BACKMIXING IN SCREW EXTRUDERS

Chris Rauwendaal, Rauwendaal Extrusion Engineering, Inc.
Paul Gramann, The Madison Group

Abstract

Mixing is a critical function in most extrusion operations. One of the most difficult mixing tasks is backmixing. An extrusion operation where good backmixing is very important is when a low percentage color concentrate, CC, is added to a virgin polymer. In this case, the initial distance between the CC pellets may be 100 mm or greater. If the final striation thickness needs to be reduced to the micron level, the reduction of the striation thickness needs to be at least five orders of magnitude – this is quite a tough task!

This paper will analyze how the velocity profiles, axial mixing, and residence time distribution are related. It will be shown why simple conveying screws have poor axial mixing capability. New mixer geometries that are specifically designed to improve backmixing will be discussed. Results from extrusion experiments will be presented.

Cross Sectional Mixing and Axial Mixing

Most analyses of mixing focus on cross sectional mixing, e.g. (1). The cross sectional mixing is determined mostly by the Couette shear rate between the rotating screw and stationary barrel. Using the flat plate approximation and assuming pure drag flow, this shear rate can be approximated by:

$$\dot{\gamma} = \frac{\pi DN}{H} \qquad (1)$$

where D is the barrel diameter, N the rotational speed, and H the channel depth.

More accurate expressions for the shear rate have been developed using cylindrical coordinates (2). Typical values of the Couette shear rate in single screw extruders range from 50 to 100 sec^{-1}. With a typical residence time in the melt conveying zone of about 20 seconds, the resulting total shear strain ranges from about 1,000 to 2,000 units. This means that the striation thickness in cross sectional mixing is reduced by about three orders of magnitude. In many cases, this is not enough to achieve a level of mixing that appears uniform to the naked eye.

Axial mixing or backmixing occurs by pressure flow (3). For pressure flow of a power law fluid between parallel plates, the dimensionless velocity $\phi = v/v_{max}$ can be written as a function of the dimensionless normal coordinate $\xi = 2y/H$ as follows:

$$\phi(\xi) = 1 - \xi^{\frac{n+1}{n}} \qquad (2)$$

Figure 1 shows the velocity distribution for several values of the power law index. The velocity profile for a Newtonian fluid (n=1.0) is a parabola. The shear rate in the center of the channel is zero. As a result, no mixing will take place there. The center region flattens as the power law index reduces. In other words, the velocity profile becomes closer to a plug flow profile as the power law index approaches zero. This means that the low shear rate region expands as the fluid becomes more shear thinning. Thus, the region with poor mixing becomes larger when the power law index reduces.

From this simple analysis it becomes clear that the situation for backmixing is substantially more difficult than for cross sectional mixing. For shear thinning fluids there is a considerable region in the center of the channel where little or no axial mixing takes place.

Residence Time Distribution

The RTD can be determined from the velocity profiles in the channel. The axial flow in a screw extruder is a pressure flow because there are no axial velocity components of the screw or barrel. As a first approximation, the axial pressure flow can be considered a flow between parallel plates. The velocity profile for pressure flow of a power law fluid between parallel plates is (4):

$$\dot{V} = \frac{nWH^2}{2(1+2n)} \left(\frac{H \Delta P}{2mL} \right)^{\frac{1}{n}} \qquad (3)$$

Where W is the width of the channel, H the height of the channel, ΔP the pressure drop over axial length L, n the power law index, and m the consistency index of the fluid.

The velocity profile can be written as:

$$v(y) = \frac{nH}{2(1+n)}\left(\frac{H\Delta P}{2mL}\right)^{\frac{1}{n}}\left[1-\left(\frac{2y}{H}\right)^{1+\frac{1}{n}}\right] \qquad (4)$$

where normal coordinate y ranges from –H/2 to +H/2.

