

**GAS CONSUMPTION, SOLIDS THROUGHPUT AND RESIDUAL CAKE  
MOISTURE – RELATION BETWEEN OPERATING EXPENSES AND PROCESS  
RESULTS OF ROTARY FILTERS**

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**ABSTRACT**

During the operation of continuously operating rotary vacuum and pressure filters a gas flow arises through the partially deliquored filter cake. This gas flow is determining significantly the energy demand of the process and thus the operating expenses. To maintain the filtration pressure for cake formation and deliquoring the gas flow must be exhausted in the case of vacuum filtration and be delivered in the case of pressure filtration by the compressor into the filter housing. The necessary expenses in form of gas consumption are corresponding with the process results in form of solids throughput and residual cake. One question to be answered is, whether a certain cake moisture can be achieved most energy efficient with regard to the ratio of demanded gas consumption per unit solids production. As variable parameters the relation of cake formation and cake deliquoring angle in the filters control head, the rotation speed of the filter, the gas pressure difference and the slurry concentration had been selected for this investigation. For the formation of approximately incompressible filter cakes, their desaturation and the parallel arising gas flow physically based model equations are known, which contain the relevant influencing parameters. These can be transferred to the operational conditions of a rotary filter. Thus the correlations between energy demand, operational parameters and process results could be estimated and compared with the results of a broadly layed out experimental parameter study. This study was based on discontinuous laboratory filter experiments with a zincsulfide slurry from the mineral processing. From the results could be concluded, that for rotary filters the ratio of gas consumption per produced solids mass is depending mainly on the residual moisture content of the filter cake. The parameter combination to achieve a certain moisture content seems to be not relevant. As a final consequence the parameter combination for a desired moisture content should be set for maximal solids production. This leads for a given slurry feed flow and constant energy demand to the smallest necessary filter area and thus the optimal solution for the given separation task.

**KEYWORDS:** solid-liquid-separation, cake filtration, rotary filter, gas flow, energy demand

## 1. INTRODUCTION AND PROBLEM

The gas flow through the partly deliquored filter cake during the operation of continuous vacuum and pressure filters decisively determines the energy consumption of the process and thus the operating expenses. Fig.1 shows exemplarily for the family of rotary filters a vacuum and a pressure drum filter [1, 2].

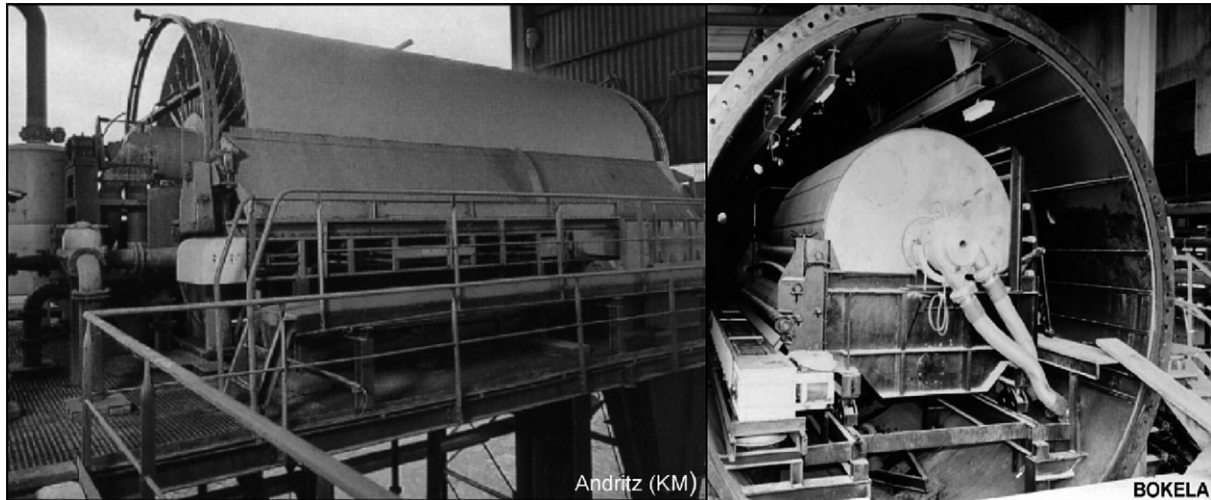


Fig.1: Vacuum and pressure drum filter

To maintain the pressure difference  $\Delta p$  for cake formation and deliquoring this gas stream must be sucked off via the filtrate receivers in the case of vacuum filtration and supplied by a compressor to the filter housing in case of pressure filtration. In both cases the appropriate technical gas compression work must be provided. As can be seen in fig.2 the gas throughput  $Q_g$  rises for const. pressure difference  $\Delta p$  with progressing deliquoring time  $t_2$  according to the complete emptying of throughgoing pores in the filter cake and thus decreasing saturation  $S$ .

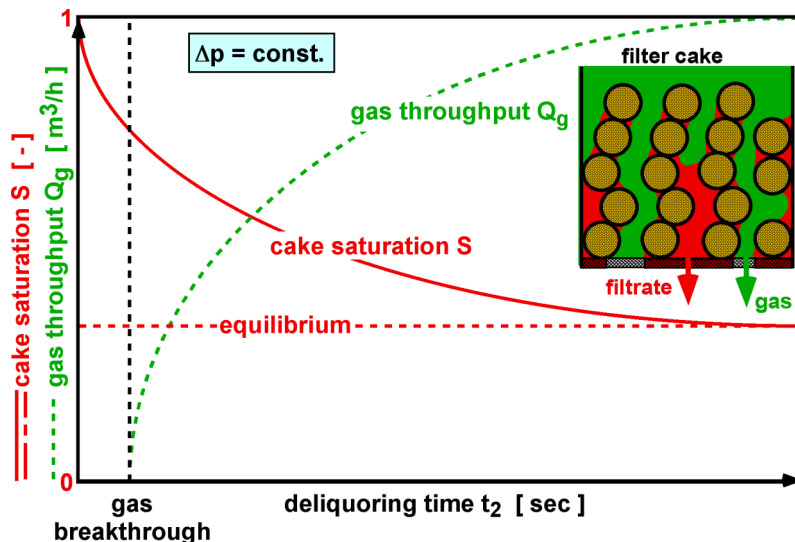


Fig.2: Kinetics of filter cake deliquoring

The principle process shown in fig.2 can be transferred according to fig.3 to a rotary filter. The cake deliquoring starts as soon as the cake appears from the slurry and ends at the moment of cake discharge. The cake formation time  $t_1$  is realized by the filter rotational speed  $n$  and the cake formation angle  $\alpha_1$ .

