Solid-liquid separation of lignocellulosic sugars from biomass by rotating ceramic disc filtration

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Table S1: Properties and analysis of DDR-EH slurry and MF permeate for manuscript section 'SLS of hydrolysate using RCD filtration'.

	DDR-EH	Permeate	Unit
% wt. total solids	14.91	11.91	wt.%
% wt. Insoluble solids	3.02	0	wt.%
Density	1.04	1.04	g/mL
pH	3.0	3.0	
Lignin	0.21	0.22	mg/mL
Cellobiose	2.4	2.4	mg/mL
Glucose	73.2	73.2	mg/mL
Xylose	34.4	34.4	mg/mL
Galactose	1.5	1.5	mg/mL
Arabinose	4.2	4.2	mg/mL
Fructose	0.0	0.0	mg/mL
Lactic acid	1.4	1.4	mg/mL
Glycerol	0.32	0.3	mg/mL
Acetic Acid	2.26	2.3	mg/mL
Ethanol	1.23	1.2	mg/mL
HMF	0.00	0.0	mg/mL
Furfural	0.00	0.0	mg/mL

Table S2: Two-stage filtration analysis

*466 mL of the stage 1 retentate and 600 mL of water

	DDR-EH Feed	Stage 1 retentate	Stage 1 permeate	Stage 2 feed	Stage 2 retentate	Stage 2 permeate	Unit
Volume	1416	466	950	1066*	450	616	mL
% wt. total solids	14.55	19.52	12.11	8.23	14.22	4.8	wt.%
% wt. Insoluble solids	2.71	8.24	0	3.60	8.53	0	wt.%
Particle size (volume basis)	36.7	35.6	n/a	37.5	31.8	n/a	µm
Density	1.04	1.04	1.05	1.01	1.02	1.03	g/mL
pH	3.20	3.24	3.26	3.40	3.34	3.37	
Cellobiose	0.35	0.37	0.47	0.14	0.17	0.22	mg/mL
Glucose	74.9	74.6	74.6	30.2	32.9	30.0	mg/mL
Xylose	30.9	32.2	30.6	12.6	16.7	16.9	mg/mL
Galactose	1.6	2.2	1.3	1.2	1.4	1.6	mg/mL
Arabinose	4.2	4.6	3.7	2.2	2.7	2.6	mg/mL
Fructose	0	1.2	0	1.1	1.1	1.6	mg/mL
Lactic acid	0.4	1.2	1.2	0.4	0.5	0	mg/mL
Glycerol	0	0	0	0	0	0	mg/mL
Acetic Acid	0.12	0.37	0.37	0.14	0.19	0	mg/mL
Ethanol	0	0	0	0	0	0	mg/mL
HMF	0	0	0	0	0	0	mg/mL
Furfural	0	0	0	0	0	0	mg/mL
Lignin	54.5	54.5	n/a	51.9	52.1	n/a	wt. %
Glucan	20.3	20.5	n/a	20.1	20.5	n/a	wt. %
Xylan	8.5	9.6	n/a	9.2	10.6	n/a	wt. %
Galactan	1.2	1.3	n/a	1.4	1.1	n/a	wt. %
Arabinan	3.1	5.4	n/a	3.0	3.7	n/a	wt. %

Table S3: Additional TEA assumptions

	Cost ($)	**Unit**
Membrane cost[1]	2000	$/m2
Membrane lifetime[1]	5	years
Electricity cost[1]	0.0572	$/kWh
Steam cost[2]	0.018	$/kg
On-stream factor	0.9	
Flocculant cost[3]	2.5	$/kg
Process water[4]	4	kGal
Cleaning cost[5]	20% of OPEX	

1. Saboe, Patrick O., et al. "Recovery of low molecular weight compounds from alkaline pretreatment liquor via membrane separations." *Green Chemistry* 24.8 (2022): 3152-3166.
2. Salvachua, Davinia, et al. "Process intensification for the biological production of the fuel precursor butyric acid from biomass." *Cell Reports Physical Science* 2.10 (2021).
3. https://www.nrel.gov/extranet/biorefinery/aspen-models/
4. https://www.energy.gov/sites/prod/files/2017/10/f38/water_wastewater_escalation_rate_study.pdf
5. Gruskevica, Kamila, and Linda Mezule. "Cleaning methods for ceramic ultrafiltration membranes affected by organic fouling." *Membranes* 11.2 (2021): 131.

