

Through Drying of Paper - A Literature Review



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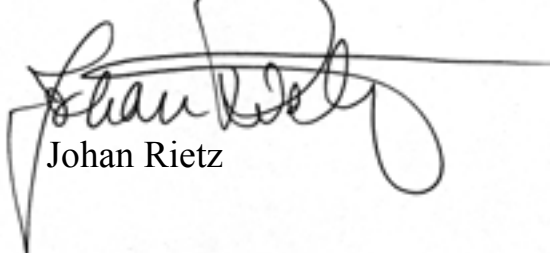
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SUMMARY

The purpose of the review is to give an overview of industrial through drying of tissue and experimental studies of through drying described in the literature. The benefits and drawbacks of through drying compared to conventional tissue drying is paid attention and some of the most extensive experimental studies on through drying are discussed. Also the transport phenomena involved in through drying of paper are discussed starting from theory originating from flow through porous media.

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1 Introduction

Over the last decades the tissue industry has seen a rapid increase in the use of through drying as a drying technique for tissue and towel grades. In the paper industry through drying is used exclusively for tissue drying because of the high operating costs (large pressure drop) associated with through drying of heavier paper grades. Until now through dryers for tissue drying have been installed mainly in the United States, while in Europe the development has been somewhat more conservative even if there has been an increase in the number of installations in recent years. In the literature the through drying process is also commonly referred to as through-air drying or TAD, however throughout this report the term “through drying” will be used.

In conventional drying of tissue and towel grades the paper web is dried on a large steam heated cylinder, a so-called Yankee cylinder, covered by a hood providing impinging hot air to the paper surface. Through drying of paper was first introduced in the late 1960's by Proctor & Gamble, USA. Until then through drying had been used predominantly for the drying of textiles and nonwovens. As the name implies, in through drying a permeable material is dried by the forced through-flow of hot air or combustion gases, which results in significantly higher drying rates than obtained with traditional drying techniques. In comparison with traditional tissue drying, the main advantage of a through drying system is the improved product quality, i.e. the higher bulk, softness and absorbency, which are achieved. A further advantage of through drying is that it virtually eliminates “one-sidedness” and cross-machine variations of the product. The positive traits of the through dried product are achieved at the cost of a relatively low dryness at the start of the thermal drying, resulting in a relatively large energy demand per unit mass of fiber. However the high bulk of the product enables the manufacturer to use a lower basis weight for a given product, which at least to some extent compensates for the large energy demand.

Because of practical reasons, most experimental studies of through drying described in the literature have been performed at conditions giving a constant flow rate through the paper. However, in industry through drying is performed such that the pressure difference is kept constant across the web, resulting in a constantly increasing air flow rate as drying proceeds and the permeability increases.

2 Industrial through drying of tissue

2.1 Extension of use

Industrial through drying of paper was introduced in the USA by Proctor & Gamble in the late 1960's. Even today most of the installed through drying machines are located in the USA, as shown by Table 2-1.1. In Europe the development has been somewhat more conservative but in recent years there has been an increase in the number of installed through drying systems.

Table 2.1-1. Number of installed through drying machines and fraction of tissue prod. for some regions.

	Number of installed through drying machines	Fraction of tissue production (%)
USA	33	20
Europe	8	8
Sweden	0	import only

2.2 Industrial setup

The tissue machine consists of a forming section, a dewatering section, a drying section and a reeling section. In conventional drying on Yankee cylinders the paper web is efficiently dewatered by pressing the web against the Yankee cylinder with a press roll (Jewitt 2000). In through drying of tissue the wet pressing of the paper web is replaced by different vacuum dewatering techniques. After dewatering the web is transferred to the drying fabric, which supports the web throughout the through drying process. After the through drying section the drying may be finished on a Yankee cylinder in order to further improve the features of the product.

In the forming section of a through drying system the paper web is formed between two fabrics by a forming box. The two fabrics are separated and the paper web follows one of the fabrics through the dewatering section. The forming fabric (the fabric which was separated from the web and the other fabric) is cleaned in water before it returns to the formation box. The web is dewatered by passage over a series of vacuum boxes before which the web is heated with steam in order to reduce the viscosity of the water and thus enhance the dewatering. The web is now transferred to the through drying fabric by a pickup roll. The final vacuum dewatering before thermal drying takes place at a so called molding box at which the web, supported by the drying fabric, is fed over a perforated box connected to a strong vacuum. Besides dewatering an equally important effect of this procedure is to draw the paper web into the fabric creating a three dimensional structure in the web. Thus in through drying the web is given a macroscopic structure very early in the process compared to conventional drying where the macroscopic structure is created through creping and

embossing after drying. The high bulk and water absorbency of through dried tissue is mainly the result of this early registration of web and fabric and not of the through-flow of air itself. The registration of web and fabric also acts efficiently to prevent shrinkage of the web during drying.

The through dryer itself normally consists of a large perforated rotary drum, up to 6.7 m in diameter and 7.6 m wide, enclosed in a hood (Polat et al. 1992a). The rotary dryer can be constructed in two principally different ways, i.e. the air flow can be directed either into or out from the cylinder. As shown in Figure 2.2-1, if the vacuum is connected to the drum the supporting fabric is running closest to the drum while in the other configuration the fabric is on top of the paper web, holding it on to the drum. As mentioned earlier, in industrial through drying the pressure difference across the web is kept constant, resulting in an increasing air flow rate through the web as drying proceeds.

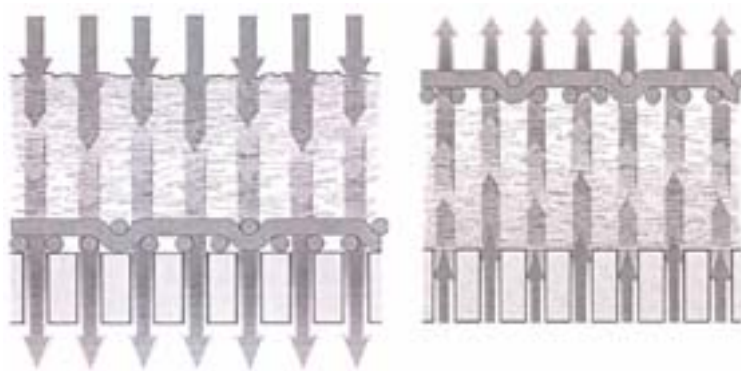


Figure 2-2.1. *Inward and outward configuration of industrial through drying.*

A through drying section normally consists of one or two rotary dryers. In some configurations the web is transferred from the drying fabric to a Yankee cylinder where the drying is finished. Drying on the Yankee cylinder gives the paper a soft and smooth surface and thus further enhances the quality of the product. By using a Yankee cylinder at the end of drying the paper quality is also improved by creping the paper from the cylinder surface, which acts to give the paper more stretch. The paper speed of industrial through dryers can be as high as 1800 m/min (30 m/s) and inlet air temperature may be as high as 430 °C (Polat et al. 1992a). Typical conditions in industrial through drying (Cui and Ramaswamy 1999) are presented in Table 2.2-1. Despite the high temperatures reported in Table 2.2-1 the temperature in through drying of tissue is generally not higher than 100-220 °C, the reason for this being the temperature limit of the fabric. On each side of the paper web there is some fabric, which is not covered by the moist paper, this part of the fabric thus heats up fast and virtually to the temperature of the drying air. The upper temperature limit for the through

drying air is thus set by the temperature limit of the fabric material. Since it is desirable to use as hot air as possible fabrics stable at high temperatures are available, but so far to a very high cost.

