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CONTINUOUS PRESSURE OR DISCONTINUOUS PRESS FILTRATION TO SEPARATE SLURRIES OF VERY SMALL PARTICLES - A THEORETICAL COMPARISON -

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ABSTRACT

State of the art to separate particles in the sub-µm range from liquids by cake filtration are filter presses. Direct cake filtration in comparison to all competing techniques exhibits the advantage of additional cake dehumidification. The principle problem consists here of very high cake resistances. The more or less compressible filter cake is responsible for huge filtration times to build up thick filter cakes, which are necessary for the technical operation of conventional filter presses. In addition the discontinuous principle of a filter press leads to a limited throughput. A significant increase of filter capacity can be expected for continuously operating processes and very thin filter cakes. In this presentation the filtration process for a discontinuous filter press and a continuous rotary pressure filter will be analyzed and compared. The theoretical comparison will be validated by experimental data. As a result filtration conditions can be identified, which would be especially advantagues for a continuous thin layer pressure filtration process. Finally first ideas and preconditions for a technical realization of such an innovative process will be introduced.

KEYWORDS

sludge filtration, cake filtration, cake resistance, filter press, rotating filter

1. Introduction

The separation of particles in the sub-µm range from liquids by cake filtration today is – if at all - only possible by using time consuming discontinuously operating press filters like chamber or diaphragm presses. Alternatively such slurries can be separated by solid bowl centrifuges like decanters, disc stack separators or tube centrifuges, by cross-flow filtration or by precoat filters.
The direct cake filtration in comparison to the competing separation techniques is characterized by the possibility of additional post treatment methods like cake squeezing or desaturation and avoids contamination of the solid product by precoat materials.
The mean spec. cake resistances $\bar{r}_C$ of conventional cake filtration processes today are placed in between of $10^{12}m^{-2}$ and $10^{16}m^{-2}$. The idea is to extend this range to spec. cake resistances of $10^{17}m^{-2}$ to $10^{18}m^{-2}$.
A principle problem of small particle filtration consists of the arising very high filter cake resistance. The filter cake can be more or less compressible depending on the stability of the slurry. For stable slurries even for very small particles an only slight compressible cake structure of minimal porosity and thus maximal flow resistance is developing. An improvement of filtrability by destabilisation of such slurries leads to the formation of agglomerates, which are building up more or less compressible cake structures. The most condensed particle layer is located at the filter medium. In the direction to the cake surface the porosity increases more and more. Particularly the great flow resistance of the particle layers near the filter medium is responsible for the huge filtration time to build up filter cakes in the cm range, which are necessary for the technical operation of conventional filter presses. If a filter cake exhibits a porosity gradient it contains after the cake formation some liquid, which can displaced by squeezing. Under circumstances the homogeneous cake structure after squeezing can be desaturated by pressurized gas.
The discontinuous operation of a separation apparatus leads to a limited throughput due to the so called „dead-time". As dead-time the sum of all steps during a filter cycle is defined, during which no filter cake is produced. This contains the time for feeding, cake post-treatment by washing and dehumidification, cake discharge and apparatus cleaning. The maximum throughput of a discontinuous filter apparatus is given for equity of cake formation time and dead time.
A significant increase of filter capacity can be expected if the separation process would be designed to work continuously and would be finished after the formation of a comparatively very thin filter cake of 1mm or less. The maximal throughput of a continuously operating rotary filter apparatus is determined by the minimal cake height, which still can be discharged.
Fig. 1 shows the basic principle of both filter systems to be compared schematically.