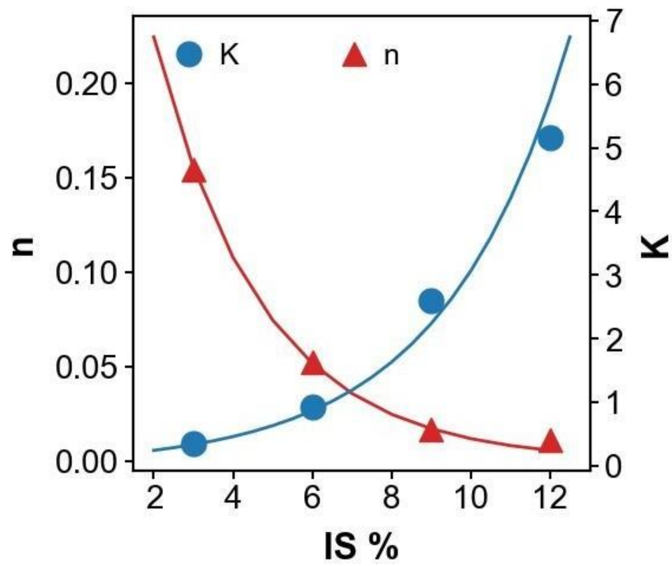

Figure S1: K and *n* parameters fit to equations S1 and S2

$$K = 0.131\exp(0.32 * IS\%) \qquad \text{(Equation S1)}$$

$$n = 0.468\exp(-0.37 * IS\%) \qquad \text{(Equation S2)}$$

The cake formation model

$$J = J_o(1 + k_c t)^{-0.5} \qquad \text{(Equation S3)}$$

J is the flux, J_o is the initial flux, k_c is the cake filtration constant ($s \cdot m^{-2}$) and t is time (s).

Flux as a function of insoluble solids concentration (Fit to Figure 3A)

$$LMH = -0.43\ \%IS^2 - 0.44\ \%IS + 132 \qquad \text{(Equation S4)}$$

LMH is the flux, %IS is the insoluble solids concentration (wt%)

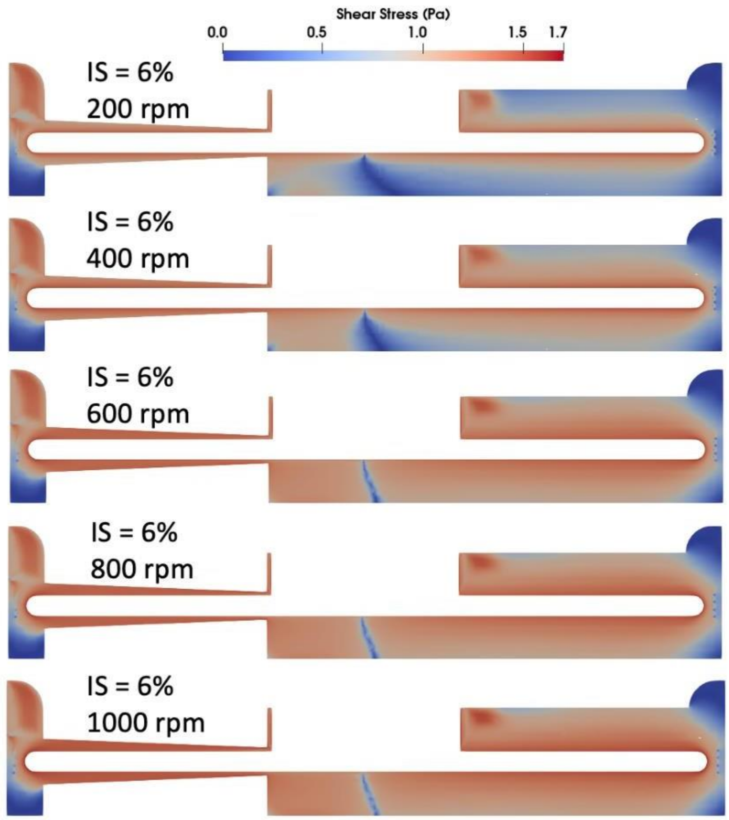

Figure S2. Impact of rotation speed on shear stress distribution.