Table 2.2-1. Typical conditions in industrial through drying (Cui and Ramaswamy 1999).

Inlet air temperature, °C	100-400
Air flow rate, kg/(m ² s)	0.07-4.0
Superficial velocity, m/s	0.05-3.0
Basis weight, g/m ²	15-100
Initial moisture ratio, kgH ₂ O/kg D.S.	2.0-3.0
Drying rate, kg/(m ² h)	90-100

Even though the drum dryer is the most abundant construction of through dryers there are also flat bed machines available. In the flat bed design the web and fabric is transported on a conveying structure between a top and a bottom air chamber. The drying-air can be applied through either of the two air chambers depending on design and process conditions. The flat bed through dryer is mainly suited for highly permeable products since the achievable pressure difference is rather low (Polat et al. 1992a).

2.3 Energy aspects and production costs

In this section the energy aspects of industrial through drying will be discussed and compared to those of conventional drying techniques (Yankee-drying).

2.3.1 Cost of dewatering and drying

In through drying, the level of vacuum dewatering and subsequently the amount of thermal drying required are determined by two opposing costs, i.e. the cost of vacuum dewatering and the cost of thermal drying. As illustrated by Fig. 2.3-1, the cost of dewatering increases dramatically with increasing dryness and the cost of drying increases strongly with decreasing dryness, therefore the dryness after dewatering is a matter of finding a minimum in total costs of these two processes. In industrial through drying the dryness, which may be reached after dewatering, is typically 26-27%. This is somewhat lower than what is shown as the optimum dryness after dewatering presented in Fig. 2.3-1. Consequently Fig. 2.3-1 is not meant to show the optimal dryness in terms of an absolute value but it gives some understanding of how the costs of dewatering and drying are influenced by the level of vacuum dewatering. The costs giving the preferred transition point between dewatering and drying depend on many different factors such as, paper properties and the relative costs between electrical and thermal energy (Jewitt 2000).

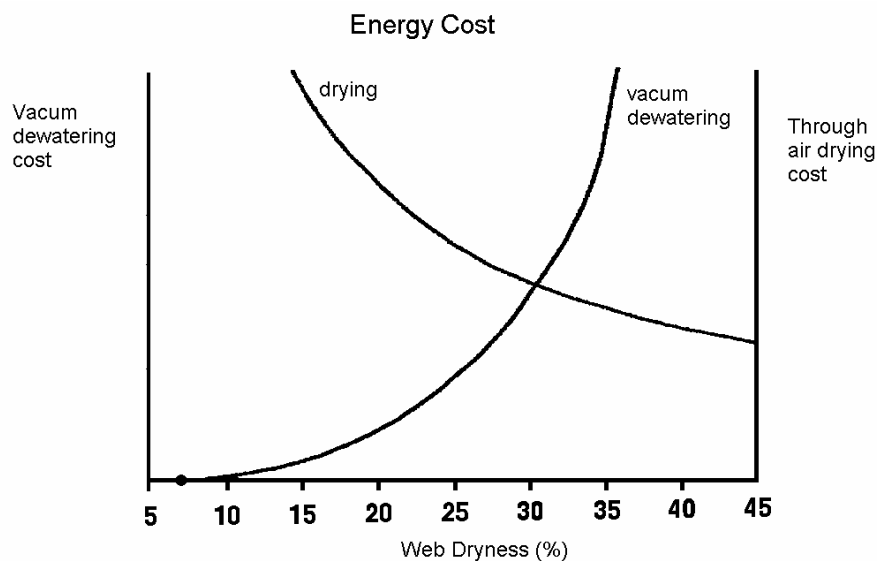


Figure 2.3-1. Energy costs for vacuum dewatering and through drying vs. dryness (Jewitt 2000).

2.3.2 Energy demand and product benefits

Because the web is dewatered by vacuum, which is connected with high costs in terms of electrical energy as discussed above, the dryness after dewatering is rather low, about 25% (3 kg H₂O/kg D.S.) compared to conventional drying with Yankee cylinders where the dryness is in the range 40-55% (0.82-1.5 kg H₂O/kg D.S.) (Jewitt 2000). Thus the total amount of water to evaporate per unit mass of fiber is much greater in through drying than in conventional Yankee drying.

The vacuum dewatering and the relatively low dryness at the start of thermal drying are the main reasons for through drying being such an energy demanding process compared to conventional drying techniques. The result however is a product of low density with superior absorbency and softness. The low density is also important from an economical point of view as it enables the manufacturer to better utilize the input of fibers by reducing the basis weight and simultaneously improving the product, i.e. produce larger amounts (surface area) of paper from each unit mass of fiber entering the process. Table 2.3-1 shows a comparison of energy demand between a conventional Yankee drying process and the through drying process. The numbers are based on typical numbers for both processes provided by Metso Paper, Karlstad, Sweden. For Yankee drying the steam consumption has been recalculated in terms of natural gas.

Table 2.3-1. Comparison between the conventional Yankee and the through drying process.

	Natural gas kJ/kg D.S.	Electricity kJ/kg D.S.	Water removed kg H ₂ O/kg D.S.	Spec. thermal energy kJ/kg H ₂ O	Spec. electr. energy kJ/kg H ₂ O	Spec. total energy kJ/kg H ₂ O
Conventional process	7500	3500	1.45	5170	2410	7580
Through drying process	20000	8000	2.82	7090	2840	9930

Table 2.3-1 clearly demonstrates the great energy requirement of the through drying process, in total 28000 kJ/kg D.S. compared to 11000 kJ/kg D.S. for the Yankee process. As mentioned previously the main reason for this is the relatively large amount of water evaporated per unit mass of fiber. At this point it is worth noting that it is suggested (Jewitt 2000) that the higher absorbency and bulk of through dried tissue makes it possible to replace a two ply toilet tissue by a one ply through dried tissue, reducing the roll weight by 26%. This example shows that from the producers' point of view it is not adequate to only look at the energy requirements per unit mass fiber, the opportunities of reducing basis weight must also be taken into account. However, from an engineering point of view a discussion about energy requirements per unit mass of fiber is still adequate. Because vacuum dewatering is the dewatering technique used in through drying the dryness after dewatering will always be relatively low for the through drying process and consequently the energy requirement is to some extent an inherent feature of the process. However, a comparison between the two processes shows that the specific thermal energy input is about 37% higher and the specific electrical energy input is about 18% higher for the through drying process. The higher specific electrical energy input is due two specific requirements of the through drying process, i.e. the energy demanding vacuum dewatering and the fan system. However the specific thermal energy for the Yankee and through drying processes is about 2.3 and 3.1 times the heat of evaporation respectively. Already the specific thermal energy requirement of the Yankee process is poor compared to a traditional cylinder dryer, where the specific thermal energy requirement is about 3000 kJ/kg H₂O (about 1.3 times the heat of evaporation). There is no apparent theoretical reason for the through drying process being even less efficient than the Yankee process, thus the poor efficiency demonstrated by the through drying process in this respect is probably due to poor energy optimization of the process. It is worth noting that by reducing the specific thermal energy requirement of the through drying process to the same level as is normal for the Yankee process, the total energy requirement would decrease with 19%, from 28000 kJ/kg D.S. to about 22600 kJ/kg D.S. The fact that some manufacturers consider the through drying process competitive enough, in spite of the great energy requirement, implies that a reduction of the total energy requirement by a magnitude as discussed above would have radical effects on the impact of through drying in tissue industry.

