

Supplementary Information for

Efficient conversion of syngas into sustainable aviation fuel

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Supplementary Methods:

Synthesis of the shaped catalyst

To meet the demands of industrial applications, the powdered CoMn/ $\beta_0.2$ catalyst was shaped into cylindrical pellets via extrusion molding. The formulation was as follows: the catalyst powder was mixed with pseudo-boehmite (PB, 20 wt%) as a binder to provide both mechanical strength and thermal conductivity, carbon powder (2 wt%) as a pore-forming agent to introduce additional mesoporosity and improve mass transfer. A dilute nitric acid solution (1 wt%) was added as a peptizing agent to the pseudo-boehmite and form a cohesive and extrudable paste. The mixture was thoroughly kneaded, extruded through a cylindrical die, and the resulting extrudates were dried at 120 °C for 12 h, followed by calcination at 350 °C for 4 h to remove the organic additives and stabilize the shaped catalyst bodies.

Catalyst tests

The FT was conducted in a fixed-bed steel reactor. Before the catalytic tests, the catalyst was ground and sieved to a particle size of 60–80 mesh. Subsequently, 1.0 g of the catalyst was diluted with silica sand at a volume ratio of 1:2 (catalyst: silica sand). Prior to the reaction, the catalyst was pretreated in pure H₂ gas (50 mL min⁻¹) at 400 °C for 8 h. Afterward, the temperature was lowered to 200 °C, and syngas was introduced under the specified reaction conditions for catalytic testing.

The FT performance of the pellet catalyst (10.0 g, size in cylindrical shape of $\Phi 2.0 \times 5.0$ mm) was investigated in a pilot-scale fixed-bed reactor (internal diameter, 19 mm; length, 1180 mm), which was equipped for operation at industrial working conditions with recycled tail gas (**Supplementary Fig. 47**). The feed compressor raised syngas pressure and offset bed pressure drop, while a downstream back-pressure regulator maintained constant system pressure. A separate recycle compressor repressurized unreacted tail gas after liquid product separation and mixed it with fresh feed for cyclic reaction. To ensure efficient heat removal from the highly exothermic FT reaction, the reactor was operated in a jacketed tubular configuration using circulating silicone oil as the heat-transfer medium,

while the catalyst bed was diluted with SiC (catalyst/SiC = 1:2, v/v) to enhance heat transfer and suppress local hot-spot formation.

The oil and wax products generated during the FT reaction were collected using a cold trap (0 °C) and a hot trap (120 °C), and analyzed using a flame ionization detector (FID) equipped on a Shimadzu GC-2010. The tail gas, consisting of H₂, CO, N₂, CO₂, and gas-phase hydrocarbons (C₁₋₇), was analyzed using a thermal conductivity detector (TCD) and an FID, also on a Shimadzu GC-2010. CO conversion was determined by the internal standard method through the 1.5% N₂ from the feed gas.

CO conversion (Conv. CO) was calculated from equation (1):

$$Conv_{CO} = \frac{CO_{in} - CO_{out}}{CO_{in}} \times 100\% \quad (1)$$

Where CO_{in} and CO_{out} represent moles of CO at the inlet and outlet, respectively.

The selectivity of an individual hydrocarbon, Sel_{C_nH_m}, among the hydrocarbons (free of CO₂) in equation (2) was calculated according to

$$Sel_{C_nH_m} = \frac{C_{nH_m out} \times n}{\sum_{i=1}^n C_{nH_m out} \times n} \times 100\% \quad (2)$$

CO₂ selectivity (Sel_{CO₂}) was calculated from equation (3):

$$Sel_{CO_2} = \frac{CO_{2out}}{CO_{in} - CO_{out}} \times 100\% \quad (3)$$

The carbon balance was calculated from equation (4):

$$C_{balance} = \frac{CO_{2out} + C_{nH_m} \times n}{CO_{in} - CO_{out}} \times 100\% \quad (4)$$

Where CO_{2out} denotes moles of CO₂ at the outlet. The carbon balance of all tests was between 95% and 105%.

Catalyst characterizations

The chemical composition of the zeolites was determined with the use of an X-ray fluorescence (XRF) spectroscopy (Rigaku ZSX Primus II, Japan) by SQX calculation. The Cobalt loading of CoMn-

based catalysts was analyzed using inductively coupled plasma-optical emission spectroscopy (ICP-OES) on a Perkin Elmer Optima 8000.

X-ray powder diffraction (XRD) measurements were performed on a Rigaku Ultima IV diffractometer. The Cu K α radiation ($\lambda = 0.15406$ nm) generated at 40 kV and 30 mA was used as the X-ray source. Brunauer–Emmett–Teller (BET) surface area and Barrett–Joyner–Halenda (BJH) pore volume of the catalysts were determined through N₂ adsorption-desorption experiments with a Micromeritics 2020 analyser.

The Fourier transform infrared spectra (FT-IR) spectrometer was obtained using a Nicolet 6700 spectrometer. Samples were mixed and with KBr (in a mass ratio of 1:10) for measurement in wavenumber range of 400–4000 cm⁻¹. Pyridine-adsorption infrared (Py-IR) spectroscopy measurements were conducted on a Thermo Scientific Nicolet 380 spectrometer equipped with an MCT detector. The sample, loaded in a thermostated cell with CaF₂ windows, was first evacuated at 450 °C to 10⁻² Pa for 2 h, exposed to pyridine vapor for 1 h to achieve saturation, and then subjected to Py-IR analysis to record the desorption spectra.

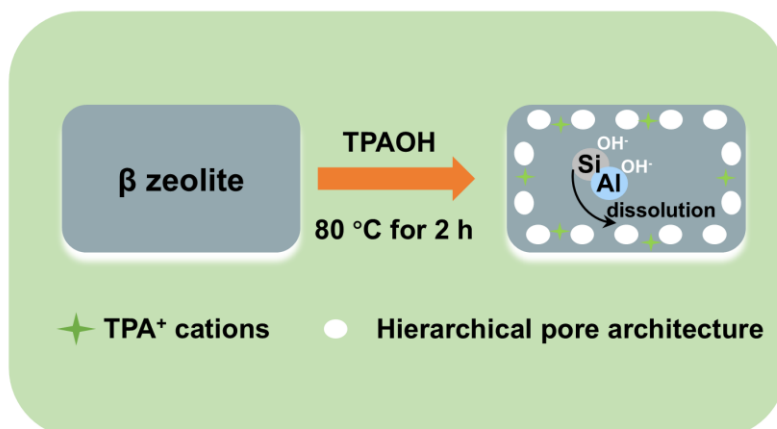
X-ray photoelectron spectroscopy (XPS) was collected on a Thermo Scientific K-Alpha with Al K α at 150 W ($h\nu = 1486.6$ eV), calibrated internally by the carbon deposit C1s ($E_b = 284.8$ eV).