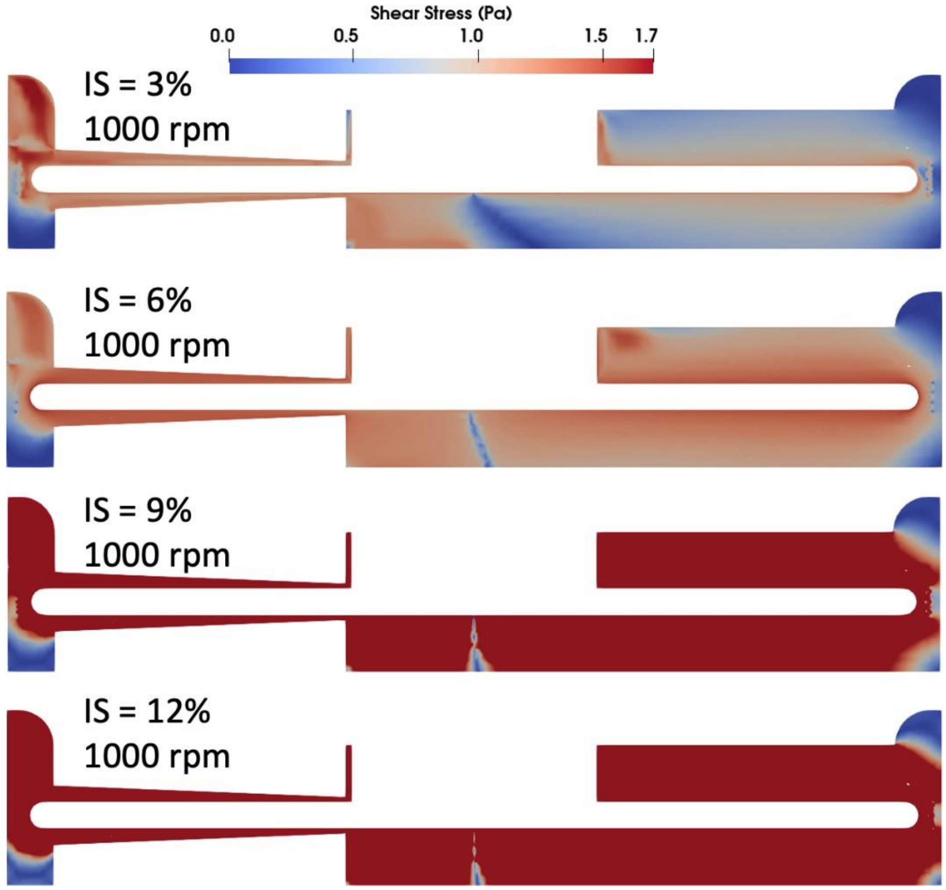

Figure S3. Shear stress distribution of the hydrolysate with different insoluble solids and with rotational speed of 1,000 rpm.