The velocity profile can be written as:

$$v(y) = v_{\max}\left[1-\left(\frac{2y}{H}\right)^{1+\frac{1}{n}}\right] \qquad (5)$$

The external RTD function *f(t)dt* can be determined from:

$$f(t)dt = \frac{d\dot{V}}{\dot{V}} = \frac{2(2n+1)}{(n+1)H}\left[1-\left(\frac{2y}{H}\right)^{1+\frac{1}{n}}\right]dy \qquad (6)$$

Coordinate y can be expressed as a function of time by the following substitution:

$$t = \frac{L}{v(y)} \qquad (7)$$

With this substitution, the external RTD function can be written as:

$$f(t)dt = \frac{n(2n+1)}{(n+1)^2}\left(1-\frac{t_0}{t}\right)^{\frac{-1}{n+1}}\frac{t_0^2}{t^3}dt \qquad (8)$$

The minimum residence time t_0 can be determined from:

$$t_0 = \frac{L}{v_{\max}} \qquad (9)$$

The cumulative RTD function *F(t)* can be found by integrating the external RTD function; it can be expressed as:

$$F(t) = \left(1-\frac{t_0}{t}\right)^{\frac{n}{n+1}}\left(1+\frac{n}{n+1}\frac{t_0}{t}\right) \qquad (10)$$

For a Newtonian fluid the RTD function becomes:

$$F(t) = \left(1-\frac{t_0}{t}\right)^{\frac{1}{2}}\left(1+\frac{t_0}{2t}\right) \qquad (11)$$

If we express the RTD as a function of the dimensionless residence time θ, where θ is the actual residence time divided by the mean residence time, we get:

$$F(\theta) = \left(1-\frac{n+1}{(2n+1)\theta}\right)^{\frac{n}{n+1}}\left(1+\frac{n}{(2n+1)\theta}\right) \qquad (12)$$

The expression above for a power law fluid has not been published before. With this expression the RTD can plotted at several values of the power law index n, see figure 2. It is clear from figure 2 that the RTD becomes narrower as the value of the power law index reduces. This means that backmixing reduces as the fluid becomes more shear thinning (lower power law index). Figure 2 confirms what we have already seen in the velocity profiles of figure 1. As the fluid becomes more shear thinning, the velocity profile becomes closer to plug flow and, consequently, the RTD becomes narrower and backmixing more problematic.

RTD in Screw Extruders

Pinto and Tadmor (5) developed expressions for the RTD in single screw extruders. The cumulative RTD function can be written as:

$$F(t) = F(\xi) = 0.5\left(3\xi^2 - 1 + (\xi - 1)\sqrt{1+2\xi-3\xi^2}\right) \qquad (13)$$

The dimensionless time θ (time divided by mean residence time) can be expressed as a function of the dimensionless normal coordinate ξ:

$$\theta = \frac{3\xi - 1 + 3\sqrt{1+2\xi-3\xi^2}}{6\xi(1-\xi+\sqrt{1+2\xi-3\xi^2})} \qquad (14)$$

The two expressions above were derived for a Newtonian fluid using the flat plate approximation considering both down- and cross-channel velocity components. Figure 3 shows the RTD for a single screw extruder as well as for pressure flow of a Newtonian fluid between flat plates. The single screw RTD is narrower than the flat plate RTD because of the recirculation of the fluid in the screw channel. Fluid spends more time in the lower portion of the channel ξ=0-2/3 than in the upper portion of the channel ξ=2/3-1.

In order to properly determine the RTD of an extruder we have to consider not only down- and cross-channel velocity components but also normal velocity components that occur at the flight flanks. This will require a numerical analysis, either FDA, FEA, or BEA. Even though the depth of the channel is usually quite small compared to the channel width, the residence time at the flight flanks is substantial because the normal velocities are quite small. Joo and Kwon (6) pointed out limitations of the Pinto analysis. As one would expect, the Pinto model underpredicts the residence times relative to a full three-dimensional analysis, particularly with large values of the axial pressure gradient.

The Pinto RTD for a single screw extruder is narrower than the RTD for pressure flow between flat plate for a Newtonian fluid. The effect of shear thinning is to further narrow the RTD as discussed earlier. These two effects explain why backmixing is such a critical issue in screw extruders.