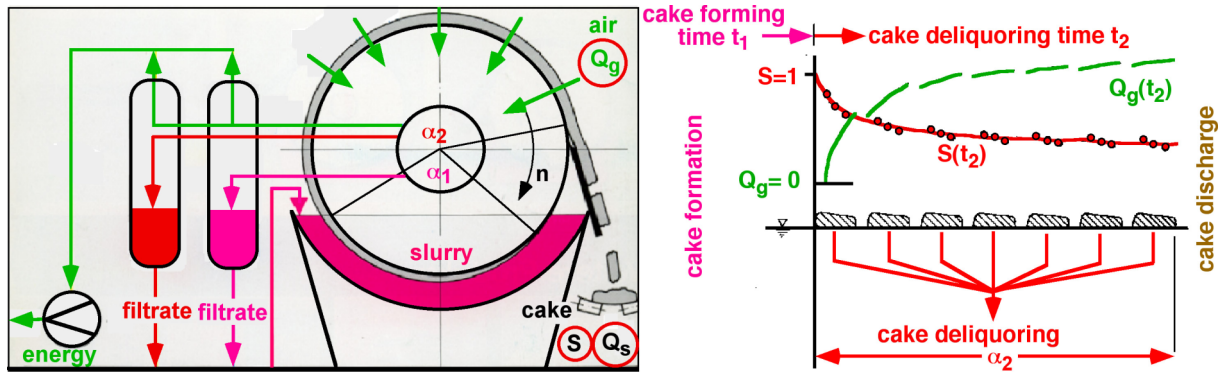


Fig.3: Cake deliquoring on a rotary filter

The cake deliquoring time  $t_2$  is given by the same rotational speed and the deliquoring angle  $\alpha_2$ . The expenditure for the filtration in form of the gas throughput  $Q_g$  corresponds with the results of the separation process in form of solids mass throughput  $Q_s$  and the resulting saturation degree  $S$  of the filter cake. In the following the correlations between expenditure and result of continuous cake filtration should be examined. Amongst others the question should be answered, whether a certain cake saturation can be achieved particularly energy saving by a special parameter combination. As variable parameters were chosen the ratio between cake forming and cake deliquoring angle in the control head of the filter, the rotation speed of the filter, the gas pressure difference and the solids volume concentration of the slurry. For the formation of approximately incompressible filter cakes, their deliquoring by desaturation and the simultaneously appearing gas throughput physically based model equations are available, which contain the decisive influencing parameters [3, 4]. These can be transferred to the operating conditions of a rotary filter. Thus the correlations between spec energy demand, filtration result and set of operational parameters could be estimated and compared with the results of an extensive experimental parameter study. This study was based on lab scale experiments with a zinc sulfide slurry from the mineral processing. There is evidence, that for rotary filters the ratio of gas consumption and produced solids mass mainly depends on the residual cake moisture or rather the resulting permeability for the gas flow. As a result it seems to be irrelevant which parameter combination this cake saturation degree had been achieved. For the purpose of a process optimization the smallest necessary filter area could be chosen for a given slurry feed flow and a required residual cake moisture, because the solids related gas consumption remained constant for all parameter combinations.

## 2. PHYSICAL BACKGROUND

The rotary filters spec. solids mass throughput  $q_s$ , which is defined as relation of absolute solids mass throughput  $Q_s$  and filter area  $A$ , is calculated from the product of spec. solids weight  $\rho_s$ , the solids volume fraction of the cake  $(1-\varepsilon)$ , the cake thickness  $h_c$  and the rotation speed of the filter  $n$  [5].

$$q_s = \frac{Q_s}{A} = \rho_s \cdot (1 - \varepsilon) \cdot h_c \cdot n \quad (1)$$

The porosity  $\varepsilon$  can be determined from the relation of void volume in the cake  $V_v$  and total cake volume  $V_{tot}$ . For this purpose in a lab scale pressure filter cell (see fig.4) a

cylindric filter cake of area  $A$ , height  $h_c$  and known spec. solids weight  $\rho_s$  is formed and its solids mass is measured.

$$\varepsilon = \frac{V_V}{V_{tot}} = \frac{V_V}{V_S + V_V} = \frac{V_{tot} - V_S}{A \cdot h_c} = \frac{(A \cdot h_c) - \frac{m_s}{\rho_s}}{A \cdot h_c} = 1 - \frac{m_s}{A \cdot h_c \cdot \rho_s} \quad (2)$$

The cake thickness  $h_c$  can be calculated from the square root of concentration parameter  $\kappa$ , gas pressure difference  $\Delta p$ , spec. cake resistance  $r_c$ , dyn. liquid viscosity  $\eta_L$  and cake forming time  $t_1$ .

$$q_s = \rho_s \cdot (1 - \varepsilon) \cdot \sqrt{\frac{2 \cdot \kappa \cdot \Delta p \cdot t_1}{r_c \cdot \eta_L}} \cdot n \quad (3)$$

The spec. cake resistance  $r_c$  is determined according to VDI guideline 2762 by measurement [6]. The concentration parameter  $\kappa$  is calculated on the base of a mass balance around the solid-liquid system from the solids volume concentration  $c_v$  and the cake porosity  $\varepsilon$ .

$$\kappa = \frac{c_v}{1 - c_v - \varepsilon} \quad (4)$$

The cake forming time  $t_1$  on the rotary filter can be expressed finally by the cake forming angle  $\alpha_1$  and the rotation speed  $n$ .

$$q_s = \rho_s \cdot (1 - \varepsilon) \cdot \sqrt{\frac{2 \cdot \kappa \cdot \Delta p \cdot \alpha_1 \cdot n}{r_c \cdot \eta_L \cdot 360^\circ}} \quad (5)$$

For the representation of the filter cake deliquoring at first must be defined, how the deliquoring result should be expressed. The measurement delivers first by an easy weighing of the moist and afterwards the dry cake the residual moisture content MC. Here the mass of the remaining liquid in the cake  $m_L$  is related to the total mass of the moist cake ( $m_s + m_L$ ).

$$MC = \frac{m_L}{m_s + m_L} \quad (6)$$

Alternatively the saturation degree  $S$  can be used as a standardised parameter. Here the remaining liquid volume in the moist cake is related to the totally available void volume in the cake  $V_V$ .

$$S = \frac{V_L}{V_V} \quad (7)$$

To determine the void volume additionally the knowledge of the cakes porosity is necessary. Moisture content MC and saturation degree  $S$  can be converted into each other.

$$S = \frac{\rho_s \cdot (1 - \varepsilon) \cdot MC}{\rho_L \cdot \varepsilon \cdot (1 - MC)} \quad (8)$$

For the representation of the experimental results here mainly the moisture content is used als primary measured data. On the saturation degree is referred, if it is sensefull for interpretation reasons. The temporal decrease of the saturation degree for

incompressible and isotropic filter cakes generally can be expressed by an „deliquoring parameter“.

$$S(t_2) = f \left[ \frac{r_c \cdot \varepsilon \cdot \eta_L}{(\Delta p - p_{cap,e})} \cdot \frac{h_c^2}{t_2} \right] = f \left[ \frac{r_c \cdot \varepsilon \cdot \eta_L}{(\Delta p - p_{cap,e})} \cdot \frac{2 \cdot \kappa \cdot \Delta p}{r_c \cdot \eta_L} \cdot \frac{t_1}{t_2} \right] \quad (9)$$

$$S(t_2) = f \left[ \frac{r_c \cdot \varepsilon \cdot \eta_L}{(\Delta p - p_{cap,e})} \cdot \frac{2 \cdot \kappa \cdot \Delta p}{r_c \cdot \eta_L} \cdot \frac{\alpha_1}{\alpha_2} \right]$$