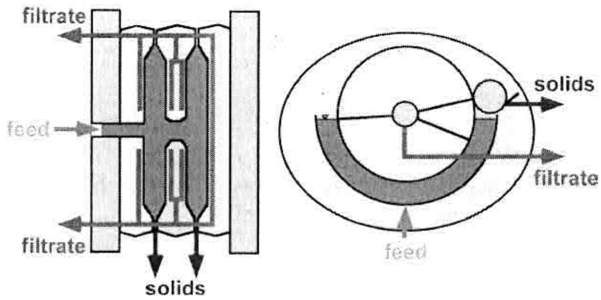

Fig.1: Schematic representation of filter press (left) and rotary pressure filter (right)

2. Basis of calculations to compare filter press and rotary pressure filter

2.1 Discontinuous filter press

For the comparison of the systems a chamber depth of 50mm for the filter press is defined. This means a cake thickness of 25mm, which grows up from both sides of the chamber. The necessary cake thickness of h_c=25mm fixes the cake formation time t_1. The cake dehumidification time t_2 in the case of the rotary filter should be equal to the cake formation time t_1. To compare both systems on the basis of same conditions also for the filter press should be valid that $t_1=t_2$. The time t_3 necessary for filling the press, opening the chambers, cake discharge and closing the press again is assumed to be 30min. The „dead time", which means the time of no filter cake production can be calculated now as:

$$t_{dead} = t_2 + t_3 \qquad \text{eq.1}$$

The spec. solids throughput of a filter press $\dot{m}_{s,FP}$ can be formulated as:

$$\dot{m}_{s,FP} = \frac{\dot{M}_s}{A \cdot (t_1 + t_{dead})} = \frac{A \cdot h_{c,FP} \cdot (1-\bar{\varepsilon}) \cdot \rho_s}{A \cdot (t_{1,FP} + t_{dead})} = \frac{h_{c,FP} \cdot (1-\bar{\varepsilon}) \cdot \rho_s}{t_{1,FP} + t_{dead}} \qquad \text{eq.2}$$

$\dot{M}_s$ abs. solids throughput $[kg/m^2 \cdot h]$
A filter area [m^2]
$h_{c,FP}$ cake thickness for filter press [mm]
$\bar{\varepsilon}$ mean cake prorsity [-]
ρ_s spec. weight of solids [kg/m^3]

The filter cake formation under the condition of neglectable filter cake resistance can be calculated according to eq.3:

$$h_{c,FP} = \sqrt{\frac{2 \cdot \bar{\kappa} \cdot \Delta p}{\bar{r}_c \cdot \eta_L} \cdot t_{1,FP}} \qquad \text{eq.3}$$

$$\bar{\kappa} = \frac{c_V}{1 - c_V - \bar{\varepsilon}} \qquad \text{eq.4}$$

$\bar{\kappa}$ concentration parameter [-]
Δp pressure difference [MPa]
c_v solids volume concentration of the slurry [-]
$\bar{r}_c$ mean spec. filter cake resistance $[m^{-2}]$
η_L dyn. liquid viscosity $[Pa \cdot sec]$

From filtration theory for compressible cake filtration can be derived, that the cake porosity ε and thus the spec. cake resistance r_c are not depending on the cake height but only on the pressure difference Δp. As a consequence the cake formation for compressible filter cakes can be calculated using the formula for incompressible cake filtration, if for porosity and spec. cake resistance the integral mean values are used.
From theoretical calculations can be derived further on, that the maximum throughput of a discontinuous filter press can be realized, if t_1 equals t_{dead}. This condition cannot be achieved here due to the fixed values of t_1, t_2 and t_3. For difficult to filter suspensions this is also valid for many practical applications, where the cake formation time exceeds the dead time by the multiple.

2.2 Continuous rotary pressure filter

The spec. solids throughput for a rotary drum filter $\dot{m}^{s}_{s,DF}$ can be calculated by eq.5:

$$\dot{m}^{s}_{s,DF} = n \cdot \rho_S \cdot h_{c,DF} \cdot (1-\bar{\varepsilon}) \qquad \text{eq.5}$$

n rotation speed of the filter drum [min^{-1}]

The rotation speed of the filter drum corresponds to the cake formation time t_1 by the cake formation angle α_1, which is defined in the filters control valve:

$$t_1 = \frac{\alpha_1}{360^{\circ}} \cdot \frac{1}{n} \qquad \text{eq.6}$$

For better understanding the process zones of a rotary filter are explained schematically in Fig.2.