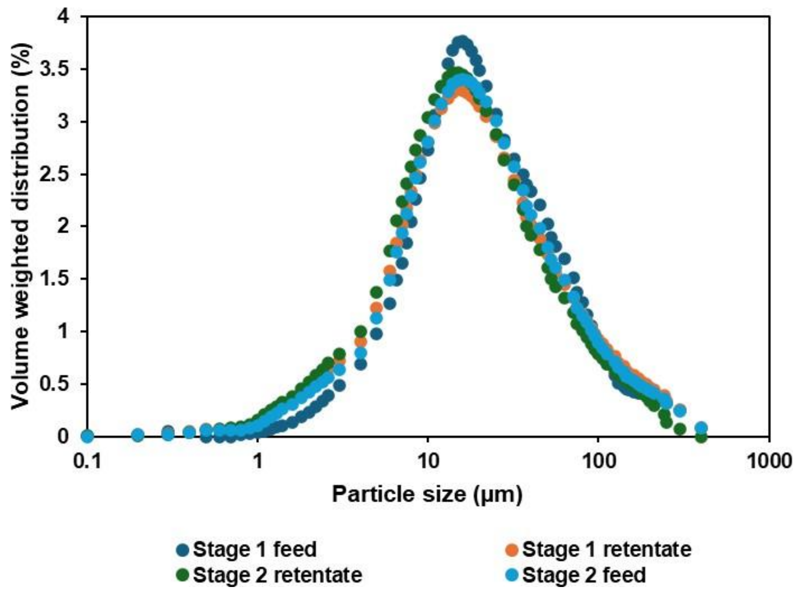

Figure S4. Particle size analysis showing the volume weighted distribution.

Cross-flow filtration energy consumption analysis

Shear-rate

From the Rabinowitsch-Mooney relationship[1] (Equation S5), the cross-flow velocity and the *n* parameter from the non-Newtonian shear thinning power-law viscosity model (Figure S1) are used calculated the wall shear rate.

$$v = \frac{\dot{\gamma}\, D}{8}\left(\frac{4n}{1+3n}\right) \qquad \text{(Equation S5)}$$

In Equation S5, v is the channel cross-flow velocity $(m \cdot s^{-1})$, $\dot{\gamma}$ is the shear rate at the membrane surface (s^{-1}), D is the channel inner diameter (m) and n is the power law parameter.

With the known cross-flow velocity, the volumetric flowrate in a single channel can be calculated:

$$Q = \pi v \left(\frac{D}{2}\right)^2 \qquad \text{(Equation S6)}$$

In Equation S6, Q is the volumetric flow rate in a single channel $(m^3 \cdot s^{-1})$, π is the mathematical constant pi, v is the channel cross-flow velocity $(m \cdot s^{-1})$, and D is the channel inner diameter (m).

Subsequently, the total volumetric flowrate through the filtration stage can be calculated:

$$Q_{Tot} = N_c N_{mod} Q \qquad \text{(Equation S6)}$$

In Equation S6, Q_{Tot} is the total volumetric flowrate through the filtration stage and Q is the volumetric flow rate in a single channel $(m^3 \cdot s^{-1})$, N_c is the number of channels in a filtration module, and N_{mod} is the number of modules required for the filtration step.

Fanning Friction Factor

To calculate this pressure drop in the membrane module, we must first solve for the modified Reynolds number for non-Newtonian fluids defined by Metzner and Reed:

$$K' = K\left[\frac{1+3n}{4n}\right]^n \qquad \text{(Equation S7)}$$

$$Re_{MR} = \frac{D^{n'} V^{2-n'} \rho}{K' 8^{n'-1}} \qquad \text{(Equation S8)}$$

In Equations S7 and S8, K' is the modified power law coefficient and K is the power law coefficient, n is the power law exponent, Re_{MR} is the modified Reynolds number, D is the channel inner diameter (m), V is the channel cross-flow velocity $(m \cdot s^{-1})$, ρ is the feed-side fluid density $(kg \cdot m^{-3})$, and $n' = n$.

Now, the fanning friction factor can be calculated:[7]

$$f^{-0.5} = 4log(Re_{MR} f^{0.5}) - 0.4 \qquad \text{(Equation S9)}$$

In Equation S9, f is the fanning friction factor and Re_{MR} is the modified Reynolds number.

Pump Power Consumption

[1] Tilton, J., Perry's Chemical Engineers' Handbook Section 6: Fluid and Particle Dynamics. *McGraw-Hill Professional* **2007,** *4*, 12-13.

The pressure drop through the module is calculated:

$$\Delta P = \frac{2\rho f V^2 L}{D} \qquad \text{(Equation S10)}$$

In Equation S10, ΔP is the pressure drop through the module (Pa), ρ is the feed-side fluid density $(kg \cdot m^{-3})$, V is the channel cross-flow velocity $(m \cdot s^{-1})$, L is the module length (m), and D is the channel inner diameter (m).