Methods to Improve Backmixing

A major concern in backmixing is the fluid in the center region of the screw channel where the axial shear strain is zero or close to zero. In a simple conveying screw the fluid in the inner recirculation region will stay within this region until it reaches the end of the screw. When this happens, the material flowing into the die will be poorly mixed.

Mixing pins and slots in the screw flights will improve axial mixing because they achieve a short term splitting and reorientation of the fluid. The effect of mixing pins on backmixing is illustrated in figure 4. These results were obtained using a three-dimensional BEM flow analysis. It is clear that one row of mixing pins has limited effect on axial mixing. Backmixing can be improved by varying the spacing between the pins, thus, intentionally creating streams of different axial velocities.

The challenge in improving axial mixing is to efficiently transfer fluid from the inner recirculation region to the outer region and vice versa. A simple but effective method of doing this is the inside-out mixer shown in figure 5. The flight in this mixer is offset so that the material in the center region is cut by the offset flight and then pushed to the screw and barrel surfaces by the normal pressure gradients that occur at the flight flank. Results of particle tracking using BEA are shown in figure 6.

The redistribution of the material is shown in figure 7. It shows how the fluid from the center region is cut by the offset flight and pushed to screw surface at the pushing side of the flight and to the barrel surface at the trailing side of the flight. Figure 7 shows the redistribution of the fluid at several axial locations from the flight offset. It is clear that a considerable axial distance is necessary to bring about the redistribution of the material.

Conclusions

Backmixing is one of the most difficult mixing tasks in screw extruders. This is due to the fact that the axial strain rates in the extrusion process are very low, particularly in the center region of the channel. The axial velocity profile is close to plug flow, particularly for strongly shear-thinning fluids. Backmixing problems are especially severe when adding a small percentage concentrate to the extruder. When both the polymer and concentrate are in pellet form, the initial striation thickness can be of the order of 100 mm. If a final striation thickness of 1 micron is required, the axial mixing has to achieve a reduction of striation thickness of at least 5 orders of magnitude.

Considering that the axial shear rate in melt conveying is close to zero in the center region of the channel, it is clear that axial mixing will be insufficient unless efficient mixing devices are used. The most efficient way to improve axial mixing is to redistribute material from the center of the channel to the outer region of the channel and vice versa. One mixer that aims to achieve such redistribution is the inside-out mixer. Another mixer that was designed specifically to improve axial mixing is the CRD7 mixer, see figure 8.

Another method of reducing mixing problems with color concentrates is to reduce the initial striation thickness. This can be done by reducing both the virgin plastic pellet size and the color concentrate pellet size. Granules are better than pellets and powder is better than granules from a mixing point of view. Smaller particle sizes may lead to other problems though, such as conveying problems and air entrapment. Adding the colorant in liquid form can also reduce the initial striation thickness. This is one of the main reason liquid colorants are used. However, they can create problems by forming a lubricating layer on the barrel surface and reducing the conveying efficiency of the extruder.

References

1. C. Rauwendaal, "Polymer Mixing, A Self-Study Guide," Hanser Publishers, Munich (1998)
2. C. Rauwendaal, T. A. Osswald, G. Tellez, and P. J. Gramann, "Flow Analysis in Screw Extruders – Effect of Kinematic Conditions," International Polymer Processing, XIII, 4 , 327-333 (1998)
3. Z. Tadmor and C.G. Gogos, "Principles of Polymer Processing," John Wiley and Sons, New York (1979)
4. R.B. Bird, W.E. Stewart, and E.N. Lightfoot, "Transport Phenomena," Wiley, New York (1960)
5. G. Pinto and Z. Tadmor, Polym. Eng. Sci., 10, 279 (1970)
6. J.W. Joo and T.H. Kwon, Polym. Eng. Sci., 33, 15, 959 (1993)