If this function is held constant for a variation of the contained parameters, the deliquoring result should remain constant too. From eq.(9) can be derived, that for rotary filters a change of the rotary speed  $n$  should have no influence on the saturation degree of the discharged cake, if the other parameters are held constant. The square of the cake height  $h_c$  is proportional to the cake formation time  $t_1$  and the ratio of cake forming and cake deliquoring time is proportional to the ratio of the cake forming and cake deliquoring angle  $\alpha_1/\alpha_2$ , because the rotation speed  $n$  can be canceled. If one is changing the ratio of the angles  $\alpha_1/\alpha_2$  this can be compensated by an adjustment of the concentration parameter  $\kappa$ . If the porosity of the cake remains constant a change of  $\kappa$  means a change of the slurry solids volume concentration  $c_v$ , which is fed to the filter. In the same way a change of the pressure difference can be treated. However the targeted residual cake moisture must be achievable against the acting capillary pressure in the filter cakes pores. Principally every change of parameters must be checked, whether it can be still realized technically on a rotary filter.

For the gas flow through the filter cake the hollow void structure accessible for the gas flow is decisive. Here basically is assumed, that liquid and gas are flowing independently and are not influencing each other. Furthermore can be supposed, that the liquid is displaced homogeneously from all locations of the filter cake. In that case no moisture gradient is forming along the filter cake height. Thus in the equilibrium state the saturation degree is no function of the cake height. In contrast to the liquid gas is a compressible fluid, which is expanding on the way through the filter cake. This is considered with the description of the gas throughput by a special pressure term and the relation to the gas entry condition (index „e“) or the gas outlet condition (index „o“).

$$Q_{g,e} = \frac{p_c \cdot p_{c,rel,g}(S)}{\eta_g} \cdot \frac{p_m \cdot \Delta p}{p_{g,e}} \cdot \frac{A}{h_c} \quad (10)$$

The saturation dependent spec. gas permeability  $p_{c,g}(S)$  results from the product of single phase permeability and a relative gas permeability  $p_{c,rel,g}$ , which varies between 0 and 100% [7].

$$p_c(S) = p_c \cdot p_{c,rel,g}(S) \quad (11)$$

Here it becomes clear once again, that the gas throughput is a result and not the cause of the diliquoring process. Decisive for the liquid displacement is only the overcoming of the capillary pressure by the applied gas pressure difference. The gas throughput measured under feed conditions during the discontinuous filter experiment in the lab scale filter cell can be transferred also on the rotary filter. For this purpose at first the gas volume flow  $Q_g(t_2)$  originating in the lab scale experiment during the deliquoring time  $t_2$  is registered. Then the deliquoring time relevant for the

rotary filter has to be determined. Over the whole deliquoring angle of the rotating filter the gas flow results together. To be able to determine this stationary average gas volume flow on the rotary filter, the time-resolved measured gas volume flow in the lab scale experiment must be integrated from the time of first gas breakthrough  $t_{2,0}$  up to the calculated deliquoring time  $t_2$  and afterwards divided by this time interval. For the reason of comparability of the data the gas throughput of the lab scale filter  $q_{g,N,lf}(t_2)$  is converted to standard conditions (index „N“) and related to one square meter of filter area. From this finally the mean spec. gas throughput of the rotary filter results.

$$q_{g,N,rf} = \frac{\int_{t_{2,0}}^{t_2} q_{g,N,lf}(t_2) \cdot dt_2}{t_2} \quad (12)$$

From eq.(10) is evident, that the spec. gas throughput for const. cake saturation degree and thus const. relative gas permeability is according to the Darcy-law inverse proportional to the cake thickness. The cake thickness itself is inverse proportional to the square root of the filters rotational speed. As a consequence the spec. gas throughput of the rotary filter should be proportional to the square root of the rotational speed.

$$q_{g,N,rf} \propto \sqrt{n} \quad (13)$$

The spec solids mass throughput is propotional to the square root of the rotational speed too. This leads to the conclusion, that the ratio of solids mass throughput and gas throughput for const pressure difference should be independent on the rotational speed and thus constant.

$$\frac{q_{g,N,rf}}{q_{m,s}} \propto \frac{\sqrt{n}}{\sqrt{n}} = 1 \quad \left[ \frac{m_N^3 / m^2 \cdot h}{kg / m^2 \cdot h} \right] = \left[ \frac{m_N^3}{kg} \right] \quad (14)$$

Then this would also have to be valid for all other rotary filter settings with which the same void structure or saturation degree of the filter cake and thus the same relative gas permeability is achieved.

To be able to estimate the economic efficiency of the process and to compare with other filtration processes, still the energy, which is to be used for the compression of the gas from standard conditions to the operating pressure, is to be estimated at the end. In a first assumption the gas is considered to be ideal and adiabatic and isentropic changes of condition are valid. The isentropic coefficient for air is set to the value  $\kappa=1.4$ . Now the technical compression work  $W_t$  for the overpressure filtration examined further on can be calculated.

$$W_t = p_N \cdot V_N \cdot \frac{\kappa}{(\kappa - 1)} \cdot \left[ 1 - \left[ \frac{p_N}{p_e} \right]^{\frac{(\kappa-1)}{\kappa}} \right] \quad (15)$$

Eq.(15) contains as additional idealization a compressor efficiency  $\eta_{\text{compr}}=1$ , which is clearly smaller in reality. Thus the compression work would have to become still divided by the compressor efficiency. To create a general basis for comparison of

results a compression work related to the separated solid mass should be defined here as spec. compression work  $w_{t,m}$ .

$$w_{t,m} = \frac{p_N \cdot V_N}{\eta_{compr} \cdot m_s} \cdot \frac{\kappa}{(\kappa - 1)} \cdot \left[ 1 - \left[ \frac{p_N}{p_e} \right]^{\frac{(\kappa - 1)}{\kappa}} \right] \quad (16)$$

The functional correlations for the spec. gas compression work are of course qualitatively the same like for the spec. gas consumption as will be shown later on.

### 3. MEANS AND METHODS

For the filtration experiments a zinc sulphide/water suspension from the mineral processing was used. The volume related mean particle diameter amounted  $x_{50,3}=26\mu\text{m}$  and the width of the particle size distribution was determined by the relation  $x_{90,3}/x_{10,3}=26$ . The measurement of the spec. particle weight resulted to  $\rho_s=4,22\text{gcm}^{-3}$ . To avoid particle segregation due to sedimentation and to guarantee swarm settling behaviour of the particles during the experiments in the lab scale filter cell the solids volume concentration of the slurry was adjusted to  $c_v=0,236$ .

For the experiments themselves according to fig.4 a lab scale pressure filter with a filter area of  $A=20\text{cm}^2$  was applied like described in the VDI guideline 2762. The filter system allows to register, to store and to evaluate the filtrate flow during cake forming and cake deliquoring time as well as the gas throughput during the cake deliquoring.

