid/liquid mixtures can be described theoretically the most important parameters of the solid/liquid separation technique must still be determined by carrying out filtration tests. Thus, all filtration parameters required for further reflection are determined by either measuring the set-up and results (throughput, cake height, rotational speed, etc.) at running filters or even better with the aid of an appropriate laboratory filter. A laboratory filtration equipment like the FILTRATEST presented in figure 2 enables quick tests providing results appropriate for scale up calculations. Conventional lab filter cells (made of porcelain) with filter papers usually neither guarantee an even and unhindered filtrate conveyance nor enable the use of filter cloths. For this reason it is strongly recommended not to use such an apparatus for scaling up a filter. However, several tests like the examination of the cloth's lifetime or the cake discharge are recommended to be carried out at running industrial filters only.

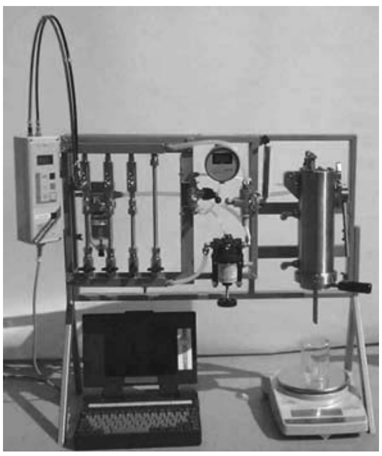


Figure 2: The laboratory pressure filter cell FILTRATEST

Basic formulas for the cake forming filtration

The following formulas and diagrams apply to continuous drum and disc filters and can also be employed for belt and pan filters.

The basic functional interrelations concerning the steady flow through a bulk are shown by the Darcy equation (equation 1).

$$\dot{V}_L = \frac{A_F \cdot \Delta p}{\eta_L \left(r_C \cdot h_C + R_M\right)} \quad \text{short:} \quad \boxed{\dot{V}_L = \frac{A_F \cdot \Delta p}{\eta_L * R}} \qquad (1)$$

This formula generally applies to the flow through already built filter cakes. An example for such a case is the filter cake washing. In order to describe the filter cake formation the Darcy equation is integrated to finally obtain the standard form of the cake formation equation (equation 2).

$$h_C = \sqrt{\frac{2 \cdot \kappa}{\eta \cdot r_C}} \cdot \sqrt{\Delta p} \cdot \sqrt{t_1} \quad \text{short:} \quad \boxed{h_C = C \cdot \sqrt{\Delta p} \cdot \sqrt{t_1}} \qquad (2)$$

This formula (2) describes the filter cake formation with a height of hc under two simplified preconditions: The filter cake is incompressible and the filter cloth resistance RM is much smaller than the cake resistance rC. These assumptions are valid for most of all practical cases so that it is not necessary to determine the real filter cloth resistance RM here. In this context, an incompressible bulk means that the cake does not shrink during the filtration process. This can easily be proven by carrying out 3-4 tests at constant pressure and different filtration times. In case of incompressible cakes the cake height hC versus the square root of the cake formation time t1 forms a straight line in the diagram (see fig. 3).

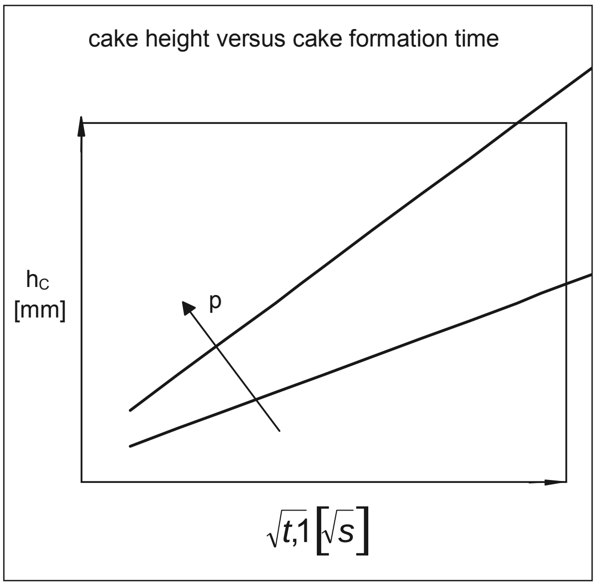


Figure 3: Interrelationship between cake height hC and filtration time t1 in case of incompressibility

Note: If the filtration graph hC above square root of filtration time t1 forms a straight line, it is allowed and very easy to calculate filtration times for other cake heights with this graph!