Thermogravimetric (TG) analysis of the spent samples was conducted on a STA449-QMS thermal analyzer, and the carbon species were examined using a Renishaw inVia confocal Raman microscope equipped with a $\times 50$ objective lens.

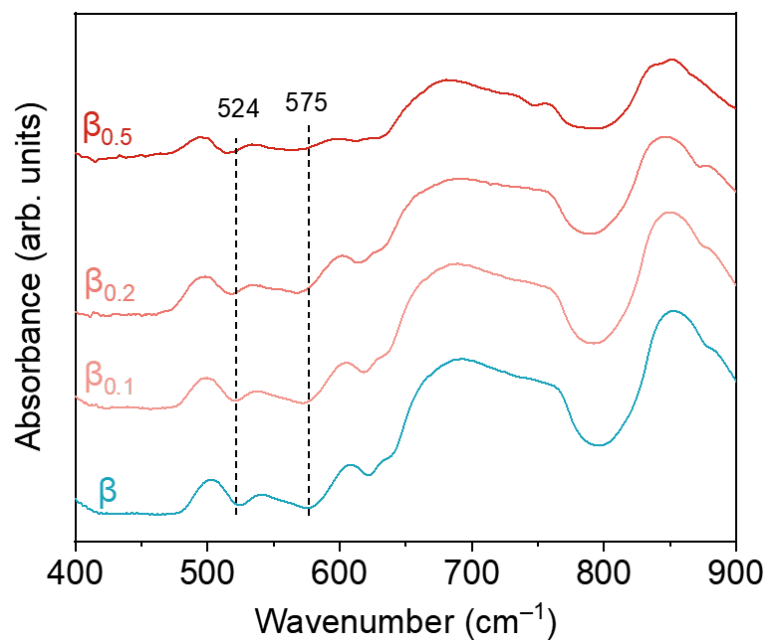
H₂ temperature-programmed reduction (H₂-TPR) was performed on a Micromeritics Autochem 2920, with tail gas detected by thermal conductivity detector (TCD). For each test, 50 mg catalyst was pretreated in Ar at 150 °C for 2h, followed by heating from 50 to 750 °C at 10 °C/min in a 5% H₂/Ar mixture. The NH₃ temperature-programmed desorption (NH₃-TPD) experiments were also conducted using a TCD. Typically, 100 mg of sample was pretreated in an Ar flow (30 mL min⁻¹) at 300 °C for 1 h. After cooling to 100 °C, the sample was saturated with NH₃, and excess physisorbed NH₃ was

removed by purging with Ar before initiating the desorption measurement. CO temperature-programmed desorption (CO-TPD) measurement was conducted on the Micromeritics Autochem 2920 connected to MS. The catalyst was saturated with CO for 0.5 h, flushed with Ar, and measured from 50 to 750 °C at 10 °C/min under Ar flow. H₂/D₂ exchange and C₂H₄ pulse hydrogenation experiments were also conducted on this platform, with the relevant mass-to-charge (m/z) signals monitored by the online MS for product analysis.

Supplementary Figures and Notes:

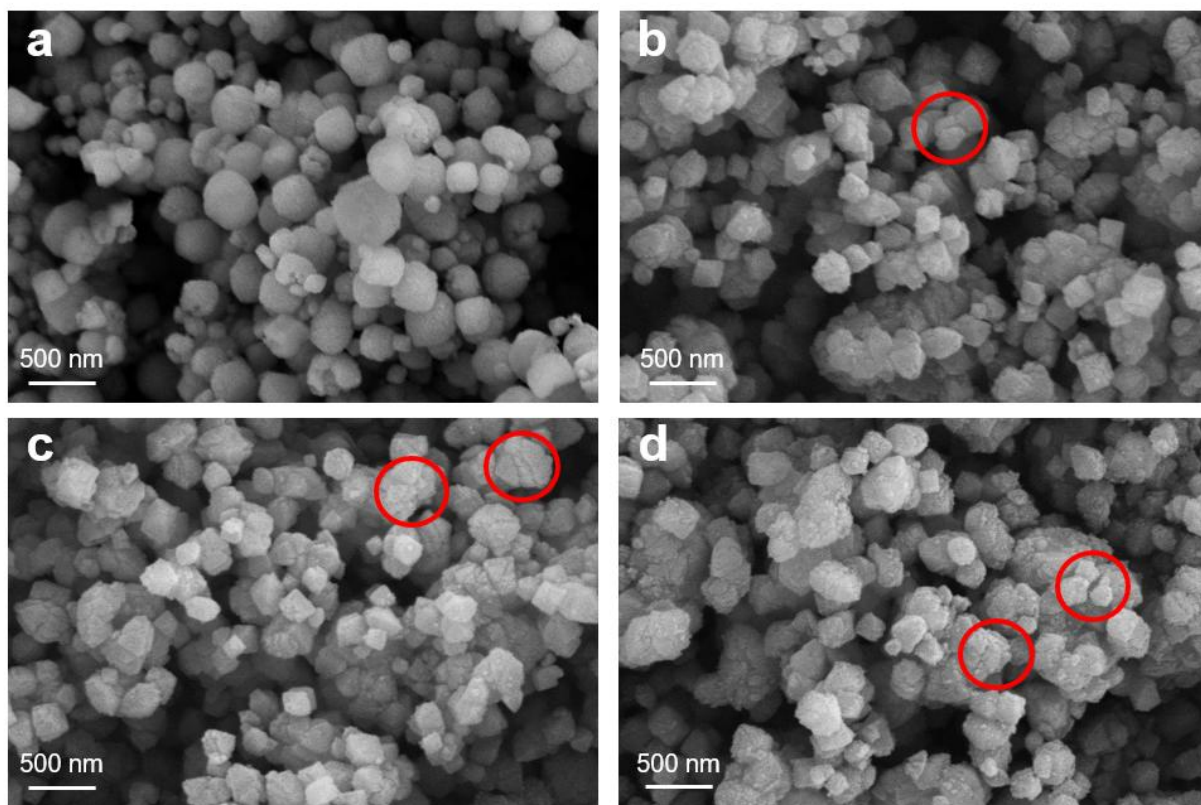


Supplementary Fig. 1 | Scheme illustration of alkali treatment with a TPAOH solution to etch β zeolite.



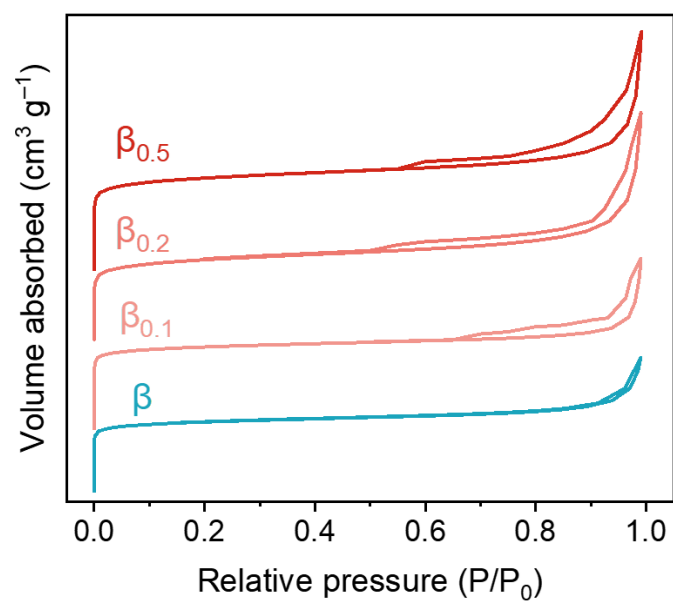
Supplementary Fig. 2 | FT-IR spectra of parent and β_a zeolites.

