The effect of stirrer geometry on fibre suspension agitation

Pavel Seichter, Luboš Pešl

TECHMIX s.r.o., Křižíkova 70, 612 00 Brno, CZ, tel.: +420-541633710, fax: +420-549254992,
E-mail: techmix@volny.cz

Key words: agitation, fibre suspension

Side entering agitators with axial stirrers are often used in paper and cellulose production plants. Here, fibre suspension mixing is one of the main problems to be solved. These media are mostly shear thinning fluids and the effect of well mixed cavern around the stirrer can appear. Model tests on laboratory size equipment just as on industrial scale apparatus were done to assess extent of the cavern, i.e. mixing zone. The effect of geometrical arrangement including various axial stirrers was investigated. Experimental results were compared with theoretical concepts for cavern size.

Filamentous paper pulp and cellulose suspensions are often very shear thinning fluids so that their flow curve can be well-fitted by the following equations (Ulbrecht and Mitschka, 1965):

$$\tau = K\, D_r^{\,m} \qquad (1)$$

$$\tau^{0,5} = (\tau_o)_c^{\,0,5} + K_c\, D_r^{\,0,5} \qquad (2)$$

or

$$\tau = \tau_o + \mu_p\, D_r \qquad (3)$$

When yield stress τ_o - value is high enough or flow index $m < 0,3$ mixing of such fluids results in the formation of a region of relatively rapid motion around the impeller (cavern) and essentially stagnant regions elsewhere in the vessel.
Several physical models for assessment of characteristic cavern size D_c have been proposed. A simple concept presented by Wichterle and Wein (1979 and 1981) determines the size of the cavern by the relationship:

$$(D_c / d)^3 = 0,375\, Po^{1/3} \left(\frac{\rho n^2 d^2}{\tau_o}\right)^{3/2} \qquad (4)$$

Other relationship was given by Solomon et all (1981) for a spherical cavern

$$(D_c / d)^3 = \left(\frac{4Po}{\pi^3}\right)\left(\frac{\rho n^2 d^2}{\tau_o}\right) \qquad (5)$$

and more complex approach comprising axial force of the impeller as a decisive factor

$$F_a = N_f\, \rho\, n^2 d^4 \qquad (6)$$

was presented by Amanullah et all (1998)

$$(D_c / d)^3 = \frac{1}{\pi}\left(\frac{\rho n^2 d^2}{\tau_o}\right) \sqrt{N_f^{\,2} + \left(\frac{4Po}{3\pi}\right)^2} \qquad (7)$$

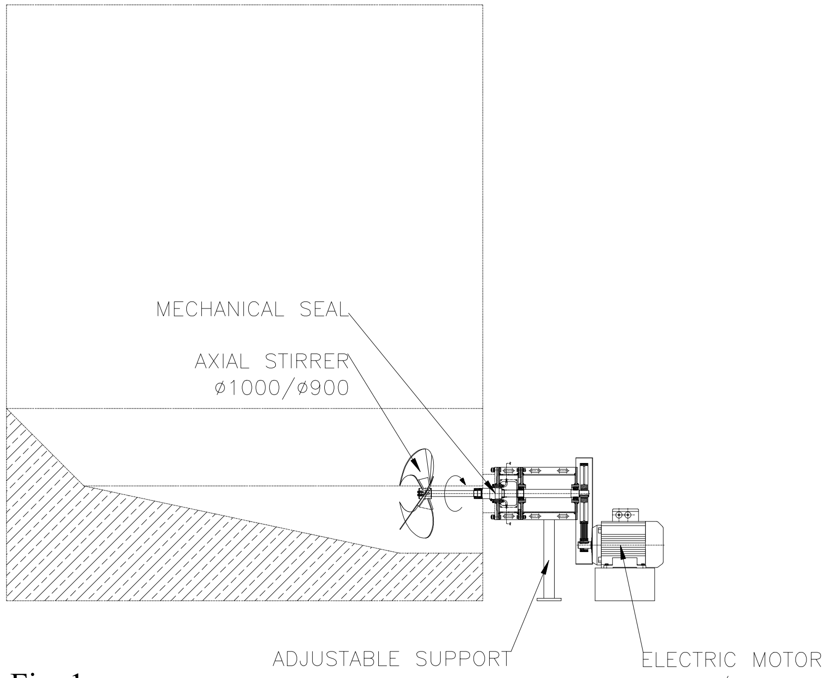


Fig. 1

Mostly spherical shape of the cavern is assumed so that the sphere volume would cover the whole mixed fluid volume.