In regard to rotary filters, the cake formation time t1 can be displayed in dependency on the cake formation angle a1 and the rotational filter speed n according to equation 3 and can further be put in the cake formation equation (equation 2).

$$t_1 = \frac{\alpha_1}{n \cdot 360°} \qquad (3)$$

The cake formation angle a1 is mostly not known and unfortunately only mentioned in very few operation manuals. For this reason it has to be determined by directly measuring the angle at the control head (must be demounted) on one hand and the immersion depth inside the filter trough on the other hand. Assuming that the suspensions' behavior does not change, which means that the physical parameters like viscosity h, the concentration parameter k (definition see appendix) and the flow resistance of the filter cake rC remain constant, equation 2 is being simplified. The simplified dependence of the cake height on the parameters Dp, a1 and n, which can be set at the filter, results from inverting equation 3 and is shown in equation 4.

In case the particle system changes (viscosity, particle size, concentration), new tests are urgently recommended in order to experimentally re-determine the proportionality of equation 2 and 4 respectively the new gradients in figure 3. It is not recommended to determine the new proportions by calculation.

Equation 4 shows that the cake formation is proportional to the square root of the pressure difference Dp and the cake forming angle a1 and also reversed proportional to the square root of the rotational speed n of the filter. These actuating variables are thus setting parameters for the filter operation.

To be applied at the filter: $\boxed{h_C \; proportional \sqrt{\Delta p} \cdot \sqrt{\frac{\alpha_1}{n}}}$

To be applied in the laboratory: $\boxed{h_C \;\; proportional \sqrt{\Delta p} \cdot \sqrt{t_1}}$ (4)

The filter surface specific solids mass throughput is calculated with equation 5. The term is equal to the bulk density of the cake and can be determined by dividing the weight of the dry cake (determined by the lab test) by the cake volume if required. In case this is to laborious, a porosity value of e = 0.5 (first approximation) can be put in as the most filter cakes have a porosity between 0.45 and 0.55. In case of finest particles or inner-porous solids there are as well values between 0.7 and 0.9.

$$\dot{m}_S = n \cdot h_C \cdot \rho_S \cdot (1 - \varepsilon) \qquad (5)$$

If equation 5 is put in equation 4, a simple but fundamental interrelation concerning the solids flow of a continuous rotary filter will result from that.

$$\dot{m}_S \; proportional \; \sqrt{n} \cdot \sqrt{\Delta p} \cdot \sqrt{\alpha_1} \qquad (6)$$

The most important control parameter for rotary filters is the rotational filter speed n. The following mnemonic sentence can be derived from equations 4 and 5:

Note: quadruple filter speed = half cake height (see equation 4) = double solids throughput – or: filter speed is the throttle of filtration (see equation 5 and figure 4)