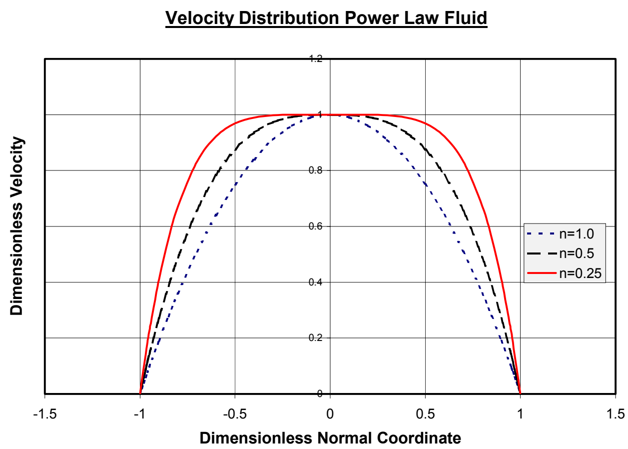

Figure 1, Velocity profiles for various power law index values

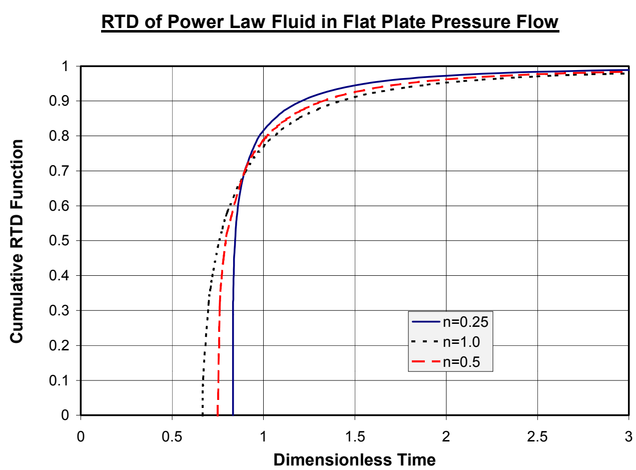

Figure 2, Residence time distribution curves for several power law index values

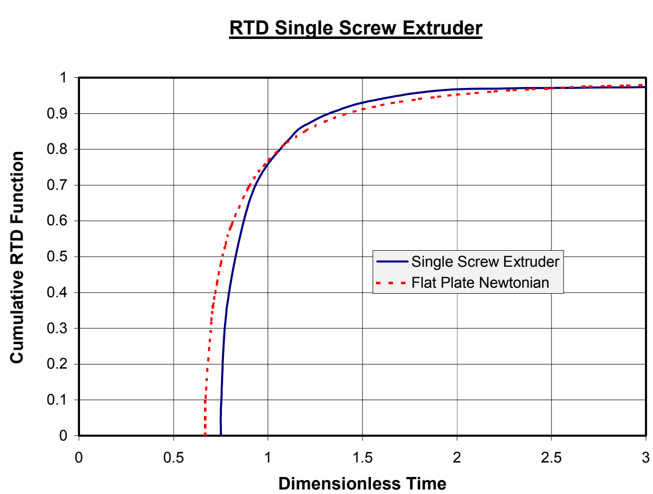

Figure 3, RTD of single screw extruder and flat plate

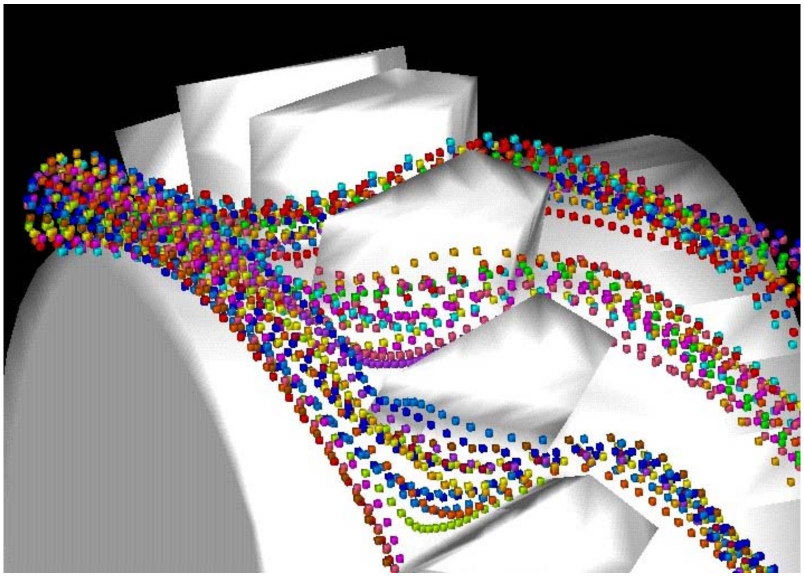

Figure 4, Particle tracking results in a mixing section with elongational mixing pins