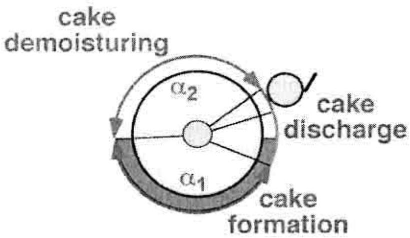

Fig.2: Process zones of a rotary pressure filter

Transferred to the formula of eq.3 for filter cake formation follows for the rotary filter:

$$h_{c,DF} = \sqrt{\frac{2 \cdot \bar{\kappa} \cdot \Delta p}{\bar{r}_c \cdot \eta_L} \cdot t_{1,DF}} = \sqrt{\frac{2 \cdot \bar{\kappa} \cdot \Delta p}{\bar{r}_c \cdot \eta_L} \cdot \frac{\alpha_1}{360^{\circ}} \cdot \frac{1}{n}} \qquad \text{eq.7}$$

Maximum throughput is given for minimum cake height. The minimum cake height is limited to the cake layer, which can be just about discharged from the filter.
For the comparative calculations the operational conditions of the rotary filter are set as follows:

$h_{c,min}$ 1mm
n_{max} $12 min^{-1}$
α_1 160°
α_2 160° ($\alpha_1 = \alpha_2$)

3. Experimental reference data

To guarantee realistic results of the calculations data from several filtration experiments have been utilized. Table 1 gives an overview about some key data of the investigated products. Soot and Kaolin are products, which are in the range of conventional press filtration. Böhmit exeeds this range and cannot be separated by conventional cake filtration techniques today.

	Soot	Soot	Kaolin	Kaolin	Böhmit (Al-Metahydr.)
$x_{50,3}$ [µm]	13.6	13.6	9.4	9.4	0.072
ρ_s [kg/m^3]	1870	1870	2600	2600	3000
ρ_L [kg/m^3]	1000 (water)	1000 (water)	1000 (water)	1000 (water)	1000 (water)
η_L [Pa sec]	0.001	0.001	0.001	0.001	0.001
c_v [-]	0.015	0.015	0.100	0.100	0.050
pH [-]	2.5	2.5	7.0	7.0	5.0
ZP [mV]	-23	-23	-23	-23	+55
Δp [MPa]	0.1	1.2	0.1	1.2	0.6
ε [-]	0.959	0.939	0.675	0.620	0.740
r_c [m^{-2}]	$2.3\ 10^{14}$	$1.1\ 10^{15}$	$1.5\ 10^{15}$	$1.8\ 10^{15}$	$1.6\ 10^{17}$
	very strong compressible	very strong compressible	compressible	compressible	low compressible

Table 1: Key data of the filtered suspensions

Fig.3, Fig.4 and Fig.5 show microscopic pictures of the particle systems.

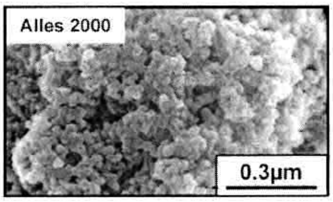

Fig.3: Soot

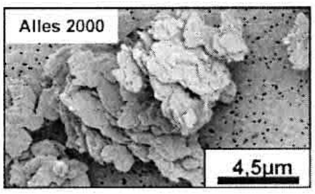

Fig.4: Kaolin

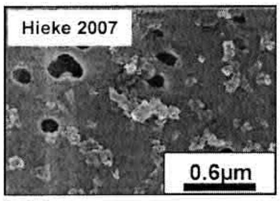

Fig.5: Böhmite

4. Comparison of the filter systems

At first the ratio of the spec. solids throughput for filter press and pressure filter has been formulated according to eq.2 and eq.5:

$$\frac{\dot{m}_{s,DF}}{\dot{m}_{s,FP}} = \frac{\rho_s \cdot (1-\bar{\varepsilon}) \cdot h_{c,DF} \cdot \frac{\alpha_1}{360^\circ} \cdot \frac{1}{t_{1,DF}}}{\rho_s \cdot (1-\bar{\varepsilon}) \cdot h_{c,FP} \cdot \frac{1}{t_{1,FP}+t_{2,FP}+t_{3,FP}}} \qquad \text{eq.8}$$