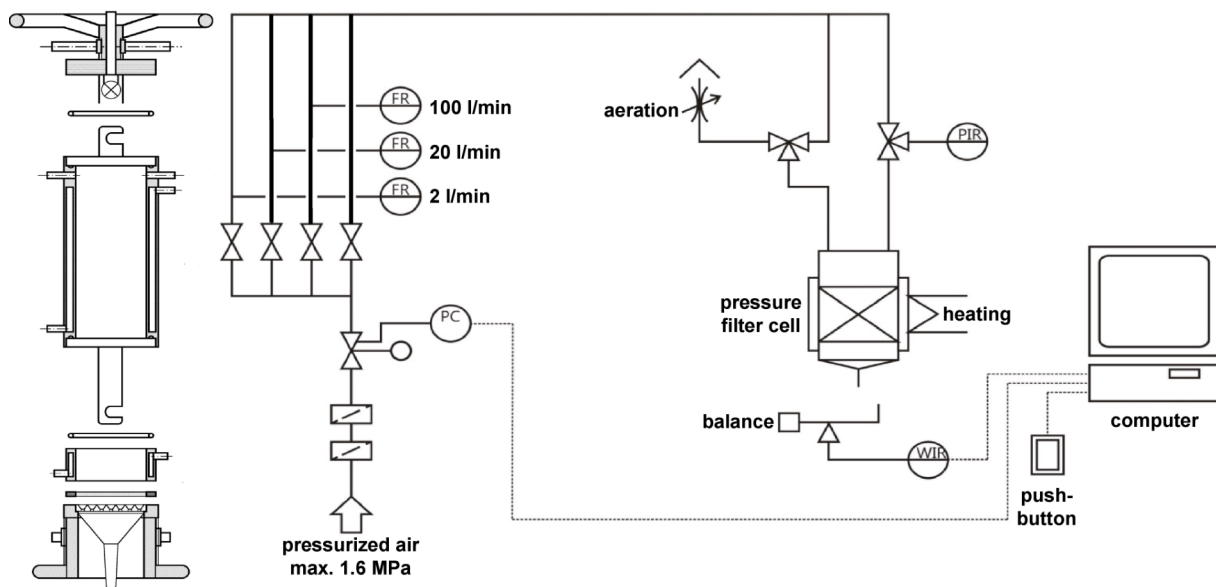


Fig.4: Lab scale filter system

To get an enough broad area to interpolate measuring data, experiments for  $\Delta p=100, 200, 300$  and  $400\text{kPa}$  and cake thicknesses of  $h_c=5, 10, 15$  and  $21\text{mm}$  have been carried out two or three times repeated each. A cake thickness of  $h_c=5\text{mm}$  was set as lower limit of a still easily removable filter cake and fixes in that way the upper limit of solids mass throughput by increasing the filters rotational speed. The deliquoring time amounted in each case  $t_2=180\text{sec}$ . Exemplarily for the results from these

measurements in fig.5 the cake formation, in fig.6 the cake desaturation and in fig.7 the gas throughput respectively the gas consumption is documented.

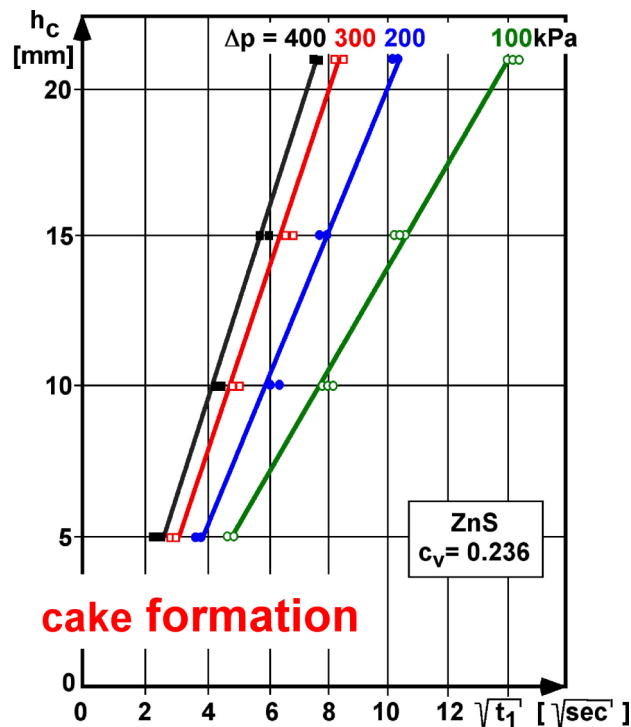


Fig.5: Results of cake formation in the lab scale

The cake thickness is plotted here against the square root of the cake formation time to make clear the validity of the correlations in eq.(5). The filter cakes turned out in the chosen pressure range to the greatest possible extent as incompressible.

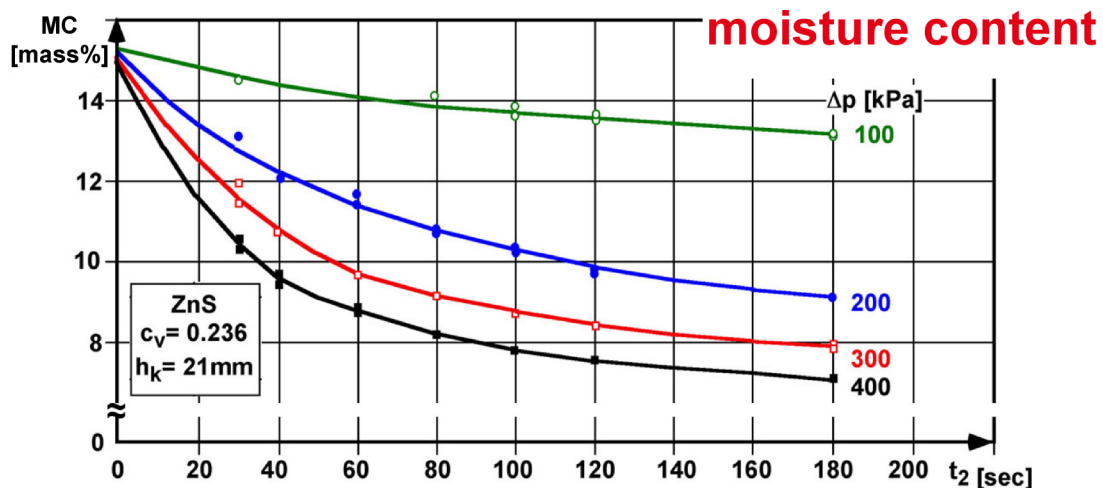


Fig.6: Results of cake deliquoring in the lab scale

The results of the cake deliquoring are documented here directly in form of the measured moisture content and not converted into the saturation degree. This conversion can be realized easily according to eq.(8). Appropriate measurements had been carried out also for cake thicknesses  $h_c=5, 10, 15$ mm. The gas consumption  $V_{g,N}(t_2)$  is calculated by integration of the deliquoring time depending gas throughput  $q_{g,N}(t_2)$ . All these data are converted to standard conditions and related to  $1\text{m}^2$  filter area. Analogous measurements had been carried out for cake heights of  $h_c=5, 15, 21$ mm.