3 Experimental studies of through drying

3.1 *Data available in the literature*

There are two different ways in which to perform through drying experiments, i.e. maintaining a constant flow rate through the web or maintaining a constant pressure difference across the web. Because permeability increases with decreasing moisture content, a constant pressure difference across the web results in an increasing through-flow rate as drying proceeds. Normally industrial through drying is performed continuously on a perforated drum and the pressure difference across the web is kept constant. As the web dries the permeability and thus the through-flow rate increases and consequently the highest through-flow rate is found at the end of the dryer. However, most through drying experiments have so far, due to the simpler equipment design, been conducted at constant through-flow rate. With respect to previous researchers, an even less satisfying fact is that all experiments in the literature have been performed at relatively low intensities (low temperatures and flow rates) compared to industrial through drying of tissue, where the paper web is dried within a few seconds at considerably higher temperatures and air flow rates. The problem of conducting experiments at industrial conditions is mainly associated with techniques for measuring the exit air moisture content. It has been found that fast enough measurements of air moisture are not yet available, at least not to an affordable price. Chen and Douglas (1997) claim that drying experiments at constant pressure difference (variable flow rate) provide no understanding of the through drying process and thus their experiments were conducted at constant flow rate. According to Tietz and Schlünder (1993a) results from constant flow experiments can be used to predict the drying behavior at constant pressure difference, according to the method presented in section 3.2 below.

Extensive experimental studies of through drying have been performed by different researchers. Especially the work of Gummel, Tietz and Schlünder at the University of Karlsruhe, Germany and the work of Polat, Chen, Hashemi and Douglas at McGill University, Montreal, Canada should be pointed out as the most extensive studies of through drying described in the literature. Gummel (1977) presented data from about 400 through drying experiments with tissue and various textiles, all performed at constant air flow rate. In Tietz's (1992) dissertation through drying of paper was further investigated. Especially prediction of through drying at industrial conditions, i.e. through drying at constant pressure difference, by use of constant flow data was paid attention as described previously. Tietz and Schlünder (1993b) have also investigated the effect of quality enhancing additives in the drying air and showed that the additives can contribute to improve the tissue quality without being left behind in the end product. In his dissertation and in a series of papers Polat et al. (1989, 1991a, 1991b, 1992a, 1992b and 1993) presented the results of about 250 through drying experiments at constant air flow rate and paid a lot of attention to the drying rate periods and

transport phenomena, as will be discussed in later sections. Polat's data for each experiment is available in the form shown in Fig 3.1-1. Polat et al. (1989, 1992a and 1993) has also extensively discussed the applicability of Darcy's law and defined a new characteristic dimension when describing flow through paper.

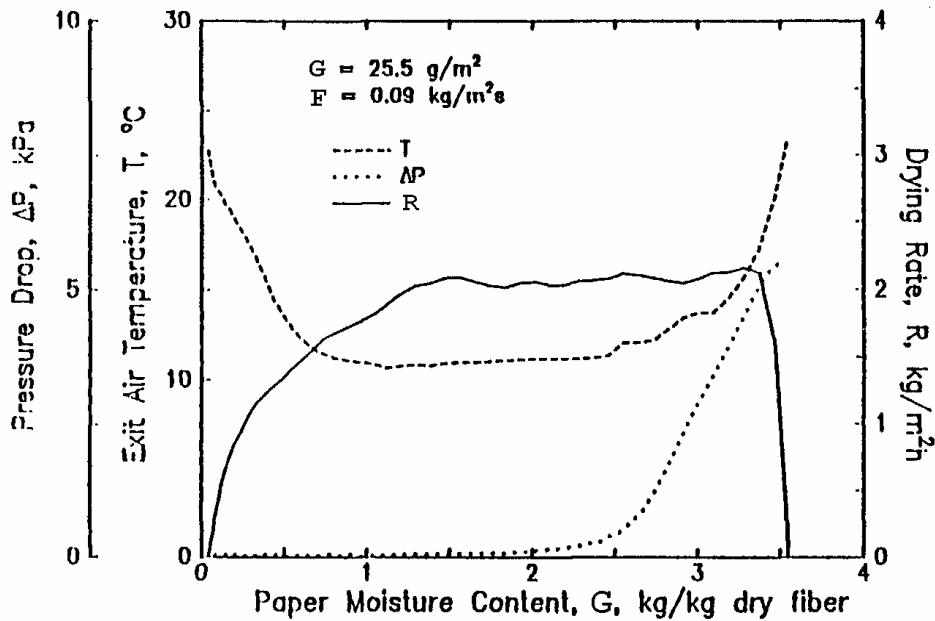


Figure 3.1-1. Example of data available from Polat (1989).

3.2 Industrial through drying – prediction from constant flow data

Most of the experimental research on through drying has for practical reasons been performed at constant air flow rate, which is not fully satisfactory even if Tietz (1992 and 1993a) has presented a method for prediction of through drying at constant pressure from these data as described in this section.

In through drying the drying efficiency (η) may be expressed as the increase in moisture ratio of the drying air as it passes through the paper web compared to the potential increase if the exiting air left saturated (Eq. 3.2-1). The drying efficiency (η) and the permeability (k) are both functions of the moisture ratio of the paper (U). Tietz and Schlünder assumed these functions to be the same for drying at both constant and variable through-flow rate. According to Tietz and Schlünder it is possible, from experiments of constant through-flow rate, to determine both these parameters as functions of paper moisture ratio, i.e. both $\eta(U)$ and $k(U)$ can be determined. The two functions are determined from constant through-flow data according to Eqs. 3.2-1 and 3.2-2.

$$\eta = \frac{X - X_{in}}{X^*(T_{WB}) - X_{in}} = f(U) \quad (3.2-1)$$

$$k = \frac{F v_{air}}{\Delta P} = f(U) \quad (3.2-2)$$

where X is the air moisture ratio, $_{WB}$ indicates the wet bulb temperature, F is the mass flow rate, v is the kinematic viscosity, and ΔP is the pressure drop across the paper web. Tietz and Schlünder claimed that the air left saturated at the beginning of drying, thus it would be more correct to use the adiabatic saturation temperature rather than the wet bulb temperature in Eq. 3.2-1.

It should be pointed out at this stage that the use of Eq. 3.2-2 (Darcy's law) when dealing with flow through paper, has been strongly criticized by Polat et al. (1989, 1992a, 1993) since it doesn't account for the inertial effects encountered even at low flow rates in flow through porous media, as will be discussed further in section 4.2.

Following the next step of Tietz and Schlünder the operating conditions for the constant pressure through drying must be known, i.e. applied external pressure difference (ΔP), inlet air temperature (T_{in}) and air moisture ratio at the inlet (X_{in}), the latter two giving the wet bulb temperature (T_{WB}).

The drying potential of the drying air may be expressed as follows:

$$\Delta X_{max} = X^*(T_{WB}) - X_{in} \quad (3.2-3)$$

The drying potential (Eq. 3.2-3) and the drying efficiency (η), which is known from constant flow data (Eq. 3.2-1) give the increase in air moisture ratio as a function of the paper moisture ratio.

$$X - X_{in} = \eta(U) \Delta X_{max} \quad (3.2-4)$$

Since the permeability is known from constant flow data (Eq. 3.2-2) the air flow rate at any given paper moisture ratio may be predicted as:

$$F(U) = k(U) \frac{\Delta P}{v_{air}} \quad (3.2-5)$$

The drying rate can now be expressed as the product of exit air moisture increase and air flow rate:

$$R = \eta(U) \Delta X_{max} F(U) \quad (3.2-6)$$

When designing a drum dryer an important factor is the required residence time, which, at a given production rate, determines the drum size. The drying rate required can be determined by solving the following differential equation:

$$\frac{dU}{dt} = \frac{F(U)}{G} \quad (2.1-7)$$

In Figs. 3.2-1 and 3.2-2 the exit air moisture increase and the drying rate as functions of paper moisture ratio, predicted by Tietz and Schlünder (1993a) according to the above method, are compared to experimental data. The exit air moisture increase as a function of paper moisture ratio was predicted rather well as illustrated by Fig. 3.2-1. However the predicted drying rate was too high during the constant rate period as illustrated by Fig. 3.2-2. In Fig. 3.2-3 it is shown that the reproducibility of drying rate as a function of time was not very good during the experiments.