Notes: Two characteristic peaks around 524 cm⁻¹ and 575 cm⁻¹ of FT-IR spectra were contributed to the typical stretching modes of six- or five-membered rings of β zeolite. The positions of these characteristic peaks in β_a remained unchanged, confirming that the crystalline phase structure of β zeolite was retained.

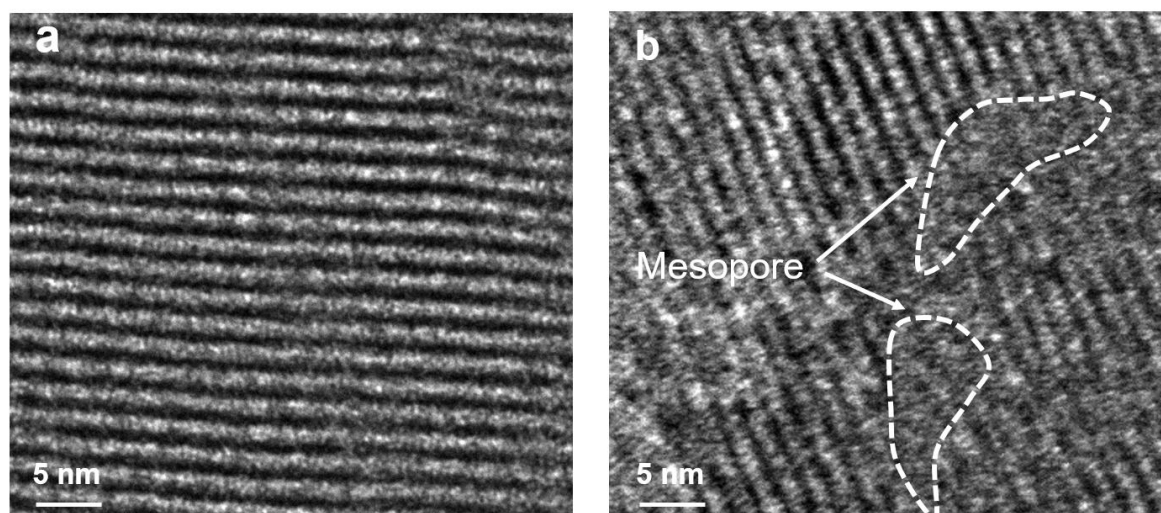


Supplementary Fig. 3 | SEM micrographs of (a) β , (b) $\beta_{0.1}$, (c) $\beta_{0.2}$, and (d) $\beta_{0.5}$.

Notes: The parent β exhibited smooth, ellipsoidal particles. After etching, the particles became progressively more faceted and cuboidal/rectangular in shape, with visibly roughened surfaces (especially highlighted by red circles in (c) and (d)). These morphological changes reflected the uniform dissolution and reconstruction of the zeolite framework induced by the mild alkaline treatment, which was consistent with the development of a hierarchical pore structure observed by N_2 physisorption and TEM.

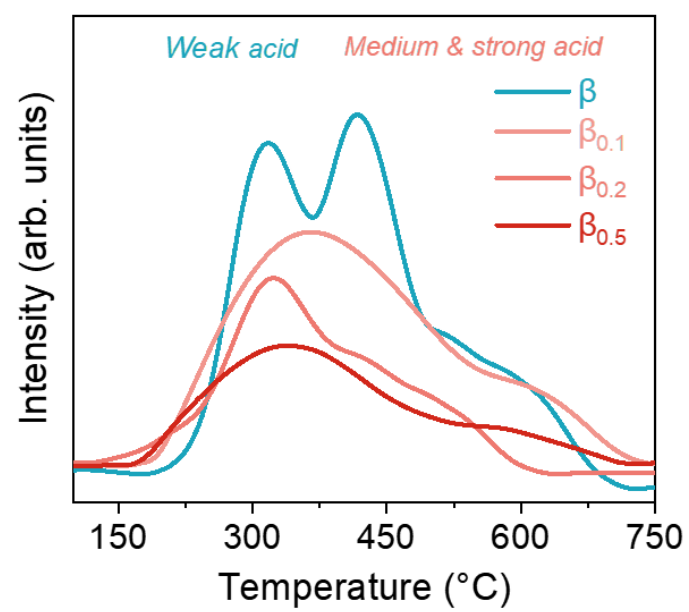


Supplementary Fig. 4 | N₂ adsorption-desorption isotherm curves of β and β_a .

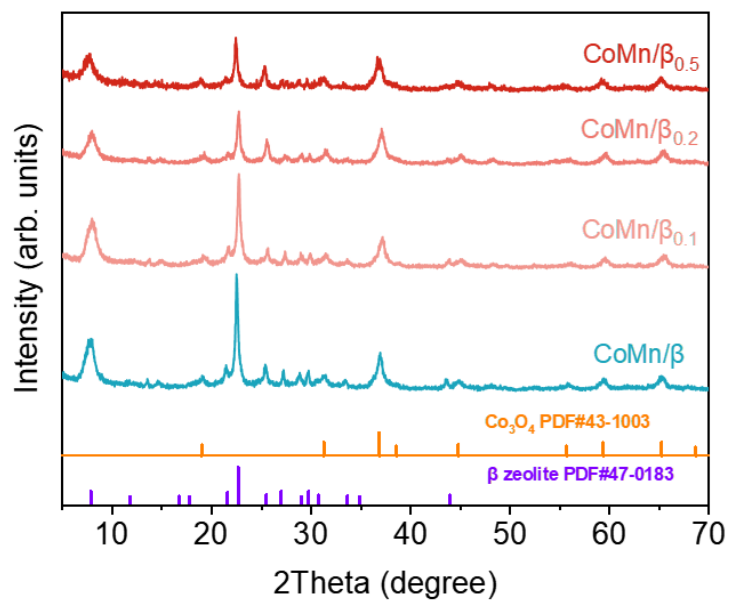


Supplementary Fig. 5 | TEM images of (a) β and (b) $\beta_{0.2}$.