Industrial agitator arrangement is shown on Fig. 1. Side entering agitator with an axial stirrer is usually situated against sloped bottom to ease suspension circulation in the tank.

Since such lay-out at model stage was rather complicated to install due to shaft sealing problems a little different stirrer location was applied.

Experimental

Model tests were carried out in two vessels: smaller one was of 300 mm diameter with a flat bottom and bigger one was of 400 mm diameter with dished bottom, see Fig. 2 and 3. Position of agitators was eccentric and in the middle of liquid height in all experiments to simulate partly industrial agitators arrangements. Suspension stirrer pumping was always directed to the bottom.

Axial stirrers used are given in the following table.

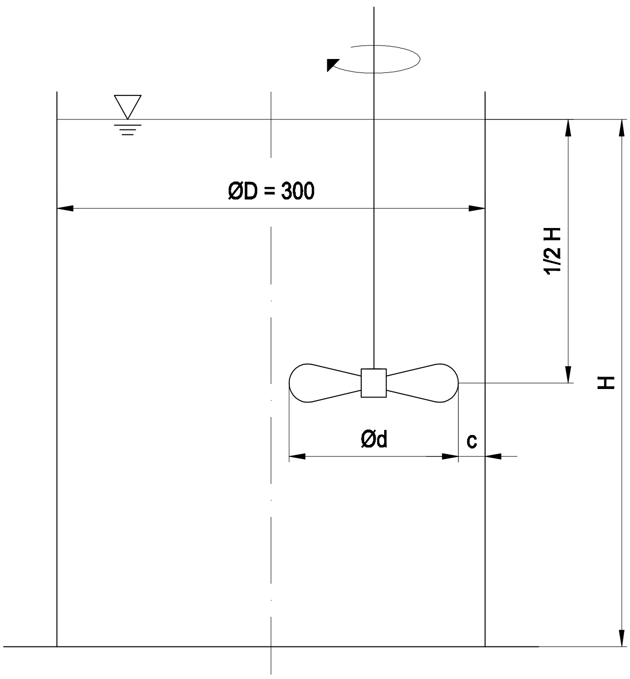


Fig. 2

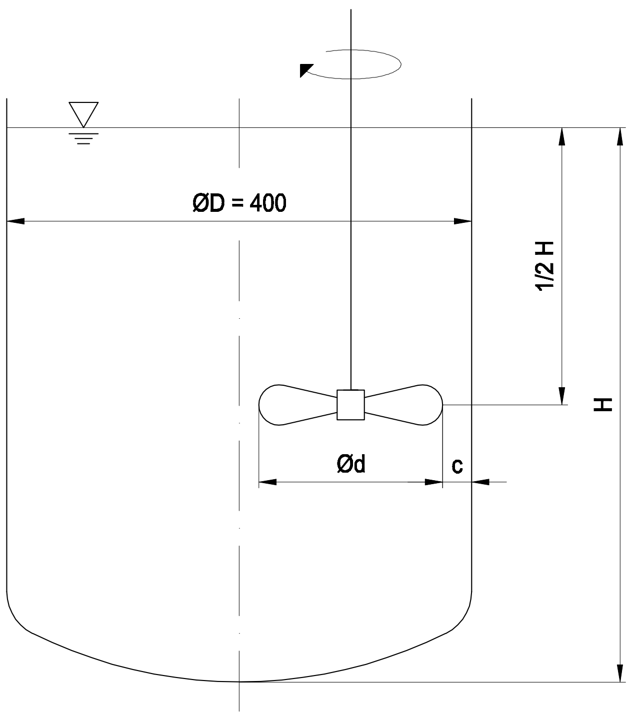


Fig. 3

Tab. I. Axial stirrers used

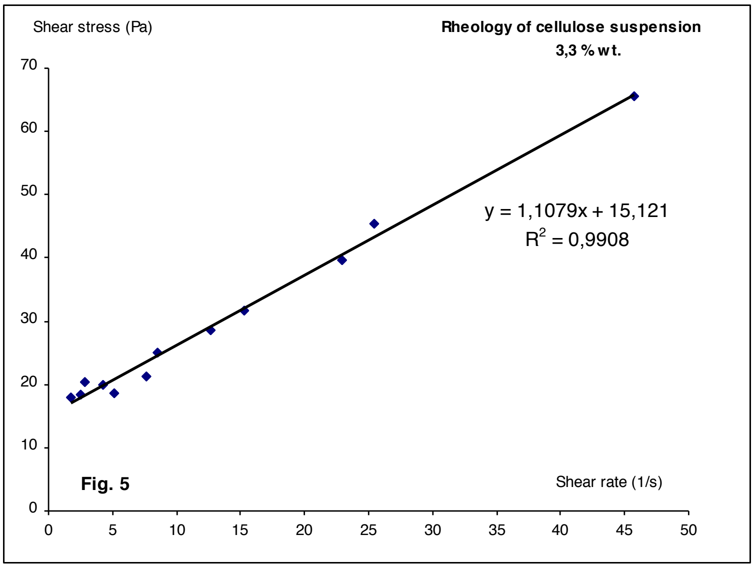
Vessel	Stirrer		
	TX 335	Propeller (ON 19)	"Majak"
ø 300	ø140	ø147	ø145
ø 400	ø182 ø168 ø140		ø145