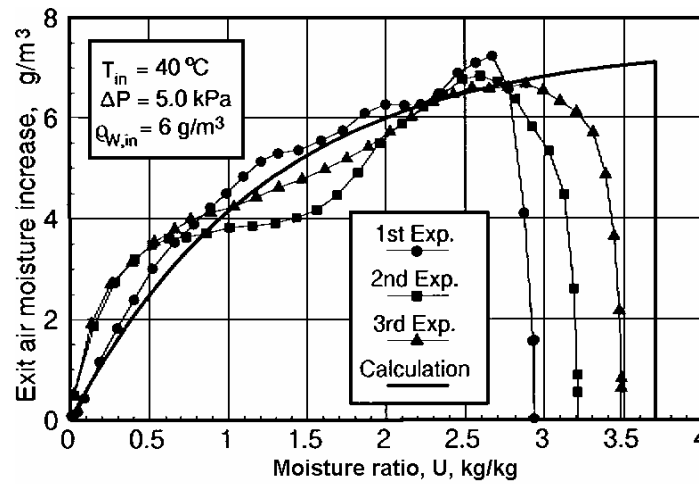


Figure 3.2-1. Predicted exit air moisture increase compared to experimental result (Tietz and Schlünder 1993).

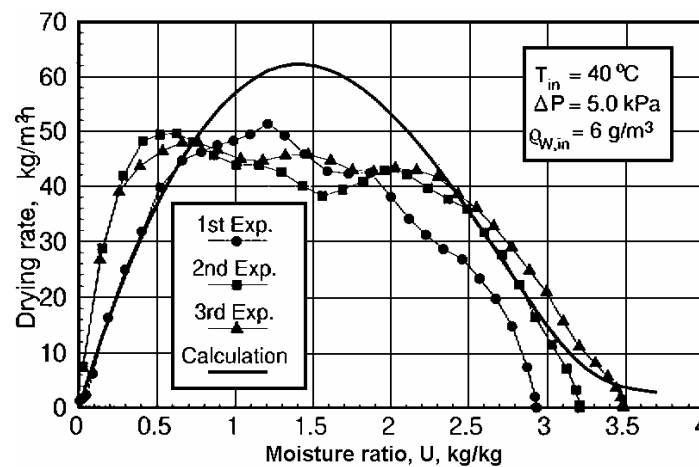


Figure 3.2-2. Predicted drying rate compared to experimental results (Tietz and Schlünder 1993).

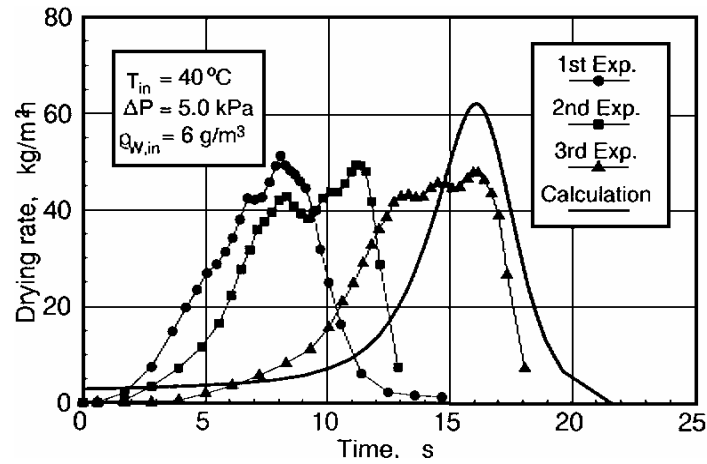


Figure 3.2.3. Drying rate vs. time – prediction compared to experimental results (Tietz and Schlünder 1993).

The research of Tietz and Schlünder also showed that in order to better predict the drying time and through-flow rate of industrial through drying, the constant through-flow experiments should be performed on paper, which has been dewatered to the initial moisture ratio by vacuum dewatering as is the case in industrial drying and not by couching, which is often the case in laboratories. According to Tietz and Schlünder, vacuum dewatering has the effect of opening pores, which in turn results in a considerable through-flow rate and thus drying rate already at the start of drying, see Figs. 3.2-4 and 3.2-5. Note that both the couched and the vacuum dewatered paper have the same initial moisture ratio. Furthermore, as shown in Fig. 3.2-5, the reproducibility of the time needed for drying was improved when the paper was vacuum dewatered.

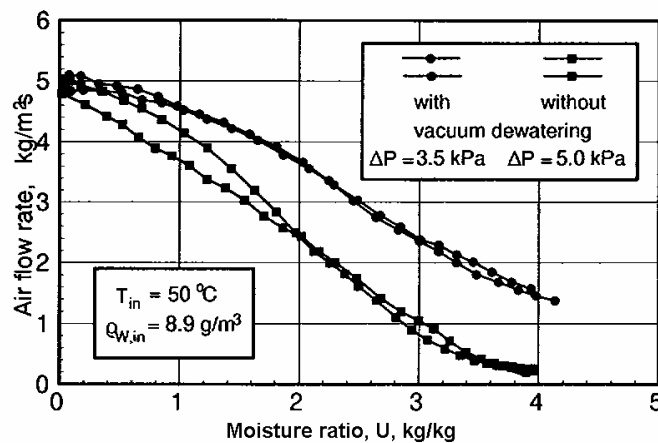


Figure 3.2-4. Effect of vacuum dewatering on air flow rate (Tietz and Schlünder 1993a).

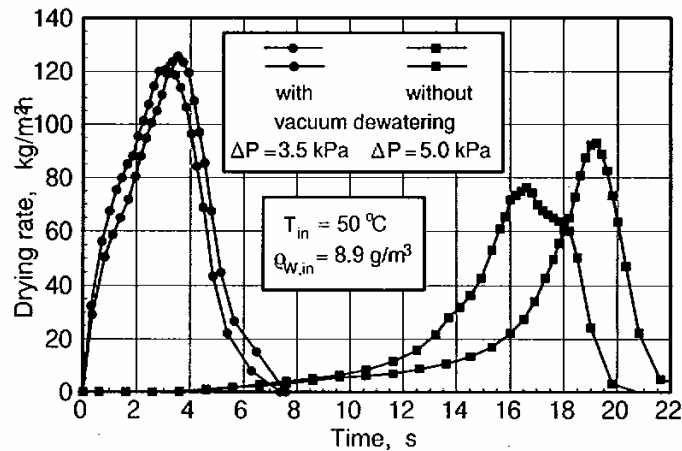


Figure 3.2-5. Effect of vacuum dewatering on drying rate (Tietz and Schlünder 1993a).