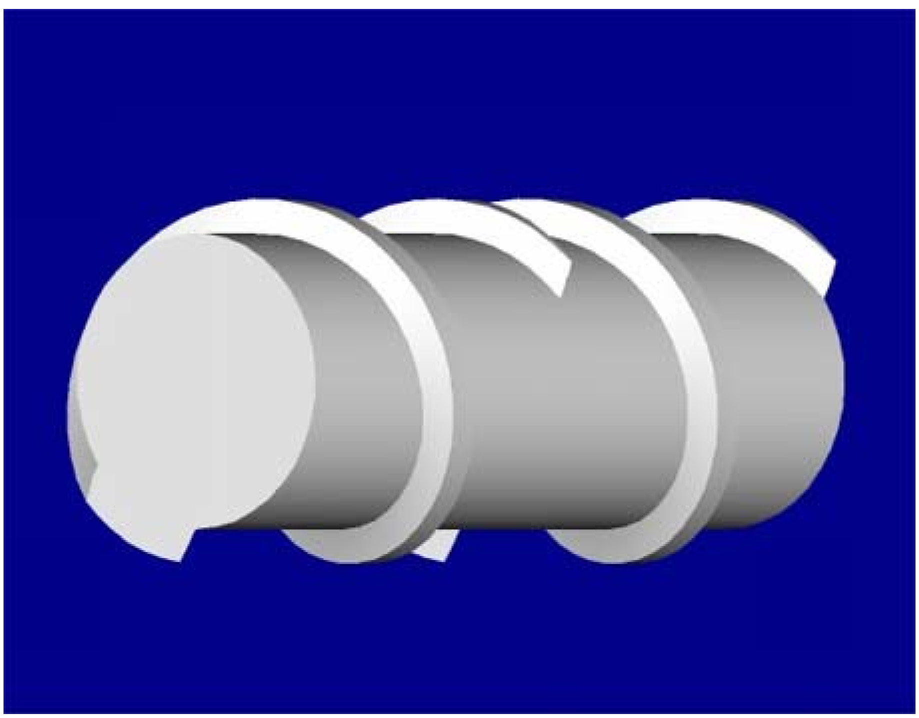

Figure 5, Solid model of Inside-Out mixer

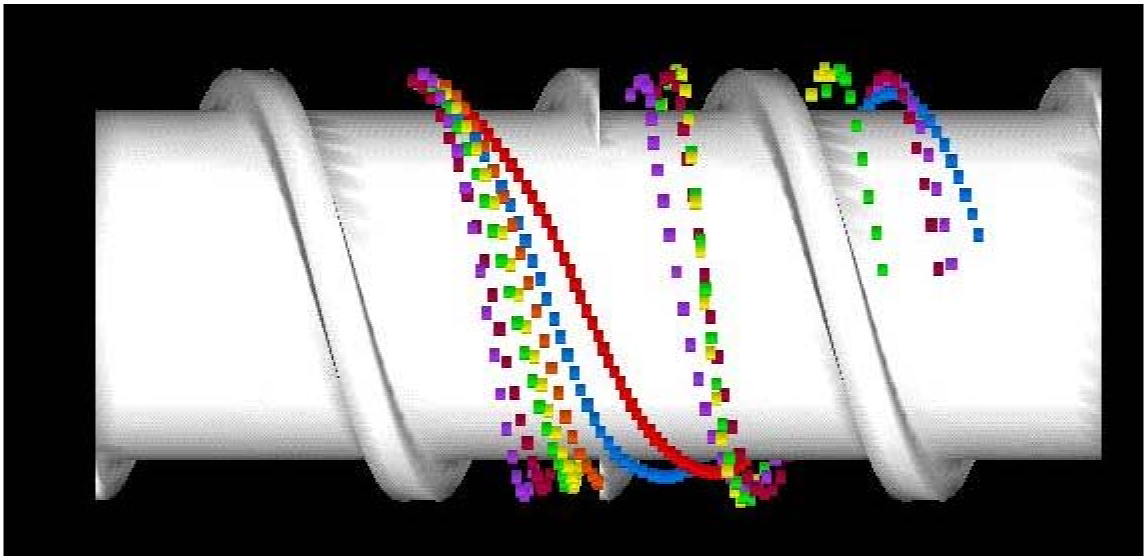

Figure 6, Particle tracking in the Inside-Out mixer