Finally, the pump power consumption can be determined:

$$P_n = \frac{Q_{Tot} \Delta P}{\xi} \qquad \text{(Equation S11)}$$

$$P = \frac{P_n}{Q_{perm}} \qquad \text{(Equation S12)}$$

In Equations S11 and S12, P_n is the net power requirement of the pump (W), Q_{Tot} is the total volumetric flowrate through the filtration stage $(m^3 \cdot s^{-1})$, ΔP is the pressure drop through the module (Pa), ξ is the pump efficiency, set to 0.6, P is the power requirement per volume permeate produced $(kWh \cdot m^{-3})$, and Q_{perm} is the permeate flow rate $(m^3 \cdot h^{-1})$.

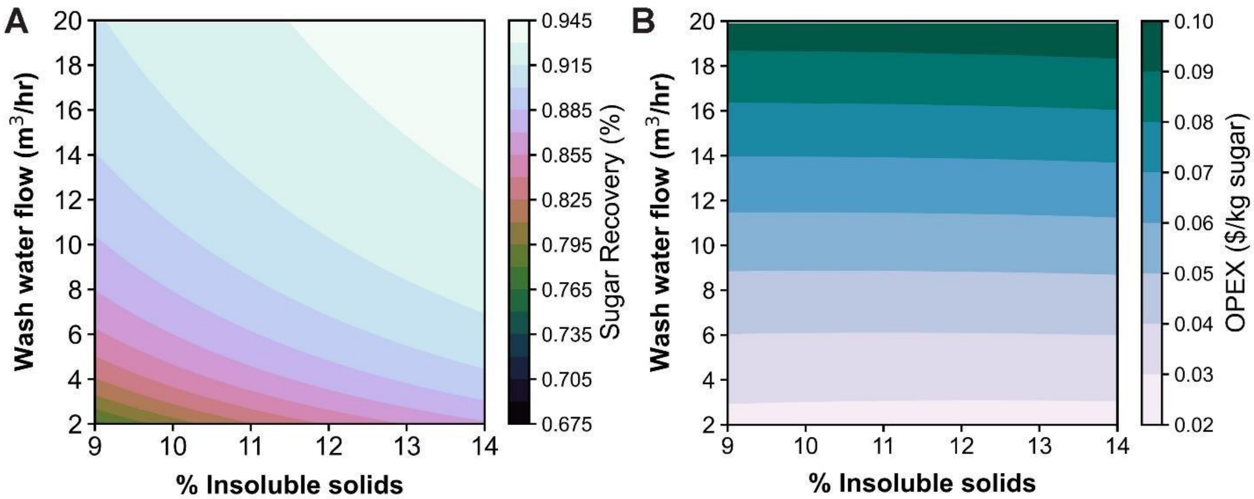

Figure S5. Process model results – flocculate case.

Multi-effect Evaporator with Integrated Mechanical Vapor Recompression

In this study, a multi-effect evaporator (MEE) with integrated mechanical vapor recompression (MVR), modeled in Aspen Plus, was adopted to concentrate sugars in the permeate streams to achieve a final concentration of 500 g/L (50 wt%) of sugars. The MEE-MVR system combines the benefits of multiple-effect evaporation and mechanical vapor recompression to achieve energy efficiency and cost savings.

The MVR-assisted evaporation process consists of the following steps: preheating, evaporation, concentration, and energy recovery. The process begins with a feed dilute sugar stream containing mixed sugars (Figure S5). The feed stream enters the first effect of the multi-effect evaporator. Each subsequent effect operates at a lower pressure and temperature than the previous one. The secondary vapor from each effect serves as the heating medium for the next effect, thus reusing latent heat and reducing steam consumption. The final effect produces a concentrated syrup containing mixed sugars. MVR is a key component in this process. It utilizes energy recovered from the condensate to recompress vapor. The vapor from the last effect is compressed using a mechanical compressor. The compressed vapor is then used as the heating medium for the first effect. By reusing the latent heat, MVR significantly reduces the need for external steam.