3.3 Rate periods in through drying

From experiences in contact drying, the drying process is normally considered to consist of three drying periods, i.e. periods of increasing, constant and falling drying rate. Experience has shown that in through drying the drying behavior is not as easy to characterize as in traditional drying. In their study of through drying of textiles and paper, Gummel and Schlünder (1980) found that no constant rate period in the conventional meaning was observed. However in most of the literature about through drying researchers have chosen to define a period of constant drying rate even if the transitions from increasing to constant rate and from constant to falling rate are gradual and the constant rate period exists only under low intensity drying. Laboratory experiments have shown that when drying becomes more intense, i.e. when temperature and flow rate approaches industrial conditions, the length of the constant rate period decreases and finally the constant rate period disappears (Polat 1989, Polat et al. 1991a, Chen and Douglas 1997, Tietz 1992, Tietz and Schlünder 1993a). The effect of drying intensity on the constant rate period is clearly demonstrated in Fig. 3.3-1 where it can be seen that the extent of the constant rate period is constantly reduced as the air flow rate (drying intensity) is increased.

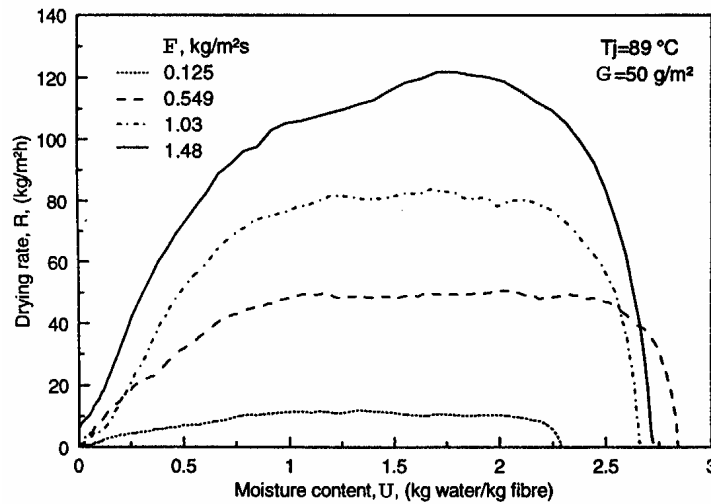


Figure 3.3-1. Effect of air flow rate on drying rate (Chen and Douglas 1997).

3.3.1 Increasing rate period

It is well known that through drying begins with a rapid increase in drying rate. According to Gummel and Schlünder (1980) and Polat et al. (1989,1991a) this is the result of free water being evaporated from the pores and the opening of new pores, leaving a larger void and internal surface area exposed to the drying air, which in turn means that the area for heat and mass transfer is constantly increasing. In through drying a substantial fraction (as much 40-50%) of the water initially present is removed during the increasing rate period (Polat 1989 and Polat et. al 1991a).

3.3.2 Constant rate period

As discussed previously, in through drying at low intensity (low temperature and flow rate) there is a period of relatively constant drying rate. Tietz and Schlünder (1993a) reported that in all their experiments the exiting air was saturated (or very close to saturated) during the constant rate period and hence they considered the period of relatively constant drying rate as thermodynamically limited rather than mass transfer controlled. Chen and Douglas (1997) state that the constant drying rate may be controlled by thermodynamics or by transport phenomena. According to Watzl and Rückert (1998) the drying rate remains constant as long as the material surface remains wet through capillary fluid transport from the interior of the material.

3.3.3 Falling rate period

Watzl and Rückert (1998) explain the falling rate period as being the result of the evaporation point receding into the interior of the fibers, causing the drying rate to be controlled by diffusion. Thus in the falling rate period, the drying rate is no longer controlled by the convective heat or mass transfer nor by thermodynamics (saturation of the drying air) but by the internal heat and mass transfer of the individual fibers. Tietz and Schlünder (1993a) claim that in through drying the falling rate period is not of classic conception, since it is due to a preferential growth of dry spots, having a higher permeability for air flow and thus resulting in a lower overall drying rate, rather than to additional resistances of dried material.

3.3.4 Drying period diagrams

In order to characterize the behavior of through drying Polat et al. (1989 and 1993a) introduced what they called drying period diagrams. The diagram was constructed in order to visualize, for a given basis weight (G) and inlet temperature (T_{in}), how the extent of the three drying periods (periods of increasing, constant and falling rate) vary with the air through-flow rate (F) and initial moisture ratio (U_0), see Fig. 3.3-2. They showed that for any combination of the variables T_{in} (inlet air temperature), F (air through-flow rate) and (paper basis weight) there exists an initial moisture ratio (U_0) under which no constant rate period exists.

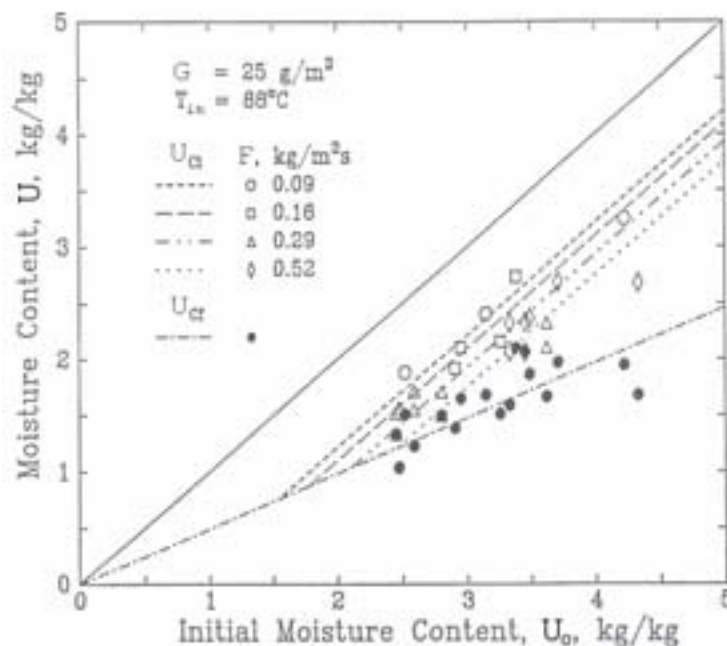


Figure 3.3-2. Drying period diagram (Polat 1989).

Polat et al. (1989, 1993a) also presented the following correlations based, on a substantial number of experiments, for the moisture ratios at the transition points from increasing to constant rate (U_{Ci}) and from constant to falling rate (U_{Cf})

$$U_{Ci} = U_0 - (T_{in} - T_s)^{0.39} F^{0.27} G^{-0.37} \quad (3.3-1)$$

$$U_{Cf} = 0.29U_0 (T_{in} - T_s)^{0.13} \quad (3.3-2)$$

where T_s is the sheet temperature. Interesting to note is that the onset of the constant rate period (U_{Ci}) was found to be sensitive to all drying variables (Eq. 3.3-1). By contrast the transition between constant and falling rate (U_{Cf}) was found to be dependant on initial moisture ratio and weakly dependent on through-flow temperature but independent of both through-flow rate and basis weight (Eq. 3.3-2).

The drying period diagrams showed that the amount of drying, which occurs in the increasing drying rate period varies significantly with drying conditions, i.e. becoming higher with increasing air temperature and through-flow rate and with decreasing basis weight. For the condition of highest intensity tested by Polat et al. (1991a), ($T_{in}=88$ °C, $F=0.52$ kg/m² s $G=25$ g/m²) and in modeling work by Polat et al. (1991b) showed that 40-50% of the water initially present was evaporated during the increasing rate period, a condition, which according to Polat et al., is possibly true even in industrial through drying.