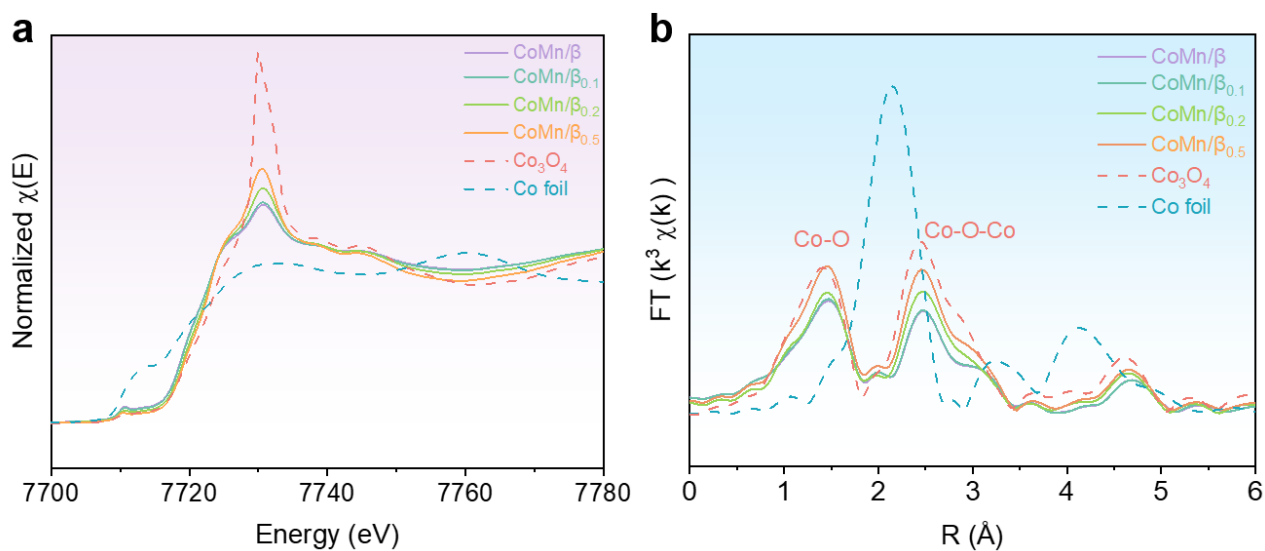
Notes: TEM analysis clearly revealed the coexistence of ordered micropores and hierarchical mesopores, corroborating the textural features observed from the N_2 adsorption–desorption isotherms. The β zeolite exhibited well-defined and highly ordered lattice fringes. In contrast, $\beta_{0.2}$ showed partially disrupted diffraction patterns and locally blurred lattice fringes, suggesting partial framework degradation and the formation of mesoporous domains with reduced long-range order¹.



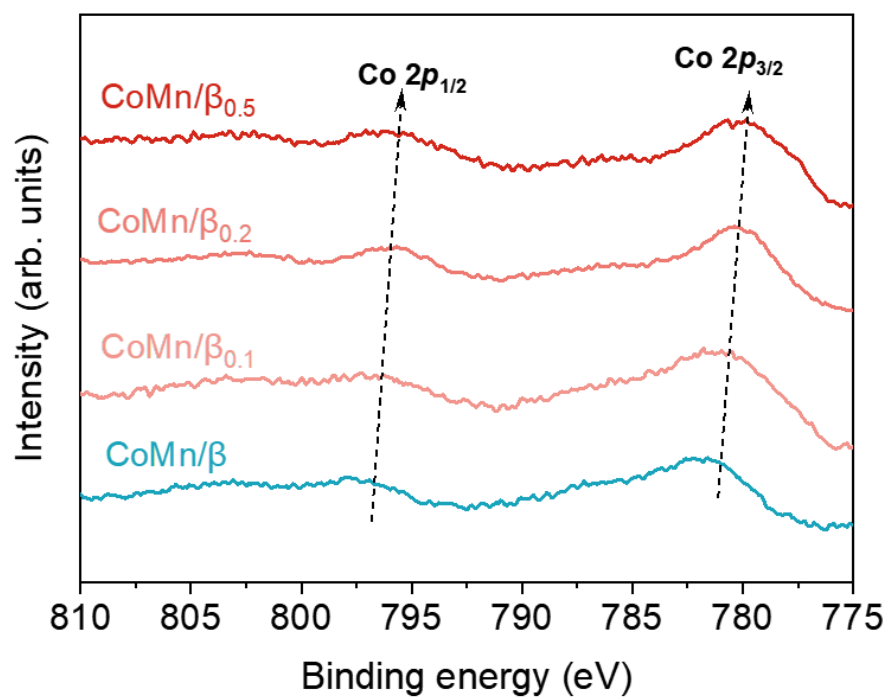
Supplementary Fig. 6 | NH₃-TPD profiles of β and β_a .



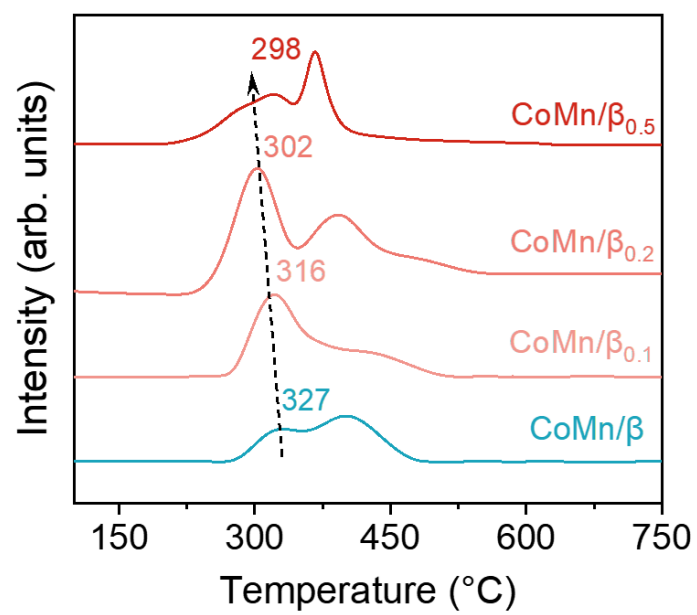
Supplementary Fig. 7 | XRD patterns of CoMn/β and CoMn/β_a catalysts.



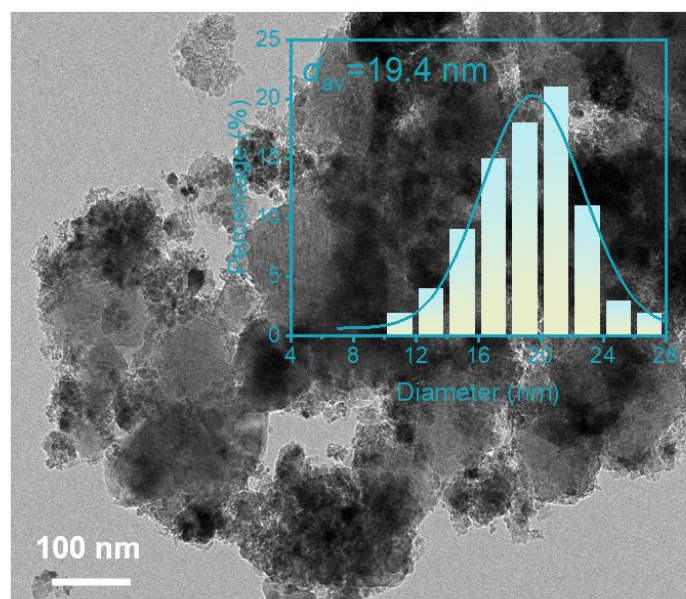
Supplementary Fig. 8 | (a) Co K-edge XANES spectra of CoMn/ β and CoMn/ β_a catalysts. (b) Co K-edge FT-EXAFS spectra of CoMn/ β and CoMn/ β_a catalysts.