The terms which contain product parameters like spec. weight ρ_s or mean porosity $\bar{\varepsilon}$ can be cancelled, because the same product is filtered in both systems at the same pressure. In addition the following data had been fixed before in chapter 2:

$h_{c,FP}$= 25mm $\quad$ $t_{3,FP}$=30min $\quad$ $t_{1,FP}=t_{2,FP}$

$h_{c,DF}$= 1mm $\quad$ $\alpha_1=\alpha_2=160^\circ \Rightarrow t_{1,DF}=t_{2,DF}$

$t_1 \propto h_c^2$

From this follows:

$$h_{c,FP} = 25 \cdot h_{c,DF} \Rightarrow t_{1,FP} = t_{1,DF} \cdot \frac{25^2 \cdot h_{c,DF}^2}{h_{c,DF}^2} = t_{1,DF} \cdot 625 \qquad \text{eq.9}$$

If all these conditions are put into eq.8 the ratio of solids throughput for both filter systems results as:

$$\frac{\dot{m}_{s,DF}}{\dot{m}_{s,FP}} = \frac{0.44 \cdot \frac{1}{t_{1,DF}}}{25 \cdot \frac{1}{2 \cdot 625 \cdot t_{1,DF} + t_3}} = 0.018 \cdot (1250 + \frac{t_3}{t_{1,DF}}) \approx 22 \qquad \text{eq.10}$$

for $t_3 \rightarrow 0$ or $t_{1,DF} \rightarrow \infty$

In table 2, table 3 and table 4 the calculated results for the reference products are presented.

<u>Soot</u>

$\bar{r}_{c,0.1MPa}$= 2.30 $10^{14}m^{-2}$, $\bar{r}_{c,1.2MPa}$= 1.14 $10^{15}m^{-2}$

Δp [MPa]	$t_{1,DF}$ [min]	$t_{1,FP}$ [min]	$\dot{m}_{s,DF}$ [kg/m³h]	$\dot{m}_{s,FP}$ [kg/m³h]	$\dot{m}_{s,DF}/\dot{m}_{s,FP}$ [-]
0.1	0.03	20.8	61.2	1.6	38.1
1.2	0.02	15.2	125.1	2.8	44.2

Table 2: Comparison of the filter systems for soot

Kaolin

$\bar{r}_{c,0.1MPa}$= 1.52 $10^{15}m^{-2}$, $\bar{r}_{c,1.2MPa}$= 1.78 $10^{15}m^{-2}$

Δp [MPa]	$t_{1,DF}$ [min]	$t_{1,FP}$ [min]	$\dot{m}_{s,DF}$ [kg/m³h]	$\dot{m}_{s,FP}$ [kg/m³h]	$\dot{m}_{s,DF}/\dot{m}_{s,FP}$ [-]
0.1	0.3	178.4	79.1	3.3	24.1
1.2	0.04	21.6	759.4	20.2	37.5

Table 3: Comparison of the filter systems for Kaolin

Böhmite

$\bar{r}_{c,0.6MPa}$= 1.6 $10^{17}m^{-2}$

Δp [MPa]	$t_{1,DF}$ [min]	$t_{1,FP}$ [min]	$\dot{m}_{s,DF}$ [kg/m³h]	$\dot{m}_{s,FP}$ [kg/m³h]	$\dot{m}_{s,DF}/\dot{m}_{s,FP}$ [-]
0.6	9.3	5836	2.2	0.1	22.3

Table 4: Comparison of the filter systems for Böhmite

As can be seen especially from table 4 huge differences exist in the cake formation time for filter press and rotary pressure filter. It needs in the range of minutes to build up the cake by the rotary filter whereby it needs days in the case of the discontinuous filter press.