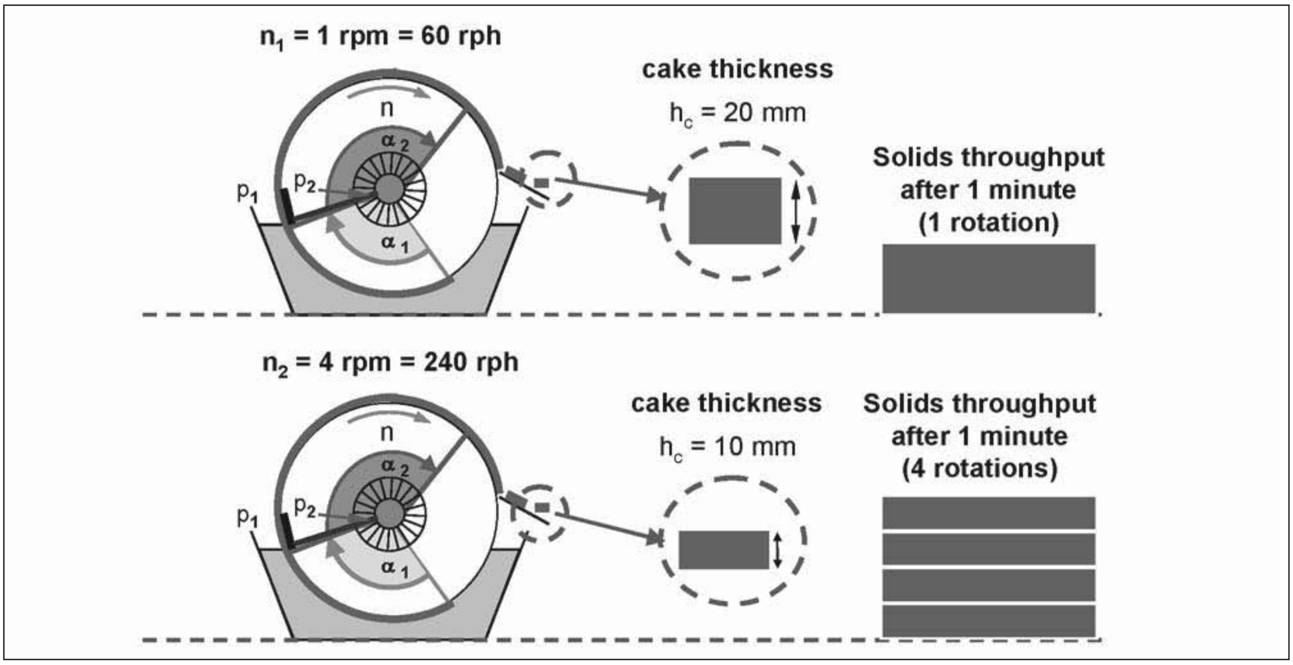


Figure 4: Filter speed is the throttle of filtration

The interrelation between cake height, rotational filter speed and solids flow is qualitatively displayed in the diagram of figure 5. The performance limit of a rotary filter is determined by the smallest cake height still dischargeable. The cakes of convenient filters from older generations are often insufficiently removed from the filter cloth at hC < 10 - 15 mm.

Reconstruction and improvement of the cake removal system is thus one of the most important improving parameters for optimizing rotary filters.

The formulas mentioned can also be applied to discontinuous filters (e.g. pressure candle filters). For the scale up of a discontinuous filter the specific solids mass flow ms has to be determined by summing up all times (cake forming-, washing-, dewatering-, load- and unload time) to one total cycle time instead of the cake forming angle a1 and the rotational filter speed n (equation 3 and 4). However, these formulas are not very useful for filter presses with compressible cakes, which means that the specific flow resistance is continuously changing. In this case the cake height does not directly depend on the filtration time but is rather pre-determined by the chamber's height.