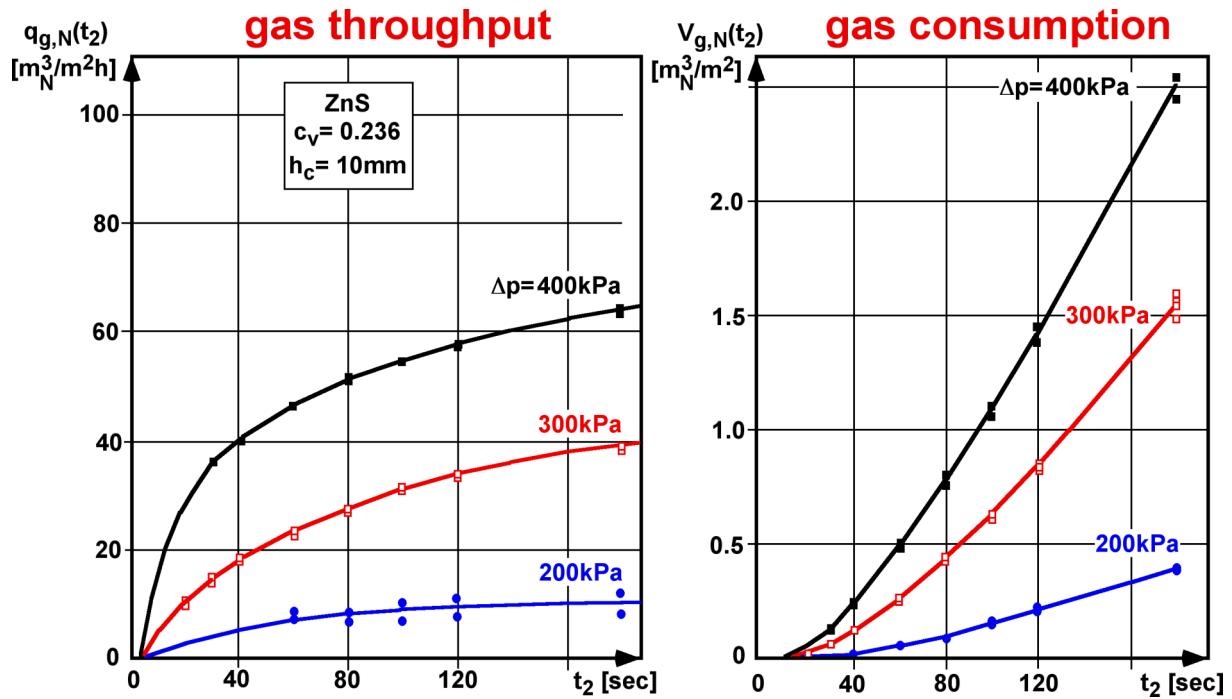


Fig.7: Results of gas flow and gas consumption in the lab scale

After these measurements a set of laboratory data was available, from which intermediary values could be generated by interpolation and which could be transferred to the conditions of a rotary filter.

#### 4. TRANSFER OF LABSCALE DATA TO A ROTARY FILTER

In the first step for the process angles  $\alpha_1$  and  $\alpha_2$  two variations had been chosen like to be made clear in fig.8.

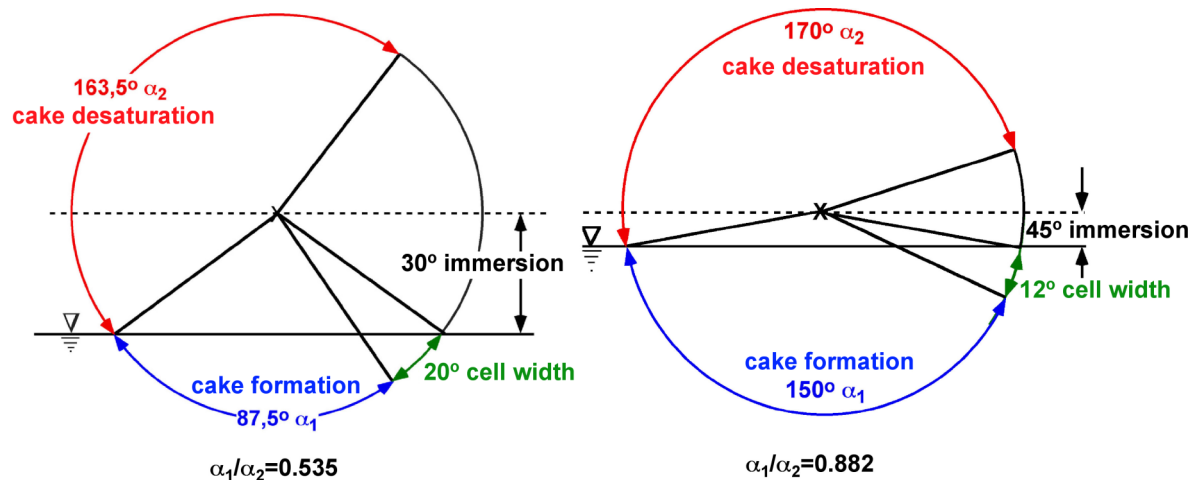


Fig.8: Angles of the rotary filters process zones

The configuration on the right side of fig.8 exhibits a considerably improved situation with regard to cake formation and cake deliquoring angle in comparison to the variation on the left side, because the filter area is much better utilized. In addition a reduction of the cell width improves the uniform distribution of the cake thickness between leading and trailing edge of the cell and enables in case of a demanded

cake washing a more sharp separation of the filtrates, which is needed for an efficient counter flow wash.

For the further evaluation firstly the more unfavourable variant of  $\alpha_1/\alpha_2=0.535$  is examined.

In the second step the rotation speed is calculated for the four cake heights from the measured cake formation times and the cake formation angle  $\alpha_1=87.5^\circ$ .

$$n = \frac{\alpha_1}{360} \cdot \frac{1}{t_1} \quad (17)$$

From these rotational speeds in combination with the known deliquoring angle the respective deliquoring times are resulting.

$$t_2 = \frac{\alpha_2}{360^\circ} \cdot \frac{1}{n} \quad (18)$$

Now the solid mass throughputs of the rotary filter can be calculated according to eq.(5) and the resulting values for the residual cake moisture content of the discharged filter cakes can be read from the diagrams like to be shown in fig. 6. The results of these operations are depicted in fig.9. The prediction from eq.(9), that the cake residual moisture content for const. pressure difference is not depending on the rotational speed of the filter can be confirmed.