4 Transport mechanisms in through drying

For any attempt of predicting a through drying process the understanding and adequate description of the transport mechanisms involved is crucial. In the literature most work presented on this topic is based on theory originally developed for flow through porous media. These theories were efficiently developed for well defined geometries, such as packed spheres, but application to the highly complex structure of paper provides new problems, such as determination and description of porosity, specific surface, pore size, pore size distribution, permeability and the choice of characteristic dimension in the transfer equations. The following subsections will deal with the characterization of the paper structure and the description of flow through porous media, especially its applicability to paper.

4.1 Structure of paper

Paper is a highly complex porous medium made up from wood fibers. The structure of paper is affected by the raw material, i.e. tree species and the pulping process, and by the paper formation technique. Within the structure, fibers are oriented approximately parallel to the sheet surface, in this plane however the fibers are more or less randomly orientated. The pore size distribution ranges from relatively large inter-fiber gaps to smaller pores of molecular dimensions. The behavior when dewatering or drying paper highly depends on the porous paper structure. The main parameters describing the porous structure of paper are porosity, specific surface area, pore size, pore size distribution and permeability. There are many different techniques for determining these parameters and it is important to be aware that the obtained results depend on both the method used and on other factors previously mentioned, such as pulp type and the paper forming process. As will be discussed below, for through drying applications, methods based on gas permeation are preferred due to the similarity to the studied process. Polat et al. (1992a) claim that in through drying the most important factor determining the drying rate is the air permeability, which is basically a result of the properties discussed above.

4.1.1 Porosity

When discussing flow through porous media an important variable is the porosity, which is the same as the volume fraction of gas in the material. When dealing with paper drying, where besides fibers and air also water is present, it is often convenient to define the porosity or volume fraction of gas as follows:

$$\varepsilon_g = 1 - \varepsilon_s - \varepsilon_w = 1 - \frac{G}{\rho_s z} - \frac{GU}{\rho_w z} \quad (4.1-1)$$

where ε_g , ε_s and ε_w are the volume fractions of gas, solid and water respectively, ρ_s is the fiber density, ρ_w is the water density G is the basis weight, U is the paper moisture ratio and z is the

paper thickness. In order to determine the porosity the fiber density (ρ_s), paper thickness (z), basis weight (G) and moisture ratio (U) have to be determined. The fiber density (is measured through fluid displacement techniques or mercury intrusion. Paper thickness (z) is measured by a caliper through a standard procedure because of the compressibility of paper. Basis weight is determined by weighing a piece of dry paper of known surface area. Finally the moisture ratio may be determined by weighing a paper sample before and after drying. Now all variables in the second part of Eq. 4.1-1 are known and the porosity may be calculated. Measured porosities may vary slightly depending on method due to the compressibility of paper and due to surface effects.

4.1.2 Specific surface area

In a porous material the interfacial area of the particles is generally normalized to the volume of the porous material, this area is referred to as the specific surface area. Because in the case of paper, basis weight is easier to determine than volume the specific surface area is often related to mass instead of volume. The specific surface of paper is usually determined through optical methods (based on reflectance) or solution or gas (N_2) adsorption, in each case in conjunction with the BET method (Braunauer, Emmet and Teller). By using some form of the Kozeny-Carman equation, which relates properties like porosity, tortuosity and specific surface area to permeability, it is possible to determine the specific surface area from gas and liquid permeation experiments. Due to internal pores the hydrodynamic surface area, the external surface area of fibers immersed in water, can be much higher than the surface area determined by air permeation measurements on water-swollen fibers. Polat (1989) calls this area, determined through gas permeation experiments, the effective specific surface area. Due to the nature of through drying, in through drying the specific surface area is preferably taken as the effective specific surface area determined by gas permeation experiments (Polat et al 1992a).

4.1.3 Pore size and pore size distribution

Obviously there is no experimental technique to directly describe the actual pore size (radius) of a paper structure since the shape of the pores are highly irregular and this parameter is not a single value but can be described only as an average pore size or by the pore size distribution within the structure. Methods for measuring the pore size distribution are X-ray small angle scattering, gas (N_2) sorption, vapor sorption (benzene), mercury intrusion and the dioxane gas-drive method. There is also a method, which involves passing a suspension of beads, with known distribution, through the porous material. The average pore size and the pore size distribution can be determined from monitoring the size distribution of beads that pass through. The obtained pore size and pore size distribution is highly depending on the method used. Another technique presented (Polat 1989 and Polat et al. 1992a), is based on calculating the average pore size from pressure drop data for through drying of paper. By this technique it

is possible to determine how the effective average pore size changes during drying. By naming this pore size the effective pore size it is implied that only the pores open to fluid flow are considered by this technique and likely this pore size is the most relevant to through drying of paper.

4.1.4 Permeability

The gas permeability is a measure of how through-flow of gas is related to the applied pressure difference across the paper. It is evident that gas permeability is not an independent parameter but a function of the parameters previously discussed. It has been suggested that different pores exist in paper, i.e. interconnected, dead end and non-interconnected pores. Naturally only the pores open to fluid flow (interconnected pores) contribute to gas permeability and therefore the effective pore space is the main variable affecting permeability.

4.2 Flow through paper

Flow through paper can generally be treated as flow through porous media and the general relations for this kind of flow are also applicable to paper. In flow through paper however, the crucial problem is how to adequately describe the geometric complexity of the flow channels and the dependence of these variables on moisture ratio, thus an analytical solution, which can be obtained for simple ordered structures, is not realistic for the case of paper.

There are two main ways of modeling flow through porous media, i.e. as flow inside conduits or as flow around objects. The conduit models have proven to be more applicable for the application of flow through paper (Polat et al. 1992a). Within the conduit models there are three different approaches, i.e. models based on geometry, statistics or on the solution of the complete Navier-Stokes equation. The statistical models require detailed descriptions of the web structure and thus they are not applicable to materials of very complex structure, such as paper.

4.2.1 Flow through porous media

The simplest and most common way to describe single-phase fluid flow through porous media is by Darcy's empirical law, which may be written as:

$$\frac{\Delta P}{L} = \frac{1}{k} \mu u \quad (4.2-1)$$

where ΔP is the pressure drop across the bed, L is the bed thickness, k is the Darcy permeability or simply permeability, μ is the dynamic viscosity, and u is the superficial velocity. For sufficiently low flow rates the permeability is uniquely determined by pore geometry.

How the permeability relates to the geometry of the porous structure at viscous flow is given by the following equation, which in its first form was presented by Kozeny and later modified to the form below by Carman

$$k = \frac{\varepsilon_g^3}{k_0 \left(\frac{L_e}{L} \right)^2 (1 - \varepsilon_g)^2 a_p^2} \quad (4.2-2)$$

where ε_g is the porosity (or volume fraction of gas), a_p is the specific surface, $k_0(L_e/L)^2$ is often referred to as the Kozeny constant (K), and L_e/L is the tortuosity defined as the ratio of the length of the effective flow path (L_e) to the bed thickness (L). According to Carman, a value of 5.0 for the constant K is appropriate to fit most data. However a value of $K=5.55$ have been reported to give very good agreement with experimental data from a study of flow through randomly packed plugs of cotton, wool, rayon and glass wool. According to Polat et al. (1992a), this value of K has then been used in many studies of flow through beds of cellulose fibers.

In order to improve the Kozeny-Carman relation suggestions of new porosity functions have been made. However, for the level of porosities found in paper, these functions generally give a value close to 5.55 for the Kozeny constant. Equation 4.2-2 is based on a model describing the porous media as consisting of channels of definite length but variable cross section. Further it is assumed that there exists no tangential component of the fluid flow, that the porosity is lower than 0.80, and that all pores are randomly distributed and open to flow. Equation 4.2-2 should be used with caution for systems made up from particles, which deviate strongly from spherical or with particles of wide size distribution or for consolidated media, specifications that all apply very well to paper.