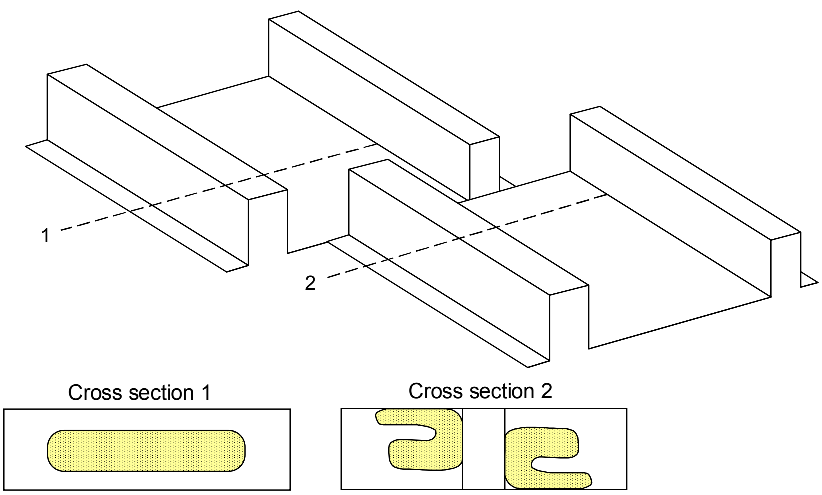

Figure 7, Redistribution in the Inside-Out mixer

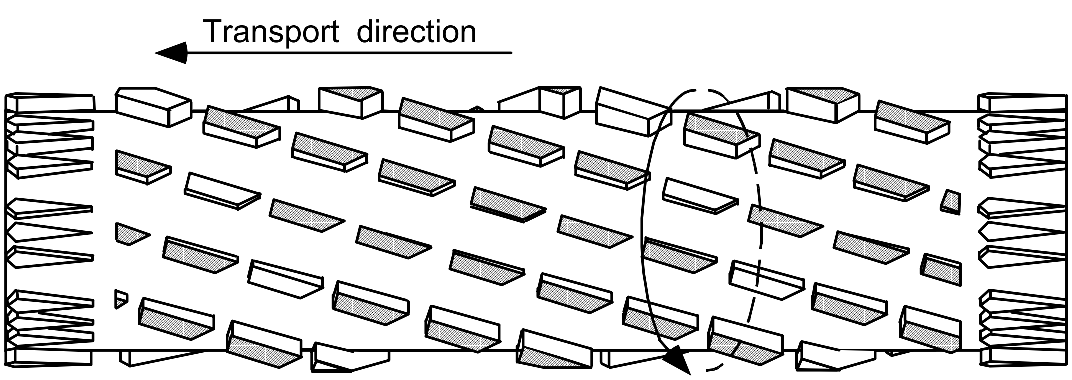

Figure 8, The CRD7 mixer