In Fig.6 results for a simulation calculation can be seen. The comparison of filter press and rotary filter was made here for different spec. cake resistances. The following data have been used for the calculation:

ρ_s = 3000kg/m³ ρ_L = 1000kg/m³ $\eta_L = 10^{-3}$Pa sec c_v = 0.1 ε = 0.7
$h_{c,DF,min}$ = 1mm α_1 = 160° $h_{c,FP}$ = 25mm $t_{3,FP}$ = 30min

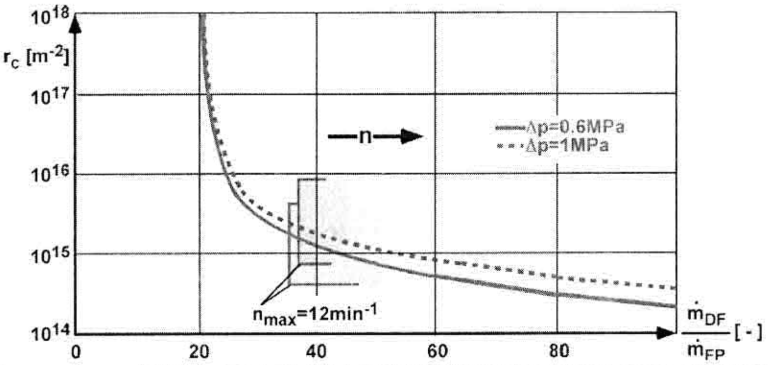

Fig.6: Comparison of filter press and rotary filter for varied spec. cake resistance

According to eq.10 the ratio of solids throughput is approaching the value 22 for very high spec. cake resistances and thus large filtration times. Towards low cake resistance the ratio becomes higher and higher. A limitation is set by the maximum rotation speed $n=12 min^{-1}$ of the rotary filter.

5. Conclusions and outlook

For very hard to filter suspensions of $\bar{r}_c > 10^{17} m^{-2}$ conventional discontinuous press filters are no option due to the huge filtration time. Here an innovative continuous rotary pressure filter could reduce the filtration time to minutes. Cake dehumidification can be realized here only by squeezing.
In the medium range of $10^{14} m^{-2} < \bar{r}_c < 10^{15} m^{-2}$ filter presses as well as rotary pressure filters can operate. In the case of possible desaturation by overcoming the capillary pressure in the cakes pores shrinkage cracks are probable. In such cases filter presses and especially diaphragm presses are advantageous. Very interesting would be more research about eventual avoiding of shrinkage cracks when forming very thin cakes.
For easy to filter products in the range of $10^{12} m^{-2} < \bar{r}_c < 10^{13} m^{-2}$ rotary filters are advantageous due to the incompressible cake structures and the possibility of cake desaturation.
A first approach for a technical realization of a continuous thin layer pressure filtration is a rotary drum filter with microporous filter media and roller discharge in a pressure vessel (hyperbar filter). The principle of a drum filter enables nearly unlimited variations of cake formation time. The drum filter is the most flexible member of the rotary filter family. The difficult discharge of the very thin filter cakes probably could be solved by an adjusted roller discharge system. For the direct filtration of extremely fine grained suspensions spezial filter media are required, which today are not yet offered by the market. However, products like shown in Fig.7 could be the right answer to the demand.

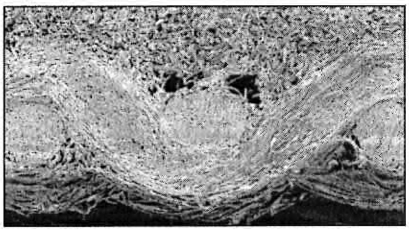

Fig.7: „Azurtex“ (Madison Filter)

This filter medium consists of a fabric with integrated microporous membrane. The present nominal pore size of such media is around 6µm. Between 0.1µm and 1µm would be needed for the process discussed here.