The stirrers are presented in Fig. 4.

Fig. 4a – TX 335

Fig. 4b – "Majak"

Fig. 4c – Propeller ON 19



Suspension of cellulose 3,3 per cent wt. (Biocel) was applied for model tests. Rheological properties of the suspension are shown on Fig. 5.
These data were obtained from rheoviscometer Rheotest 2 equipped and calibrated with a radial impeller placed at the bottom of the measuring cylindrical vessel 0,8 litre of volume.
The suspension rheology corresponds well to equation (3) i.e. Bingham plastic with τ_o = 15,1 Pa and μ_p = 1,1 Pas.
Quality of agitation was assessed by measuring of passage time of a floating particle drifted on the suspension level (on the distance 150 mm) during mixing process.

Results of model tests

Fig. 6: Data and curves here show dependencies of necessary specific power input P/V to achieve required passage time of the "flow follower" particle across 15 cm distance. The slower movement of the particle the less energy for agitation is needed.
Considerably higher power input is recorded for axial stirrer "Majak" (stirrer frequently used in Russian cellulose plants). The difference between two other stirrers is not very important.

Fig. 7: Presents the effect of the clearance C between the wall and stirrer tip. It is not unambiguous - for faster particle passage i.e. more intensive agitation smaller gap (C) seems to be better.

Fig. 8: Shows the effect of stirrer blade angle at the hub for hydrofoil stirrer TX 335.

Smaller angle 30° of its blade is accompanied with more energy consumption to reach higher suspension pumping rate.

Fig. 9: The effect of stirrer size is evaluated here. It can be stated that the bigger stirrer the less power input is necessary to dissipate for a given suspension flow rate. Again worst result presents the axial stirrer "Majak" as compared with axial hydrofoil ones TX 335, despite its greater diameter (145 mm) than the smallest TX 335 (141 mm).

Fig. 10: Verification of simplex D/d effect on necessary specific power input for stirrers pumping activity is presented here - results in bigger vessel D = 400 mm with dished bottom.