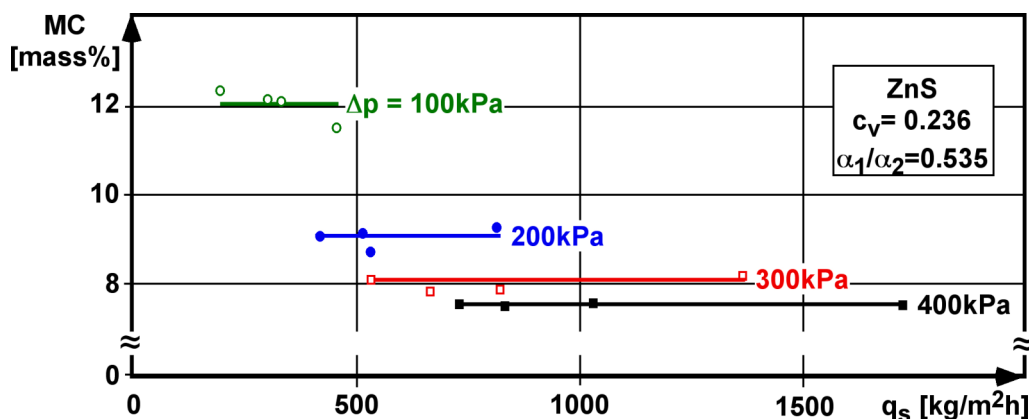


Fig.9: Residual cake moisture and solids throughput of the rotary filter

A certain target moisture can be reached not only by means of a certain pressure difference, what allows further possibilities of optimization of the separation problem with regard to an increased spec. solids mass throughput or a minimization of the necessary filter area. Fig.10 is demonstrating this for the example of a target moisture content of MC=9.1Mass% like it is reached in fig.9 for a pressure difference of  $\Delta p=200\text{kPa}$  and a ratio of process angles of  $\alpha_1/\alpha_2=0.535$ . For varied pressure differences and cake heights the slurry concentration must be adjusted now according to eq.(9) in that way, that the resulting cake moisture remains constant. To get the right slurry concentration at first the deliquoring times are taken from the experimental data, which lead to the targeted residual cake moisture for the measured pressure differences and cake heights. With these deliquoring times the rotation speed of the filter is calculated according to eq.(18). From the rotation speed accordingly eq.(17) the related cake formation time has to be determined. Now from eq.(3) the associated concentration parameter and finally from eq.(4) the necessary solids volume concentration of the feed slurry can be calculated. Under these

conditions in fig.10 the correlation between pressure difference and solids mass throughput becomes linear for each cake height. These straight lines are starting all from the same point on the ordinate. This point is according to fig.11 characterized by the capillary pressure of  $p_{cap}=136\text{kPa}$ , which enables to reach a cake moisture content of  $MC=9.1\text{mass}\%$  in the equilibrium state [8].

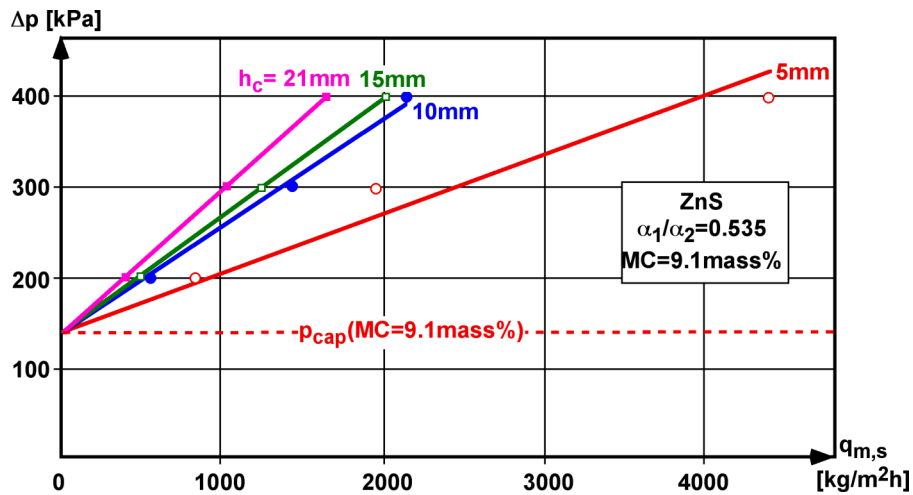


Fig.10: Pressure dependency of the filter performance for a targeted cake moisture

Further more from fig.10 the information can be taken, that for the minimal detachable cake thickness of  $h_c=5\text{mm}$  for  $\Delta p=400\text{kPa}$  an increase of spec solids mass throughput of 4:1 in comparison to the original filtration pressure of  $\Delta p=200\text{kPa}$  can be reached.

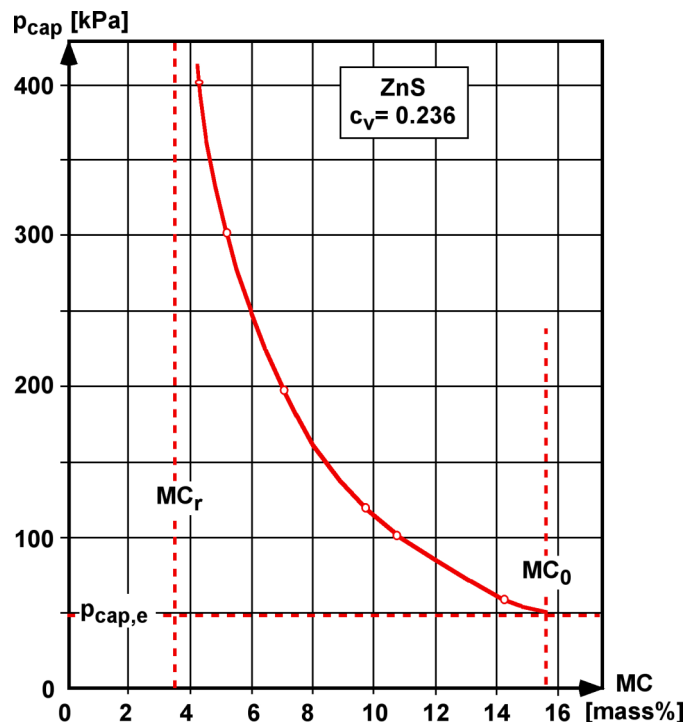


Fig.11: Capillary pressure distribution for ZnS filter cakes

Now the general question arises, how the results for the solids throughput are corresponding with the required gas throughput. For this as characterizing number the gas volume in  $\text{m}^3$  and standard conditions is related to 1kg of dry solids. To

answer the question fig.12 summates the results of all speed, pressure difference, slurry concentration and process angle variations for the rotary filter.