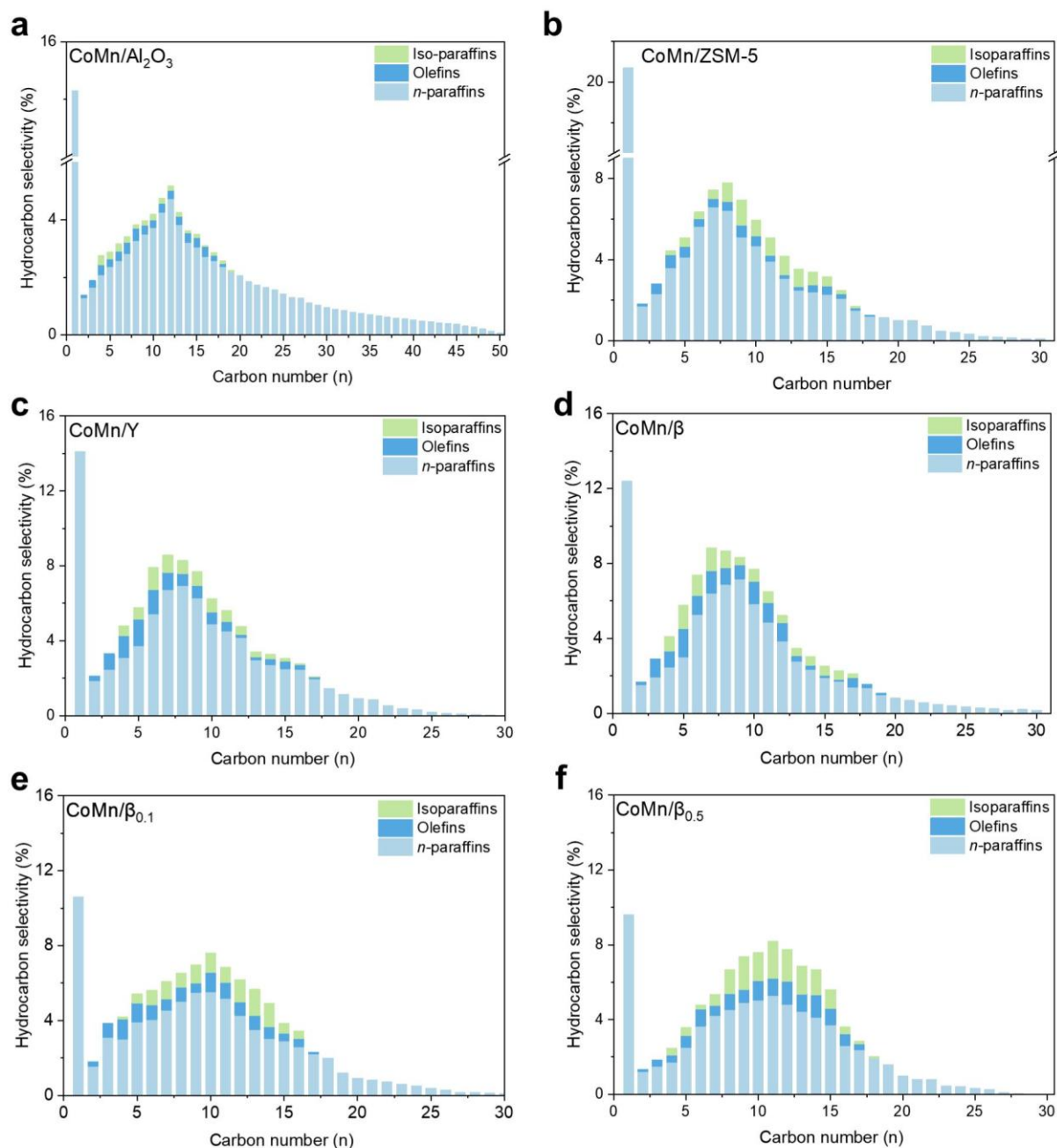
Supplementary Fig. 9 | XPS spectra of Co 2p.



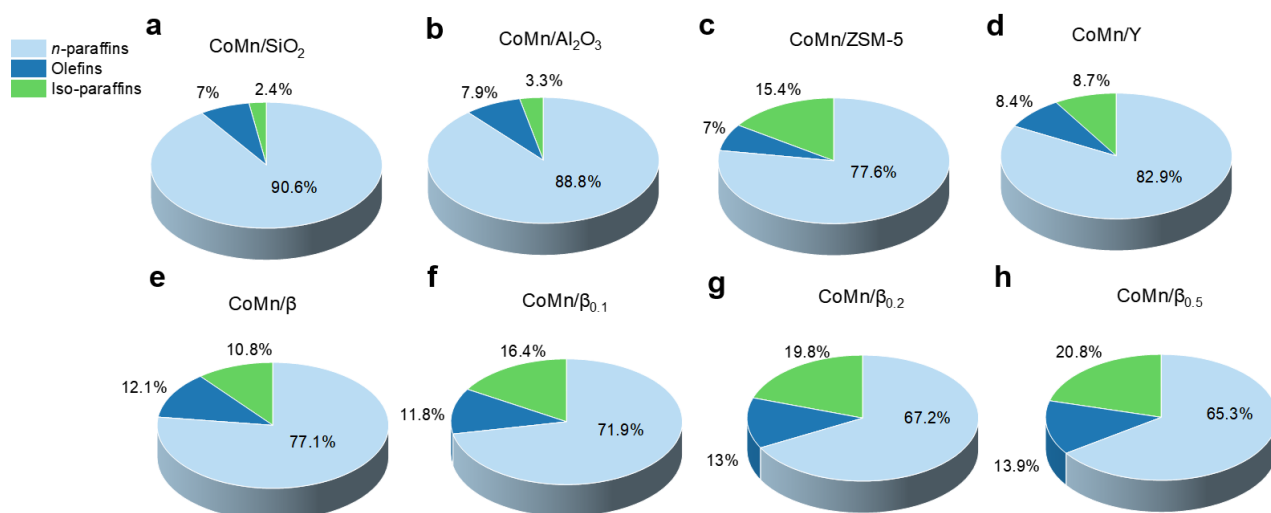
Supplementary Fig. 10 | H₂-TPR profiles of CoMn/β and CoMn/β_a catalysts.