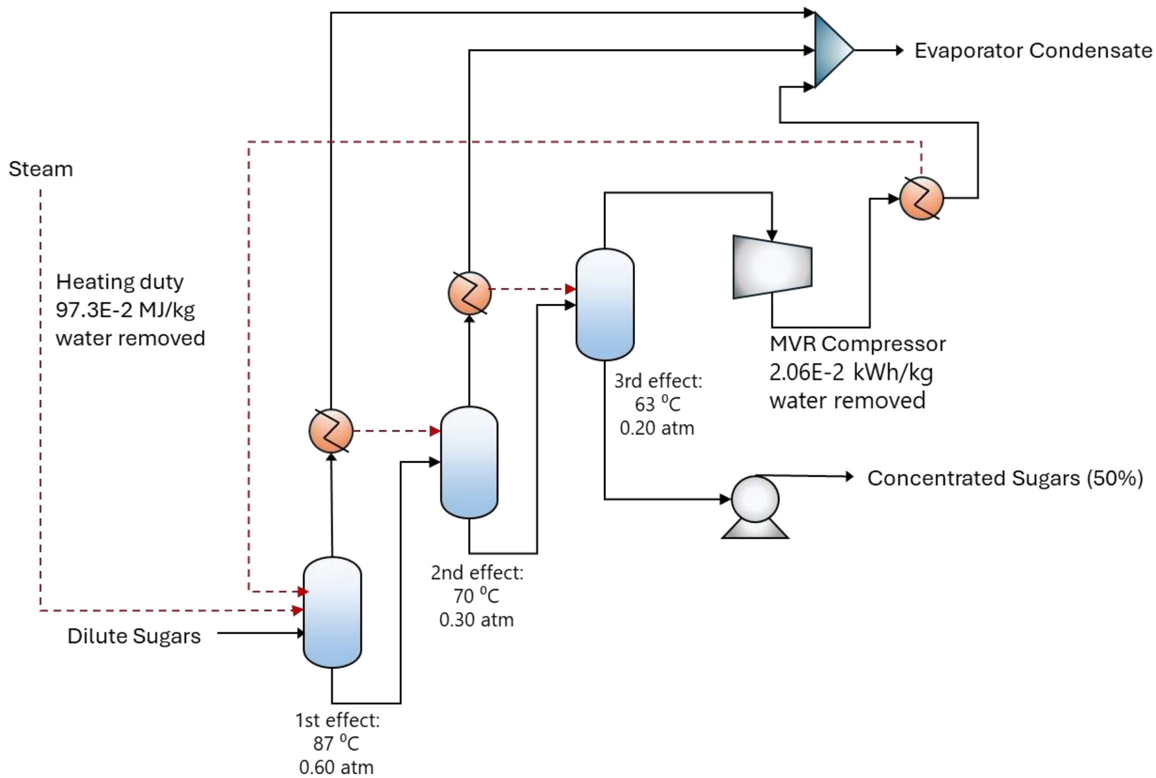

Figure S6. Sugars concentration via multi-effect evaporator integrated with mechanical vapor recompression.

Process Model Operating Guide

A two-stage filtration process model including downstream evaporation is available on GitHub (https://github.com/NREL-SEPCON). This file calculates the sugar, water, and energy balance across the system to calculate the OPEX of the system on a per kg of sugar recovered basis.

The following system specifications are given for the process. To solve the sugar and water mass balance, required input parameters include (1) the flowrate of feed, (2) the flowrate of wash water, (3) the insoluble solids content in the retentate, (4) the insoluble solids content in the feed, and (5) the sugar concentration in the feed, (6) the sugar concentration in the wash water.

To generate data for the main text, we varied the flowrate of the wash water from 0 to 2-fold the feed flow rate. The insoluble solids content in the retentate was varied between 6 and 15 wt%. The insoluble solids content in the feed was set to 3%. The sugar concentration was set at 100 g/L in the feed and 0 g/L in the wash water. The evaporation operation is defined so that the incoming permeate stream is concentrated to 500 g/L of sugars leaving the evaporator.

The process model code is specific for a rotating disc module operating at 1200 rpm with DDR-EH as the feed. However, the general mass balance equations are the same for other feed slurries. To use the code for another feed slurry, information on the energy consumption of the MF process is needed which is dependent on the viscosity and IS content of the slurry at the operating condition. Lower disc speeds can be applied in the model to reduce energy consumption. However, the impact of lower disc speeds on the permeability is currently unknown over a long duration. The membrane area required for the process was estimated based on the average flux measurements.