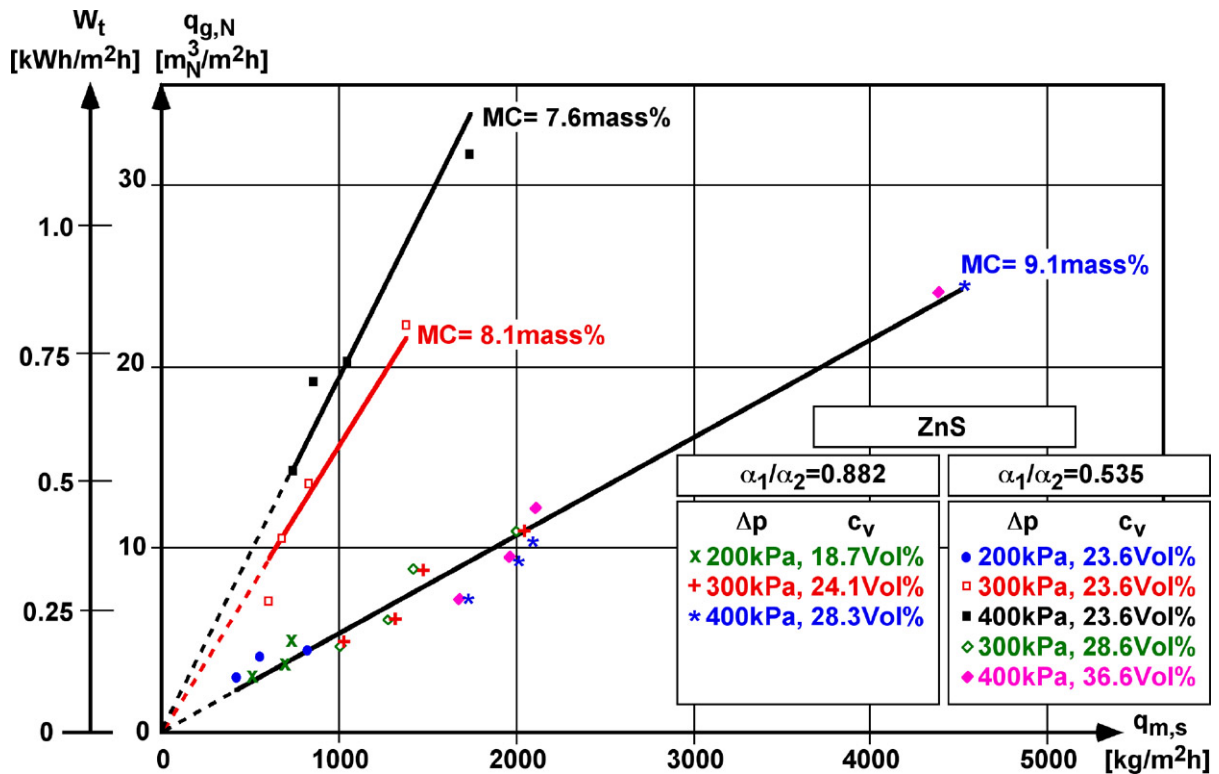


Fig.12: Gas consumption and solids throughput for a rotary filter

As general result can be observed, that for each value of residual cake moisture a linear correlation exists between spec. gas throughput and spec. solids throughput. As can be seen from the second ordinate, this linear correlation is valid in an analogous way for the technical gas compression work according to eq.(15). This means, that for constant cake moisture the ratio of spent gas volume and solids mass unit remains constant. Thus the solids throughput can be increased up to the technical limit without increase of the spec. gas demand. On the example of a targeted residual cake moisture of MC=9.1Mass% is shown, that the spec. gas consume remains constant for a variation of rotation speed, pressure difference, slurry concentration and process angles. Only in the case of changing cake moisture the spec. gas consume is also changing.

From these findings directly the question comes up, how the spec. gas consumption is correlated with the residual cake moisture content. Eq.(10) answers this question with the functional correlation between relative gas permeability and saturation degree or residual moisture content. The relative gas permeability had been experimentally determined for the particle system investigated here and is shown in fig.13. To get more supporting points for the correlation between spec. gas consumption and rel. gas permeability analogue to the representation in fig.12 further data for residual cake moisture are taken from the lab scale experiments and investigated.

These results are presented in fig.14. As expected also for the additionally chosen values of residual cake moisture linear correlations between spec. gas throughput and spec. solids throughput resulted. For better orientation the numbers for residual cake moisture content as well as the saturation degree are given here.