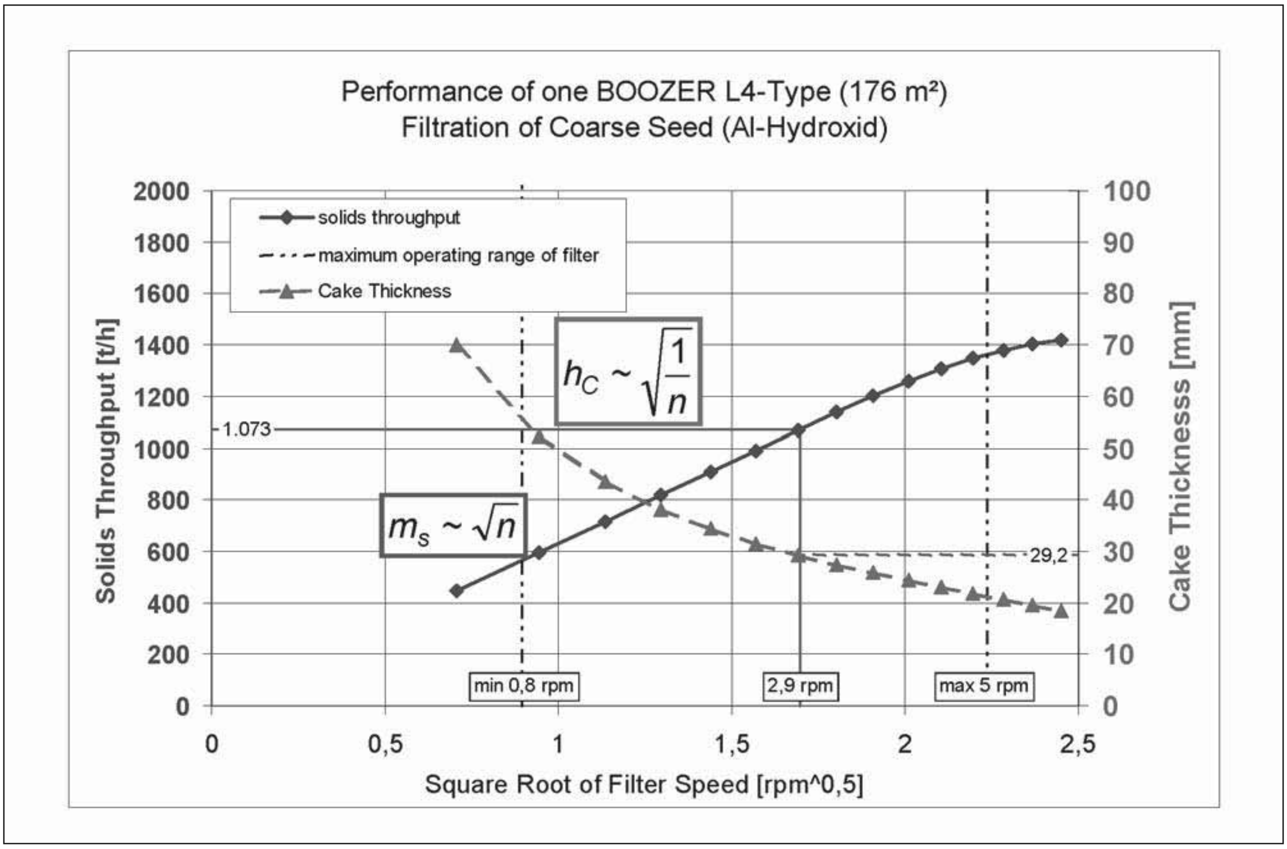


Figure 5: Solids throughput of a Boozer disc filter versus square root of filter speed n and cake height hc (Al-hydrate coarse seed filtration)

2. Filter cake dewatering

With the formulas known so far, it is hardly possible to predict the residual moisture of the cake without any test results. A rather reliable method for scaling up filters is to determine the residual moisture by a laboratory test at different cake heights and dewatering times and in case of need at different pressure differences.

The laboratory test results can be applied to rotary filters on the following assumptions:

- The flow resistance of the filter cloth is negligible.
- The pressure difference in the laboratory is equal to the actual value at the filter. In case of high gas flow this is not guaranteed for many filters due to deficient hydraulic of the filtrate pipes.
- The cakes are incompressible, which means there is no cake shrinking. In case of shrinkage crack forming and lifting of cake pieces the residual moisture at rotary filters is mostly very poor.
- The remoistening during cake discharge is low. Remoistening has a stronger impact in case of lower cake heights.