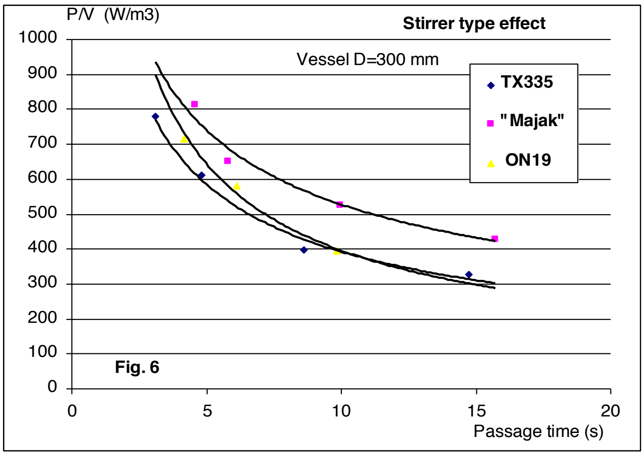


Fig. 6

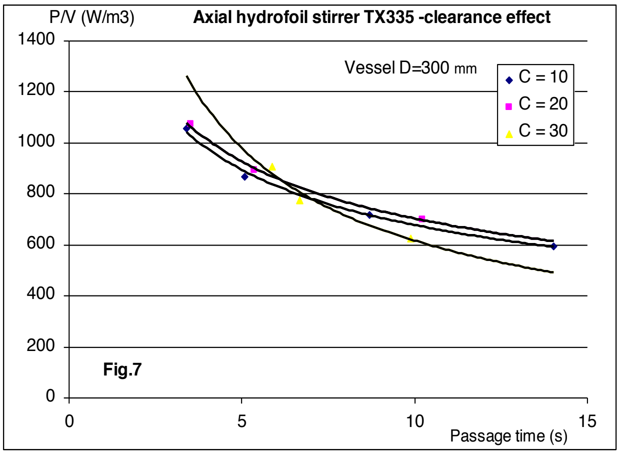


Fig.7

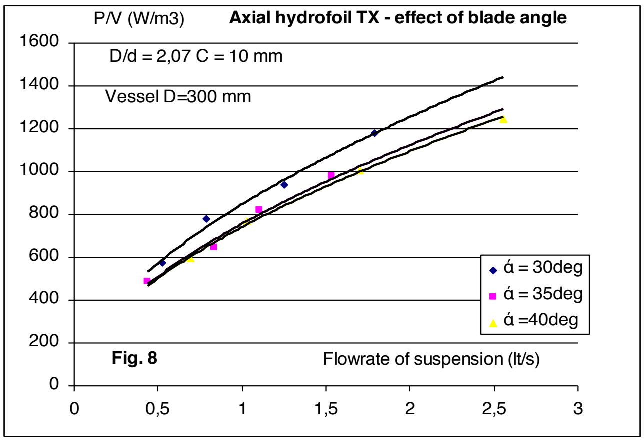


Fig. 8

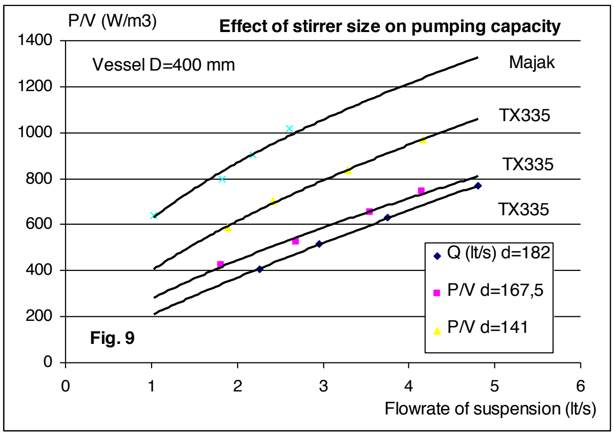


Fig. 9

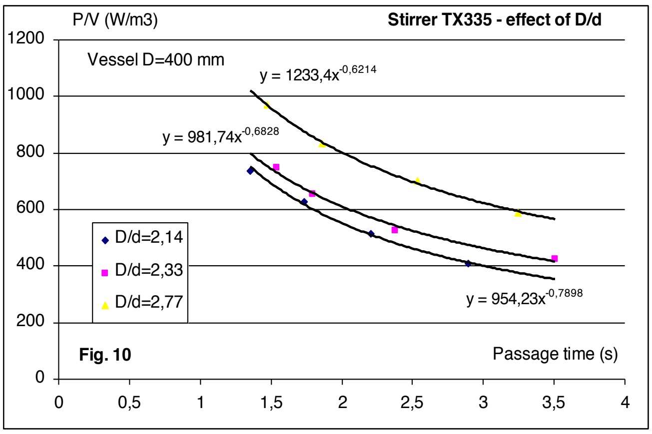


Fig. 10

Large scale test

Tests on industrial size vessel 30 m^3 (see Fig. 1 and 11), where fibre suspension of disintegrated waste paper was agitated by axial hydrofoil stirrer TX 335, were carried out at Cellulose plant Dabrowica, Poland. Rheology of the mixed medium sample was tested on the same instrument as at laboratory experiments.

Rheological behaviour of this suspension can be expressed by several models:

Bingham model:

$$\tau = \tau_o + 1{,}366\, D_r \qquad (8)$$

with: $\tau_o = 18{,}71$ Pa
$R^2 = 0{,}851$

Casson model:

$$\tau^{0,5} = \tau_o^{0,5} + 0{,}832\ D_r^{0,5} \quad (9)$$

with: $\tau_o = 10{,}59$ Pa
$R^2 = 0{,}907$

Herschel - Bulkey model:

$$\tau = \tau_o + 2{,}51\ D_r^{0,905} \quad (10)$$

with: $\tau_o = 12$ Pa
$R^2 = 0{,}943$

Fig. 11 – V

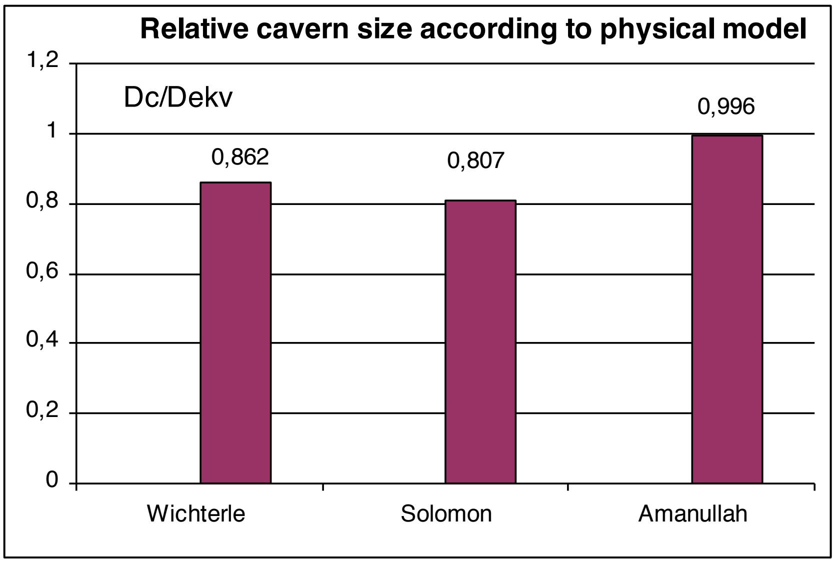
The last presented relationship seems to express flow properties of the suspension best.
Equipment parameters:

Vessel diameter:	3,2 m
Batch height:	3,67 m
Batch volume:	31 m^3
Stirrer diameter:	900 mm
Stirrer speed:	144 rpm

Motor power input: 15 kW
Impeller power consumption: 10,3 kW
Agitation intensity: Appreciated as fully convenient with very intensive the whole batch circulation

Parameters of this agitator were used to verify the calculations according to the presented physical models for the cavern size (Eq. 4, 5 and 7). As a characteristic dimension of the mixed suspension batch the sphere diameter D_{ekv} corresponding to the total batch volume was applied.

Fig. 12: presents the results of these calculations.



It is obvious that Amanullah (1998) model gives here the best fit of cavern diameter D_c to equivalent large scale batch D_{ekv} for the tested industrial 30 m^3 case. Other two models gave quite plausible approximations considering a little vague assessment of the intensity of suspension batch mixing.

Conclusion

The results of small scale experiments showed several effects of agitator geometry. Particularly, bigger size of the stirrer exhibited less energy consumption for the achievement of a required intensity of agitation. The influences of stirrer gap at the vessel wall and axial blades angle are not very pronounced. On the other hand, large solidity stirrer TX 335 seems to be a good choice for agitation of non-Newtonian fibre suspension fluids.
Physical model by Amanullah et all (1998) appears to be a good tool for the assessment of the requested intensity agitation on industrial size apparatus with shear thinning fluids.

Notation

C stirrer clearance to vessel wall, m
D vessel diameter, m
D_c cavern diameter, m
D_r shear rate, s^{-1}
d stirrer diameter, m
F_a axial force, N
H suspension filling height, m
K fluid consistency coefficient, $Nm^{-2}\ s^m$
m flow behaviour index
N_f axial force number (dimensionless: $N_f = F_a / \rho\ n^2\ d^4$)
n stirrer speed, s^{-1}
P stirrer power input, W
Po power number (dimensionless: $Po = P / \rho\ n^3\ d^5$)
V suspension batch volume, m^3

Greek symbols

μ fluid viscosity, Pas

ρ fluid density, kg m^{-3}

τ shear stress, N m^{-2}

Subscripts

ekv equivalent

o yield stress

p plastic state

References

Amanullah A., Hjorth S.A. and Nienow A.W. (1998) A new mathematical model to predict cavern diameters in highly shear thinning, power law liquids using axial flow impellers. *Chem. Eng. Sci.* 53, 455 - 469.

Solomon J., Elson T.P., Nienow A.W. and Pace G.W. (1981) Cavern sizes in agitated fluids with a yield stress. *Chem. Engng. Commun.* 11, 143 - 164.

Wichterle K. and Wein O. (1981) Threshold of mixing of non-Newtonian liquids. *Int. Chem. Engng.* 21, 116 - 121.

Wichterle K. and Wein O. (1979) Práh míchání nenewtonských kapalin. *Chem. Průmysl* (in Czech) 29, 113 - 116.

Ulbrecht J. and Mitschka P. (1965) Chemical engineering of non-Newtonian liquids. NČSAV Praha.

This research has been subsidized by the Research Project of Ministry of Industry and Trade of the Czech Republic FR-TI1/005.