Reynolds in 1900 proposed the following equation, in which the resistances to flow is considered as consisting of two parts, i.e. the viscous drag at the surface of particles and the loss due to turbulent eddies and at sudden changes in the cross section of channels:

$$\frac{\Delta P}{L} = \alpha \mu u + \beta \rho u^2 \quad (4.2-3)$$

Today Eq. 4.2-3 is known as the Forchheimer relation, even though first presented by Reynolds.

The Forchheimer equation has also been derived by solving the complete Navier-Stokes equation. Using a volume averaging theorem Dullien and Azzam's solution gave the following equations for α and β (Polat et al. 1992a):

$$\alpha = \frac{1}{d_p^2 V} \iiint \nabla^2 \mathbf{v} dV, \quad \beta = \frac{1}{d_p V} \left(\iiint \mathbf{v} \nabla \mathbf{v} dV + \iint \frac{P}{\rho u^2} \mathbf{n} dA \right)$$

where d_p is the pore diameter, V is the averaging volume, \mathbf{v} is the interstitial velocity vector, \mathbf{n} is a unit vector, u is the superficial velocity and A is the surface area of the averaging volume. By accounting for the average and fluctuating components of velocities and pressures the solution of Ahmed and Sunada gave the following equations for α and β :

$$\alpha = \frac{1}{d_p^2 \varepsilon_g V_b} \iiint \nabla^2 \mathbf{v} (\varepsilon_g dV_b), \quad \beta = \frac{1}{d_p \varepsilon_g V_b} \left(\iiint (\mathbf{v} \nabla \mathbf{v} - \mathbf{v}' \nabla \mathbf{v}') (\varepsilon_g dV_b) \right)$$

where ε_g is the porosity of the material, V_b is the bulk volume and \mathbf{v}' is the fluctuating component of the interstitial velocity vector.

In 1937 Missbach presented the following equation

$$\frac{\Delta P}{L} = au^n \quad (4.2-4)$$

where the value of the exponent n generally lies between 1 and 2. When $n=1$, Eq. 4.2-4 is equivalent to Darcy's law (Eq. 4.2-1), with $a=-\mu/k$. With $n=1$, a may also be interpreted as the combined constant $\alpha\mu$ in the viscous part of the Forchheimer relation (Eq. 4.2-3). With $n=2$, a is equivalent to the combined constant $\beta\rho$ in the part of the Forchheimer relation which is due to turbulent losses. When the value of n is somewhere in between the limits 1 and 2, the interpretation is somewhat less obvious and a simply acts as a regression constant in the description of combined viscous and inertial effects.

The Forchheimer relation (Eq. 4.2-3) has been modified in many different ways of which the Ergun equation (Eq. 5.2-5) for both viscous and turbulent flow is the best known. The constants 150 and 1.75 in this equation were determined for flow through ring packings (Richardson and Harker, 1991).

$$\frac{\Delta P}{L} = \left(\frac{150}{d_p^2} \right) \left(\frac{(1-\varepsilon_g)^2}{\varepsilon_g^3} \right) \mu u + \left(\frac{1.75}{d_p} \right) \left(\frac{(1-\varepsilon_g)}{\varepsilon_g^3} \right) \rho u^2 \quad (4.2-5)$$

At very low flow rates the second-order term of the Ergun equation approaches zero and the equation reduces to the well-known Kozeny-Carman or Blake-Kozeny equation. At very high flow rates, the second order term dominates and the equation reduces to the so-called Burke-Plummer equation. At purely viscous flow, the constant 150 corresponds to a Kozeny constant (K) of 4.16 in Eq. 4.2-2.

4.2.2 Inertial effects and turbulent flow

The application of Darcy's law has been the predominant way of describing air flow through paper, i.e. flow through paper has generally been considered as being purely viscous. However, later research has shown that when pinholes become evident the flow is no longer purely viscous and thus Darcy's law does no longer apply. According to Polat et al. (1989 and

1992a) since the findings that flow through thin paper is not always purely viscous, many researchers have instead of treating this fact, made sure to perform their experiments at viscous flow, i.e. by increasing paper thickness or decreasing flow rate.

It has been found that Darcy's law is applicable only at very low Reynolds numbers and that it breaks down already at Reynolds numbers, which would normally correspond to viscous flow (Polat 1989, Polat et al. 1992a and 1993). Polat et al. (1989, 1991a and 1993) clearly demonstrate that air flow through paper cannot be treated as purely viscous even at a flow rate of 0.6 m/s for 150 g/m² paper, see Table 4.2-1 where the exponent n of Eq. 4.2-4 is presented for flow through dry paper of different basis weights. The results presented in Table 4.2-1 also indicate that when the air is replaced by helium, which has a kinematic viscosity about 7.5 times higher than that of air, the inertial contribution to the pressure drop decreases. It is however evident that even for helium flow there is a substantial inertial contribution to the pressure drop for the lowest basis weight (25 g/m²).

Table 4.2-1. Values of the exponent n in Eq. 4.2-4 for flow through dry paper (Polat 1989)

	Basis weight, g/m ²				
	25	50	100	150	250
Air flow	1.24	1.07	1.05	1.02	1.01
Helium flow	1.08	1.01	1.00	1.00	1.00

It has been claimed that for moist tissue paper the use of Darcy's law may give an error in permeability of as much as 50% for low basis weights and high through-flow rate (Polat et al. 1992a). The Reynolds number at which Darcy's law is no longer applicable ranges between 0.1-75 depending the porous structure and the choice of characteristic dimension used (Polat et al. 1992a). Since the transition from viscous to turbulent flow should occur at an even higher Reynolds number the breakdown of Darcy's law must be due to some other process than the transition from viscous to turbulent flow. Polat et al. (1992a) explain the brake down of Darcy's law as being associated with inertial effects occurring when the streamlines of the flowing medium are distorted due to changes in direction of motion big enough for inertial forces to become significant compared to viscous forces. The mechanisms of losses due to inertial effects at relatively low Reynolds numbers and due to turbulent losses are identical. The mechanism behind inertial effects in porous media differs from the effects of turbulence only in that the change in direction of motion is induced by the structure of the porous medium rather than being the result of directional change due to turbulent eddies. Because of this mechanistic similarity, flow through paper in regions where inertial effects are important but the flow is not yet turbulent can be described by the same equations as those used for turbulent flow through porous media. Thus the Forchheimer relation (Eq. 4.2-3) and the

equation of Missbach (Eq. 4.2-4) are both in principal applicable to flow through paper. The only difference is the interpretation of the second order term in the former and the deviation from unity of the exponent, n in the latter, which for flow through paper at moderate Reynolds numbers should be interpreted as the effect of inertial forces due to deflections in the structure rather than describing the effect of inertial forces due to turbulent eddies.

4.2.3 Characteristic dimension

Because of the highly random structure of paper one of the main problems when dealing with transport phenomena in paper is the definition of the characteristic dimension needed for the Reynolds number. In unconsolidated beds it is natural to use a dimension based on particle size. In consolidated beds the two most widely used characteristic dimensions are the reciprocal of specific surface and the square root of permeability. Other characteristic dimensions have been suggested in the literature.