Supplementary Fig. 11 | TEM image of CoMn/β.

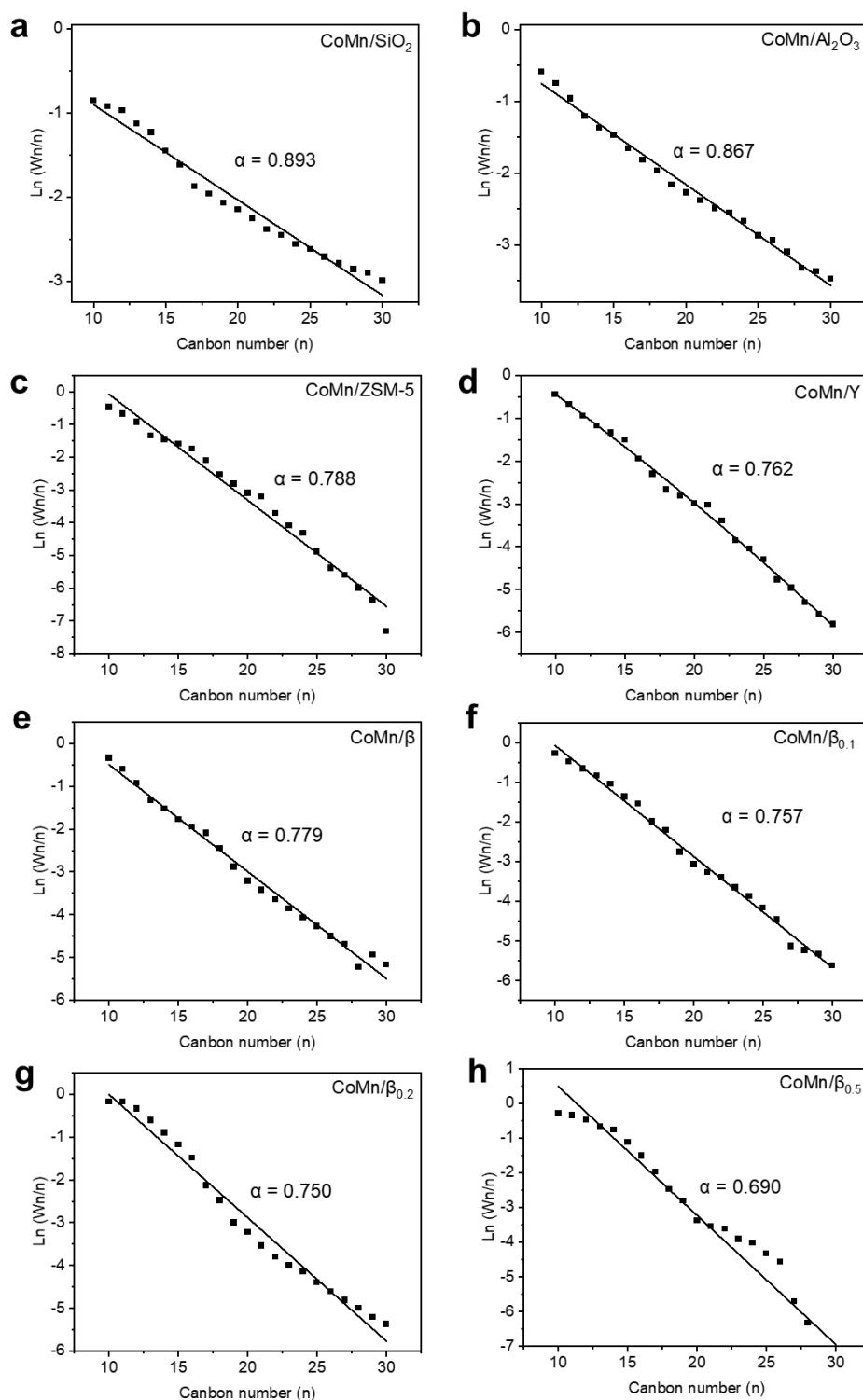


Supplementary Fig. 12 | Hydrocarbon distribution after 48 h reaction over (a) CoMn/Al₂O₃, (b) CoMn/ZSM-5, (c) CoMn/Y, (d) CoMn/ β , (e), and (f) CoMn/ $\beta_{0.5}$ catalysts. Reaction condition: 1.0 g of catalyst, H₂/CO = 2, P = 1.5 MPa, T = 230 °C, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

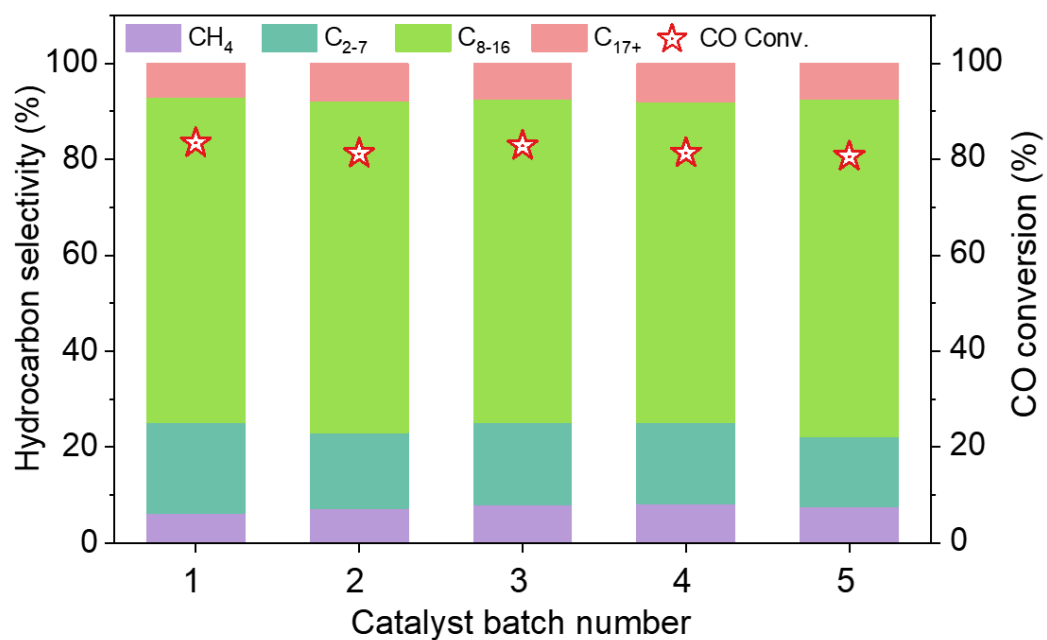


Supplementary Fig. 13 | Detailed hydrocarbon composition in the C_{8–16} over (a) CoMn/SiO₂, (b) CoMn/Al₂O₃, (c) CoMn/ZSM-5, (d) CoMn/Y, (e) CoMn/β, (f) CoMn/β_{0.1}, (g) CoMn/β_{0.1}, and (h) CoMn/β_{0.2} catalysts. Reaction condition: 1.0 g of catalyst, H₂/CO = 2, P = 1.5 MPa, T = 230 °C, WHSV = 3000 mL g_{cat}^{−1} h^{−1}.