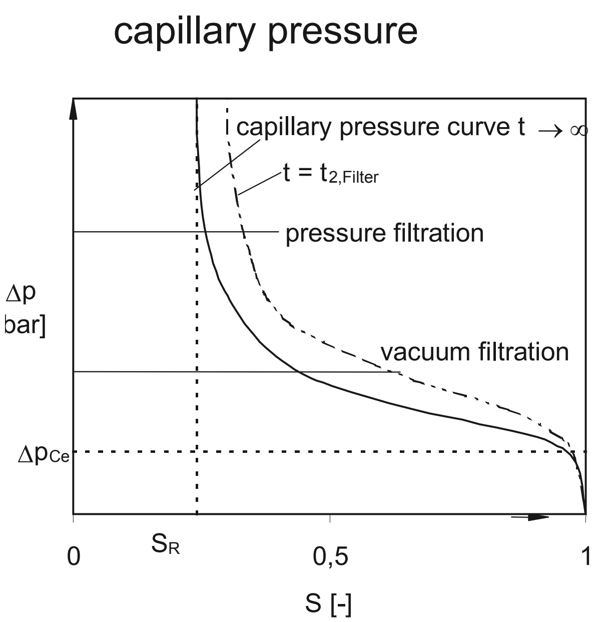


Figure 6: Capillary pressure curve

$$\frac{S-S_R}{1-S_R}=\left[1+A\cdot K\right]^{-B} \text{ with } K=\frac{(\Delta p-p_{CE})\cdot t_L}{R_C\cdot\varepsilon\cdot\eta_L\cdot h_C^2} \qquad (6)$$

A conversion results in equation 7:

$$S=\frac{rm}{\frac{\rho_L}{\rho_S}\cdot\frac{\varepsilon}{1-\varepsilon}\cdot(1-rm)} \quad (7) \text{ or } rm=\frac{S}{\frac{\rho_S}{\rho_L}\cdot\frac{1-\varepsilon}{\varepsilon}+S} \qquad (7)$$

It is however not recommended to calculate the residual moisture with equation 6 and 7 (like the calculation of cake forming). These equations are rather useful to make a simplification: In case the residual moisture mc1 or the saturation S1 at the setting hC1 and tL1 (dewatering time) are known from the lab test or the rotary filter, there is consequently a simplified proportionality between two filter settings (under keeping the residual moisture constant). This is displayed in equation 8.

$$\frac{t_{L1}}{h_{C1}^2}=\frac{t_{L2}}{h_{C2}^2} \qquad (8)$$

For rotary filters the dewatering- and cake forming time (tL and t1) is connected to the interrelation of the cake forming- and dewatering angle and to the filter speed n as well. This is shown in equation 9.

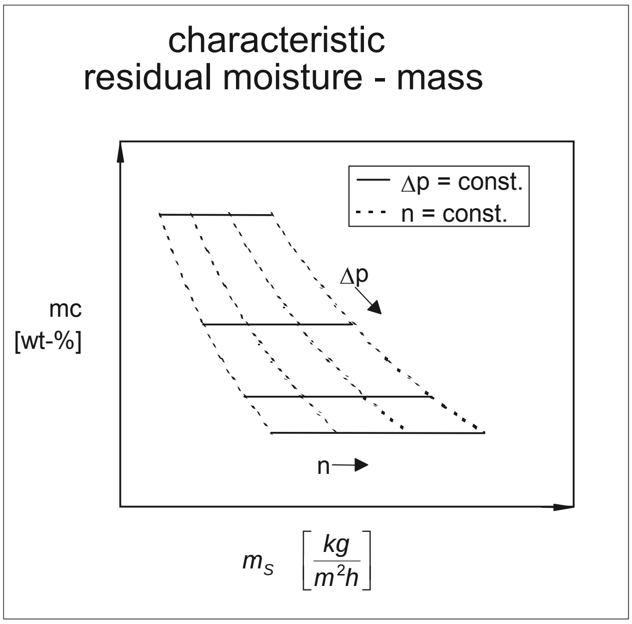
$$n=\frac{\alpha_1}{t_1\cdot 360°}=\frac{\alpha_L}{t_L\cdot 360°} \qquad (9)$$