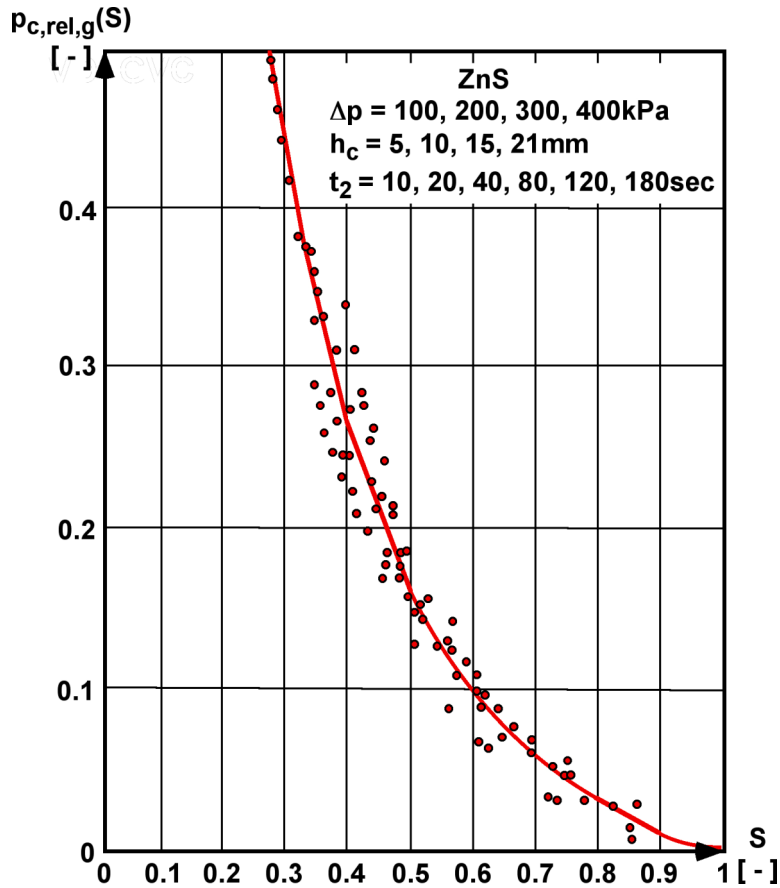


Fig.13: Relative gas permeability as function of cake saturation

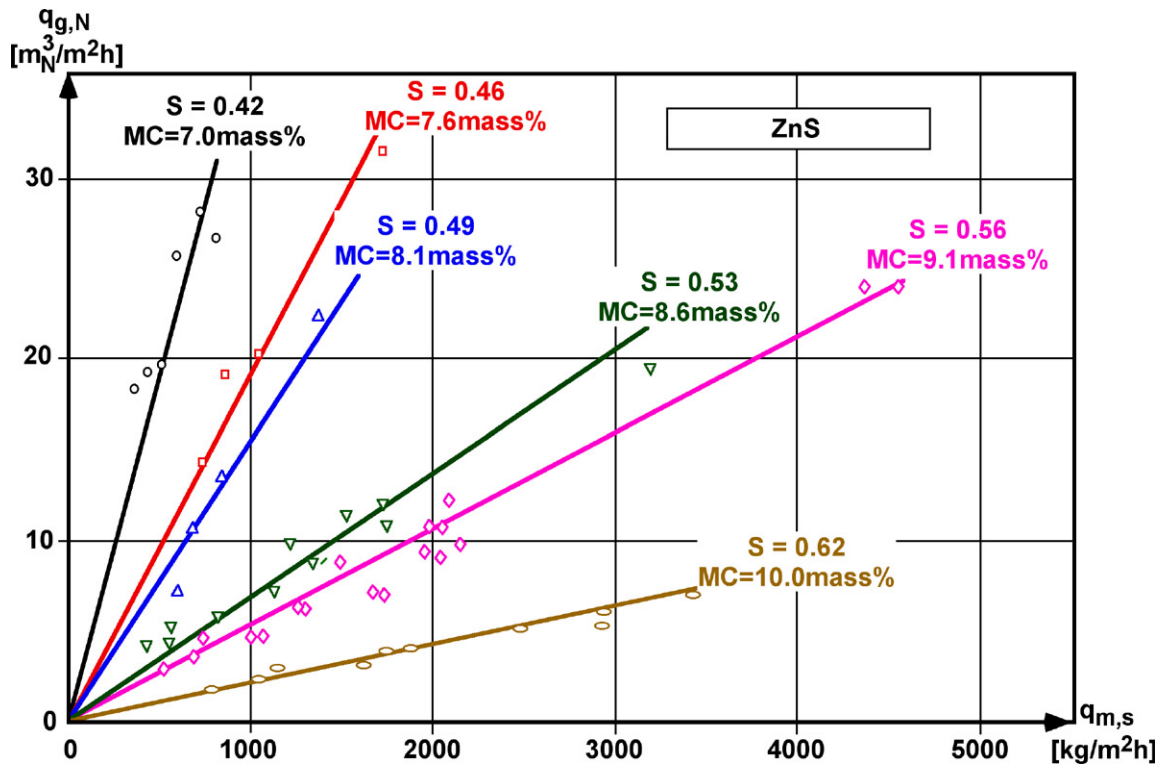


Fig.14: Gas consumption and solids throughput for a rotary filter (extended)

Each value for cake moisture is correlated with a constant spec. gas consumption and according to fig.13 a value for the rel. gas permeability. Corresponding to eq.(10)

the gas throughput is proportional to the rel. gas permeability. In fig.15 the correlation between spec. gas consumption and rel. gas permeability is demonstrated.

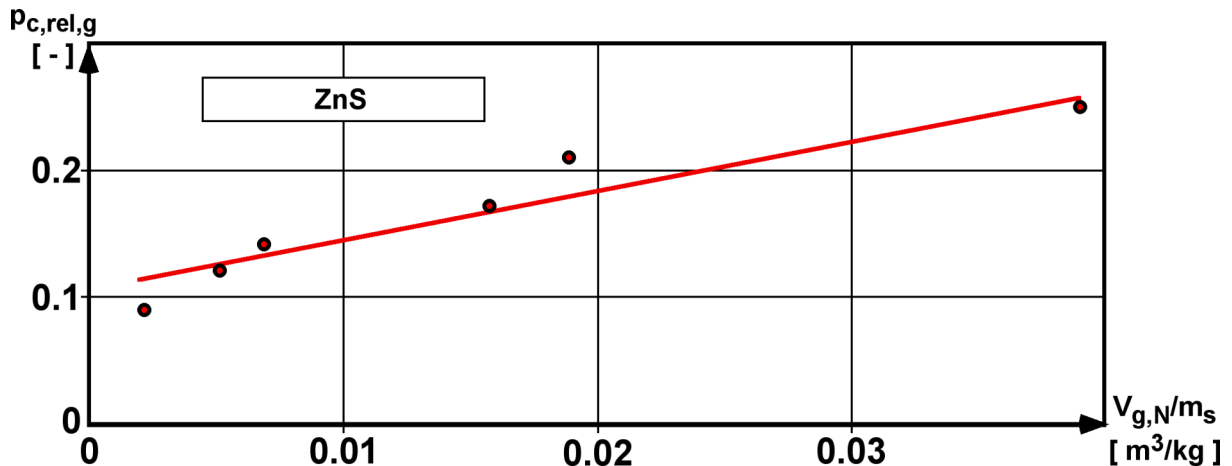


Fig.15: Spec. gas consumption for varied rel. gas permeability

It turns out a linear function and thus a confirmation of the theoretical considerations made before.

## 5. SUMMARY

On the basis of the actual state of the art to describe the kinetics of constant pressure filter cake formation, deliquoring and gas throughput and a wide spread experimental parameter study using a discontinuous lab scale filter cell the correlation of operation expenditure and process results for a continuously operating rotary filter had been investigated. The ratio of expenditure and result was defined as ratio of gas consumption and produced dry solids mass. The main result from this research work is the finding, that the solid-mass related gas consumption for all operational settings of a rotary filter remains constant, if the residual moisture content of the discharged cake remains constant. As a consequence for a certain targeted cake moisture and a given feed slurry concentration one should strive for the largest possible solids mass throughput, because then the smallest filter area is needed. This means, that for a constant ratio of expenditure and result the smallest possible filter can be applied.

## 6. NOTATION

### latin characters

A	filter area	[m <sup>2</sup> ]
$c_v$	volume concentration	[-]
$h_c$	cake thickness	[m]
$m_L$	liquid mass	[kg]
$m_s$	solids mass	[kg]
MC	moisture content	[-]
n	rotary frequency	[s <sup>-1</sup> ]
$p_c$	spec. permeability	[m <sup>2</sup> ]
$p_{c,rel,g}$	rel. gas permeability	[-]
$p_{cap}$	capillary pressure	[Pa]
$p_{cap,e}$	capillary entry pressure	[Pa]

$Q_g$	gas volume flow	$[m^3 \cdot s^{-1}]$
$q_{g,N}$	spec. gas flow (standard condit.)	$[m^3 \cdot m^{-2} \cdot s^{-1}]$
$Q_s$	solids throughput	$[kg \cdot s^{-1}]$
$q_s$	spec. solids flow	$[kg \cdot m^{-2} \cdot s^{-1}]$
$r_c$	spec. cake resistance	$[m^{-2}]$
$S$	saturation	$[-]$
$S_r$	mech.I limit of desaturation	$[-]$
$t_1$	cake formation time	$[s]$
$t_2$	cake deliquoring time	$[s]$
$V_v$	void volume	$[m^3]$
$V_s$	solids volume	$[m^3]$
$W_t$	techn compression work	$[W \cdot s]$
$x_{50,3}$	mean volume related particle diameter	$[m]$

#### greek characters

$\alpha_1$	cake formation angle	$[^\circ]$
$\alpha_2$	deliquoring angle	$[^\circ]$
$\Delta p$	pressure difference	$[Pa]$
$\varepsilon$	porosity	$[-]$
$\eta_g$	dyn.gas viscosity	$[Pa \cdot s]$
$\eta_L$	dyn. liquid viscosity	$[Pa \cdot s]$
$\eta_{verd}$	compressor efficiency	$[-]$
$\kappa$	concentration parameter	$[-]$
$\kappa$	isotropic coefficient	$[-]$
$\rho_L$	spec. liquid weight	$[kg \cdot m^{-3}]$
$\rho_s$	spec. solids weight	$[kg \cdot m^{-3}]$

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