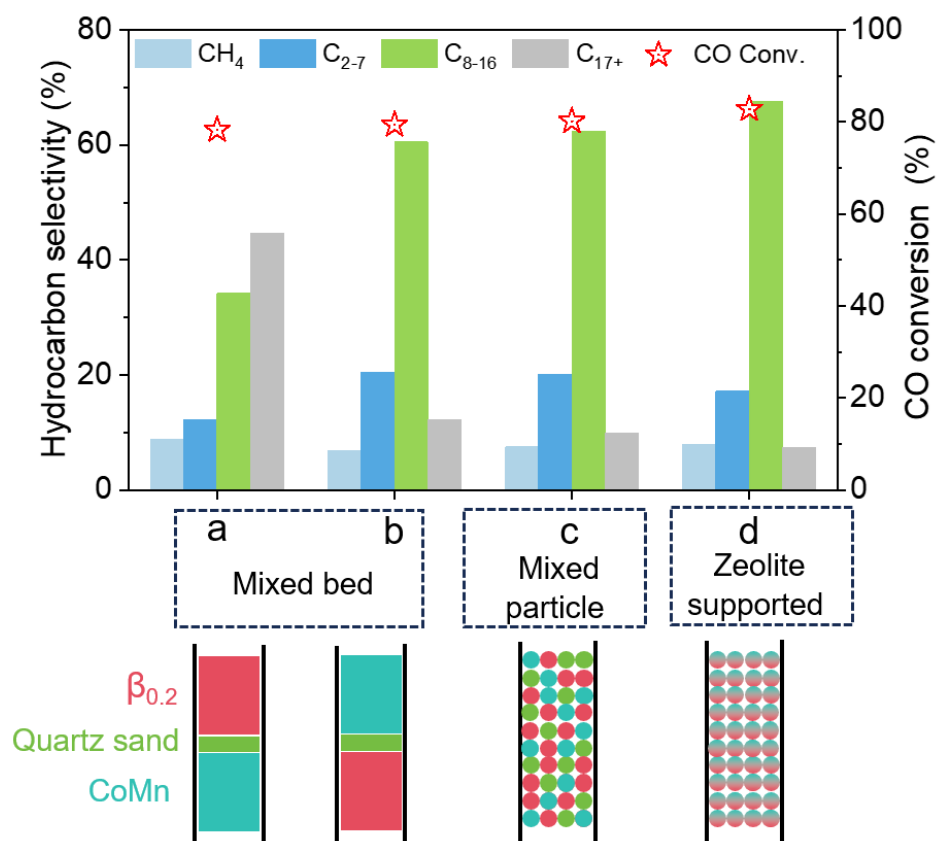
Notes: The detailed hydrocarbon distribution and composition in C_{8–16} clearly illustrated the critical influence of support properties: conventional oxide supports such as SiO₂ and Al₂O₃ produced almost no iso-paraffins, yielding predominantly linear hydrocarbons. In contrast, zeolite supports, particularly the TPAOH-etched β zeolites, generated a significant fraction of iso-paraffins, and this proportion increased systematically with higher etching concentration. This demonstrated that the optimized mesoporous structure synergizes with the moderated acid sites to effectively promote secondary isomerization of the reaction intermediates, thereby precisely steering product selectivity toward the jet fuel range.



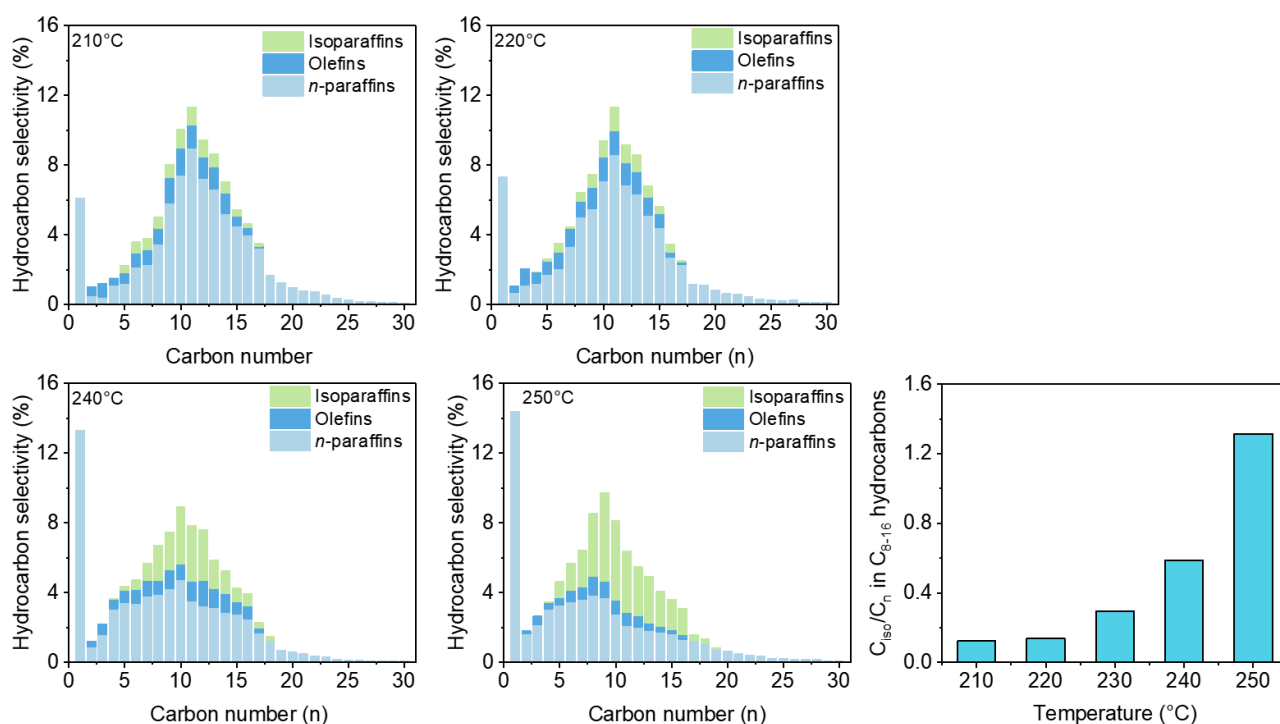
Supplementary Fig. 14 | ASF distribution plots for (a) CoMn/SiO₂, (b) CoMn/Al₂O₃, (c) CoMn/ZSM-5, (d) CoMn/Y, (e) CoMn/ β , (f) CoMn/ $\beta_{0.1}$, (g) CoMn/ $\beta_{0.2}$, and (h) CoMn/ $\beta_{0.5}$ catalysts. The chain growth probability (α) evaluated from the slope of straight lines using C_{10–30} products.