As a result of equation 8 and 9 and the cake forming equation (2) the residual moisture does not depend on the filter speed (figure 7). This surprising interrelation is repeatedly proven during practical operation.

$$mc\neq f(n) \qquad (10)$$

In order to show how the dewatering time influences the residual moisture it is recommended to display the residual moisture in a function as follows: a2/a1 or t2/t1 (figure 8). These interrelations do not depend on the filter speed. This way of outlining can also be useful to easily compare residual moistures of different cake heights (cakes formed at different filter speeds).

The residual moisture can however rise at higher filter speed because the cake becomes thinner then and the quantity of remoistening liquid (filtrate coming from insufficient cell emptying) increases proportionally during cake discharge.



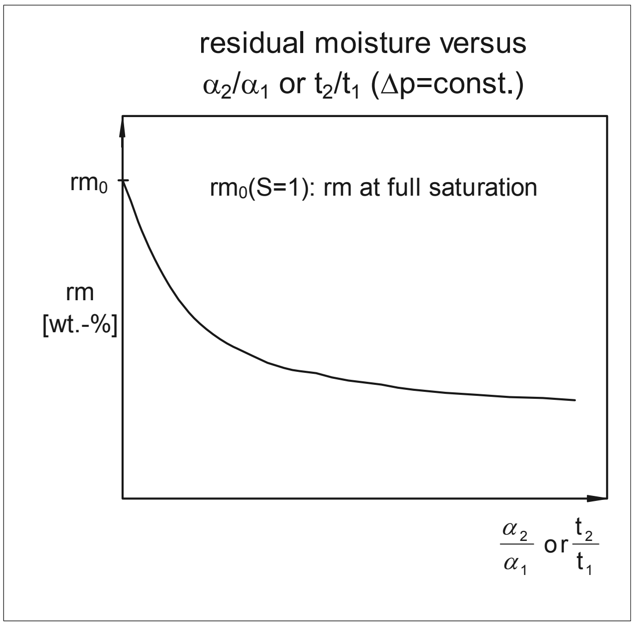


Figure 7 and 8: Residual moisture displayed as throughput function and as function of the interrelation between dewatering- and cake forming time

Note: The residual moisture of the cake does not depend on the filter speed. The throughput can thus be controlled by the filter speed without influencing the residual moisture.

3. Gas flow during dewatering

In general the gas flow increases when the pressure difference is increased as well. If a vacuum pump is already operating at its performance limit, the available pressure difference will drop upon raising the filter speed. This will immediately lead to lower residual moisture. Especially for disc filters a considerable improvement of the residual moisture (and thus the dewatering) can be achieved by homogenizing the cake height by heighten the immersion depth inside the trough for example. In case of shrinkage crack forming most of the air uselessly streams through these cracks into the filtration system. In order to avoid these cracks it is often helpful to lower the cake height.

It is recommended to determine the gas flow by a lab test using an appropriate measuring device (figure 2). It is mostly impossible to measure the gas flow during operation. Measuring the actual pressure difference is rather helpful here as well as a comparison to the manufacturer's performance curve of the vacuum pump.

4. Filter cake washing

The flow of wash water through the filter cake during cake wash happens according to Darcy's law (see equation 1). Accordingly, the flow of the washing fluid is influenced linearly by the cake height, which means twice the cake height requires twice the washing time. A constant washing ratio (the amount of washing fluid related to the amount of the filter cake) however, which is meaningful, requires the double amount of washing liquid and the washing time becomes 4 times as long. The interrelation between double cake height and quadruple time also applies to the cake forming (see equation 2) what means that the interrelation between wash fluid and cake amount (= wash ratio) is not influenced by changing the filter speed n but keeps constant.

Note: Changing the filter speed does not change the wash ratio. The throughput of a rotary filter can thus be controlled by the filter speed n without influencing the wash result. The residual moisture remains constant as well.