The characteristic dimension, $d_p = \beta/\alpha$, where α and β are the viscous and inertial parameters of the Forchheimer equation. (Eq. 4.2-3), has been suggested for flow through dry and moist paper (Polat 1989 and Polat et al. 1992a). Table 4.2-2 shows that the characteristic dimensions $1/a_p$ and $k^{0.5}$ are in range of intra-fiber pore size rather than inter-fiber pore size even though both a_p and k were determined by through-flow techniques and thus should relate to the size of inter-fiber pores. In contrast, the characteristic dimension $d_p = \beta/\alpha$ in Table 4.2-2 has been found to coincide better with the size range of the inter-fiber pores studied on electron micrographs (Polat 1989 and Polat et al. 1992a).

Table 4.2-2. Characteristic dimensions of dry paper (Polat 1989)

Basis weight, g/m ²	$1/a_p$, μm	$k^{0.5}$, μm	β/α , μm
25	1.32	1.23	16.8
50	0.96	0.88	5.4
100	0.84	0.76	5.0
150	0.75	0.72	4.7
250	0.69	0.59	4.2

Polat (1989 and 1992a) also showed how the characteristic dimension β/α changes as drying proceeds, see Fig. 4.2-1. At the beginning of drying when only pinholes are open to flow the characteristic dimension is much bigger than at the end of drying when many smaller pores are also open to flow and the average pore size is lower. The dependence of d_p on moisture ratio was also shown to agree with measurements by scanning electron microscopy.

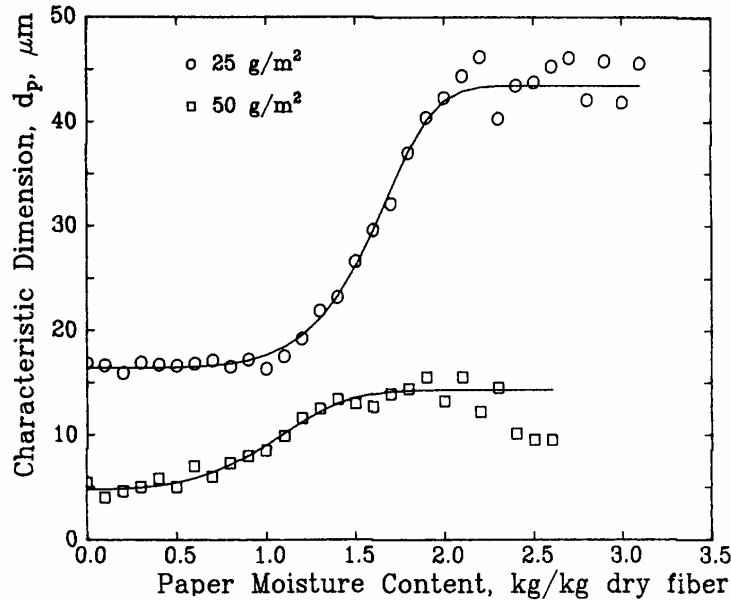


Figure 4.2-1. Characteristic dimension, β/α as a function of moisture ratio.

4.2.4 Friction factor

An alternate way of describing the pressure drop associated with flow through porous media is by introducing a dimensionless friction factor, which is a function of the Reynolds number, as commonly done for flow through tubes. For flow through porous media this friction factor is generally defined analogously to the friction factor for flow through tubes

$$f = \frac{\Delta P}{L} \frac{d_p}{2\rho u^2} \quad (4.2-6)$$

where f is the fanning friction factor. Since there are different definitions of the characteristic dimension d_p , subsequently there are many varieties of the above definition.

Other definitions of the friction factor are for example the rearrangement of the Ergun equation (Eq. 4.2-5) on the dimensionless form below.

$$f = \frac{\Delta P}{L} \frac{d_p}{2\rho u^2} \frac{\varepsilon_g^3}{1 - \varepsilon_g} = 1.75 + 150 \frac{1 - \varepsilon_g}{\text{Re}} \quad (4.2-7)$$

Other work has dealt with trying to improve the porosity function of the Ergun equation but since no satisfying results for paper have been found, the analysis of flow through paper has generally been treated using the Darcy permeability, a treatment that is not appropriate at the high flow rates found in industrial through drying as discussed in section 4.2-2.

Another definition of the friction factor may be derived from the Forchheimer relation (Eq. 4.2-3), which may be written on dimensionless form to obtain the following equation.

$$\frac{\Delta P}{L} \frac{1}{\beta \rho u^2} = 1 + \frac{\alpha \mu}{\beta \rho u} \quad (4.2-8)$$

With $d_p = \beta/\alpha$, $G = \rho u$, the definition of Reynolds number and recognizing that the left hand side of Eq 4.2-8 is the definition of the friction factor (ratio of total energy loss to the kinetic energy loss) Eq. 4.2-8 becomes:

$$f = 1 + \frac{1}{Re} \quad (4.2-9)$$

Polat et al. (1989 and 1992a) showed that their data for moist and dry paper could be represented very well by Eq. 4.2-9 as shown by Fig. 4.2-2. Polat considered $Re = 0.05$ as the upper limit for Darcy's law approximation since at $Re > 0.05$ the deviation from Darcy's law was found to be more than 5%.

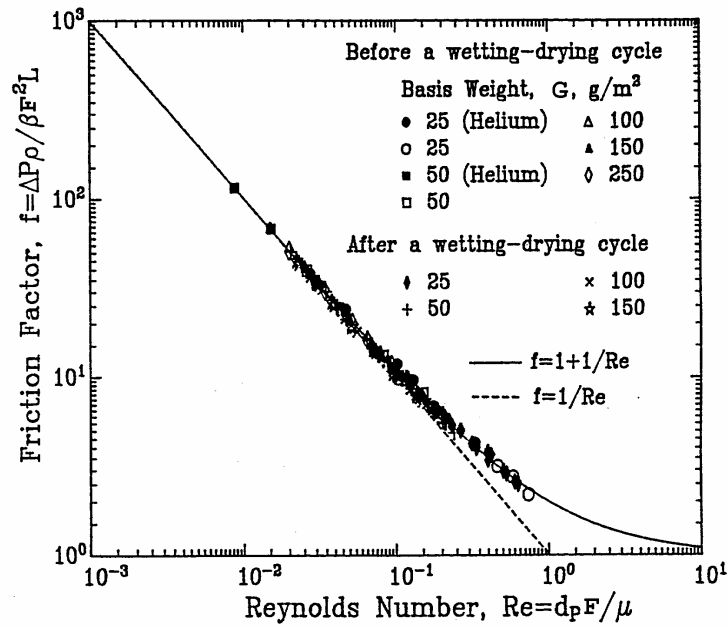


Figure 4.2-2. Pressure drop across dry paper (Polat 1989)

5 Nomenclature

Variables

a_p	specific surface	m^2/m^3
F	mass flux	$\text{kg}/(\text{m}^2\text{s})$
G	basis weight	kg/m^2
k	permeability	m
k_0	a constant	-
L	bed thickness	m
ΔP	pressure drop	Pa
R	drying rate	$\text{kg}/(\text{m}^2\text{s})$
t	time	s
T	temperature	$^{\circ}\text{C}$
u	superficial velocity	m/s
U	paper moisture ratio	kg/kg
X	air moisture ratio	kg/kg
ε	volume fraction	-
η	drying efficiency	-
μ	dynamic viscosity	$\text{kg}/(\text{ms})$
ν	kinematic viscosity	m^2/s
ρ	density	kg/m^3

Subscripts

0	initial
air	air
ci	indicates transition point between increasing and constant rate periods
cf	indicates transition point between constant and falling rate periods
e	effective
g	gas
in	inlet
max	maximum
S	sheet
s	solid (fiber)
w	water
WB	wet bulb

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