In addition, is has been experienced in practical operation that better washing results are achieved with thick cakes rather than with thin cakes. Of course this only applies to the following assumptions:

- constant wash ratio
- no crack forming
- no cake compacting upon increasing cake height

The diagram in figure 9 shows the typical course of a washing curve. The normalized load of pollution (e.g. salt) X* is plotted versus the wash ratio WVv. The wash ratio displays the interrelation between the added volume of washing fluid and the existing volume of the pores inside the filter cake. At WVv = 1 the pores of the filter cake are filled with washing fluid exactly one (1) time. Since the porosity of the filter cake is not exactly known in practical operation, the mass related washing ratio WVm is a more practical value (washing fluid volume related to solids mass).

In the displacement area (see figure 9) the polluted mother liquor inside the filter cake is almost completely replaced by the washing liquid (plug flow). In the transition area the washing is not only influenced by convection but also by diffusion processes in a considerable way. At higher wash ratios (in the diffusion area) the increasing of the wash water quantity does not have any influence on the washout as the system requires time for balancing of concentration differences by diffusion.

For this reason the cake washing is only meaningful within the range between washing ratio 1 and 3.

If a higher washing degree (a lower final load X) than 1:20 – 1:100 (maximum value depends on the product) shall be achieved, the cake will have to be re-slurried and then re-filtered and re-washed in a second filtration step.

Installed filters often do not achieve the required washing result. This is mostly due to crack forming in the cake or insufficient filtrate cell emptying and thus, remoistening of the cake with residual mother liquor. Both of these phenomena can efficiently be eliminated by a specific filter optimization.

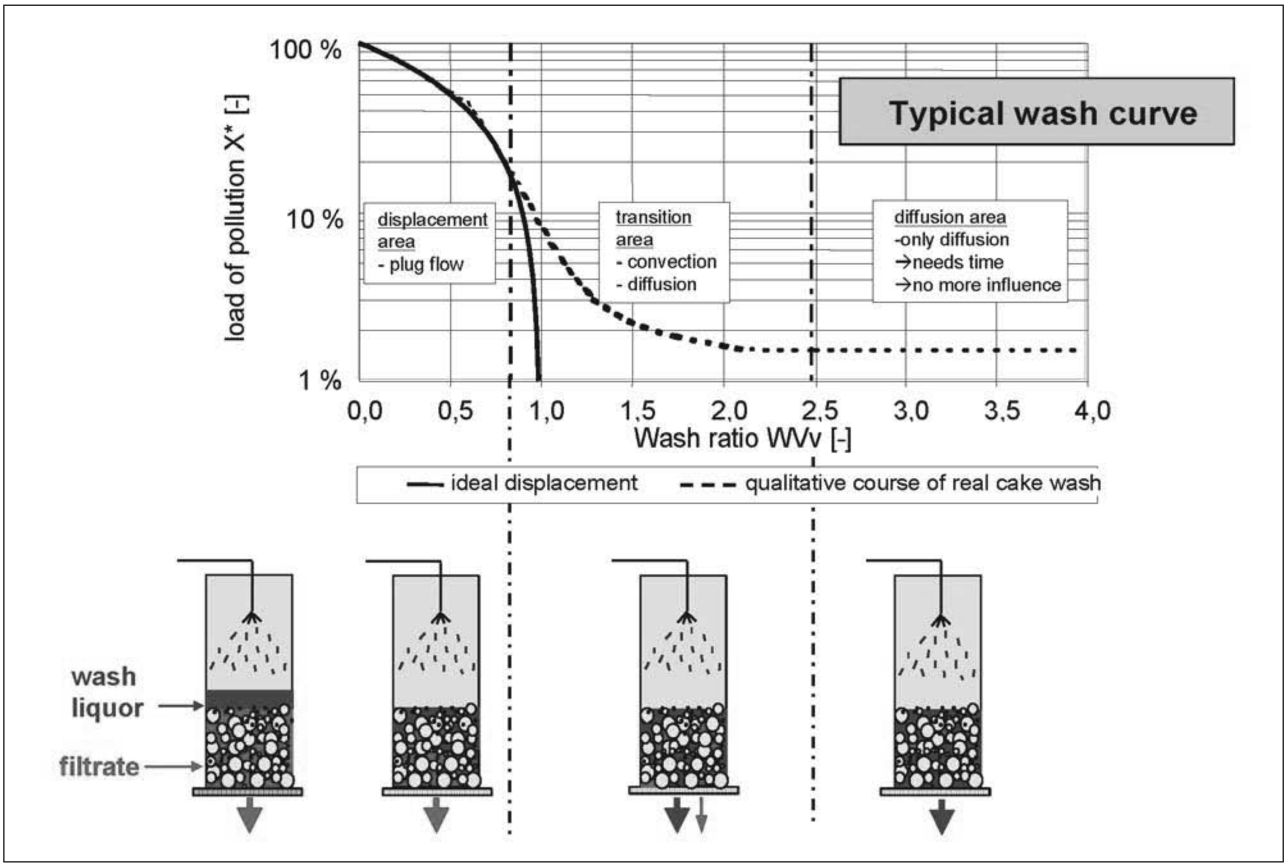


Figure 9: **Typical course of a washing curve**

5. Cake dewatering with steam and air flow

So far, it is not or hardly possible to predict or describe residual moisture with steam and air mathematically. The following statements result from own experience:

- The values determined by lab tests can be applied to rotary filters just like the scale-up calculation of the air pressure dewatering.
- Conversions concerning different cake heights or different times (tD and tL) are not recommended. The effect of the steaming- and the dewatering time with air must directly be determined by a test.

It is also not possible to convert the dewatering result to different pressure levels (e.g. from over-pressure to vacuum) as the saturated steam temperature changes along with the pressure level. The temperature of the dewatered cake ranges from 40 – 70 °C after the final throughput of pressurized air. The thumb-rule for the steam consumption and the dewatering result in comparison to the pressure filtration is:

Thumb-rule for the steam consumption:

$$mc_{SPF} \approx (0.3-0.8)\cdot mc_{PF}(\dot{m}_S = const.)$$
$$\dot{M}_{Steam} \approx 0.05 - 0.1 \cdot \dot{M}_{filter\ cake}$$

* SPF = steam pressure filtration, PF = pressure filtration

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6. Symbols

A, B, C	constants	
AF	filtering surface	[m^2]
HC	cake height	[mm]
$\dot{m}_s$	surface specific solids mass flow	[kg/m^2h]
n = a1 / (360° t1)	filter speed with cake forming time t1	[1/min]
Dp	efficient pressure difference	[bar]
DpCE	capillary entry pressure (pressure at first cake pore discharge)	[bar]
rC	specific cake resistance	[$1/m^2$]
rm	residual moisture	[wt.-%]
rm0	residual moisture at full saturation S=1	[wt.-%]
RC	cake resistance (=rK hK)	[1/m]
RM	flow resistance of filter medium at hK = 0	[1/m]
S	bulk saturation	
SR	residual bulk saturation (maximum mechanically achievable dewatering)	
t1	cake forming time	[sec]
tL	dewatering time with air	[sec]
$\dot{V}_l$	filtrate volume flow	[m^3/h]
WVv	volume related washing ratio	
WVm	mass related washing ratio	
X	load X = Nsalt/Ms, e. g. salt related to solid	
X*	normalized load of the polluted cake X*= X / X0 (X0 initial load of pollution of solids before washing)	
a1	actual cake forming angle	[°]
aL or a2	dewatering or air demoisturing angle	[°]
hL	fluid viscosity	[kg/ms]
k = cV /(1 - cV - e)	concentration parameter (cV = volume solids concentration of the suspension)	[-]
e	bulk porosity	[-]
rL	solids density	[g/cm^3]
rL	fluid density	[g/cm^3]