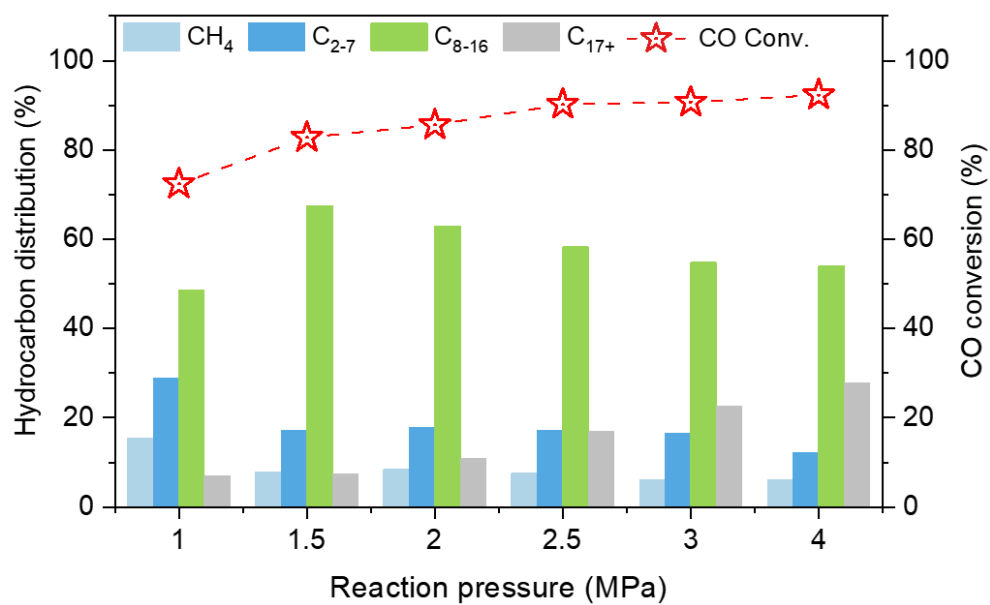
Supplementary Fig. 15 | Reproducibility of FT performance over different batches of the CoMn/ β _{0.2} catalyst. Reaction condition: 1.0 g of catalyst, H₂/CO = 2, T = 230 °C, P = 1.5 MPa, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.



Supplementary Fig. 16 | Effects of metal-zeolite assembly strategy on catalytic properties under the same reaction conditions. Equal total cobalt mass was loaded into the reactor for zeolite-supported CoMn/ $\beta_{0.2}$, dual-bed and mixed-particle catalysts for controlled comparison.

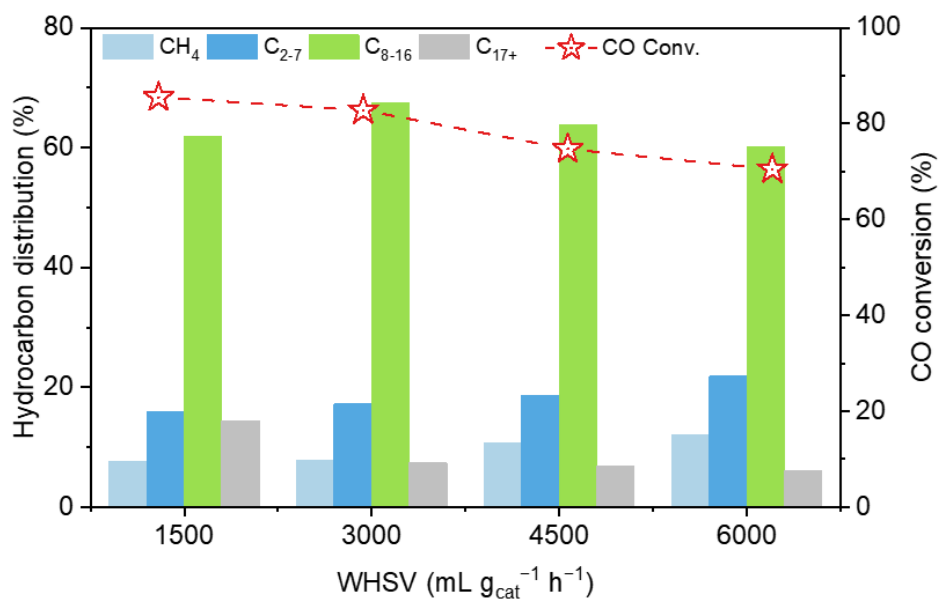


Supplementary Fig. 17 | Effects of reaction temperature on hydrocarbon distribution and the ratio of isoparaffins (C_{iso}) to n -paraffins (C_n) in C_{8-16} over the CoMn/ $\beta_{0.2}$ catalyst. Reaction conditions: 1.0 g of catalyst, $H_2/CO = 2$, $P = 1.5$ MPa, $WHSV = 3000$ mL $g_{cat}^{-1} h^{-1}$.

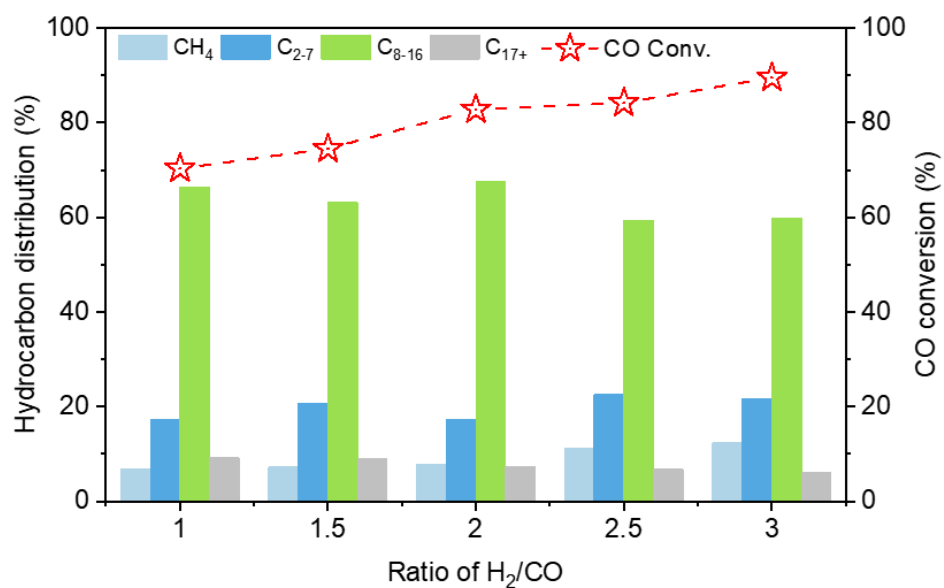


Supplementary Fig. 18 | CO conversion and hydrocarbon distribution at different reaction pressure.

Reaction condition: 1.0 g of catalyst, H₂/CO = 2, T = 230 °C, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

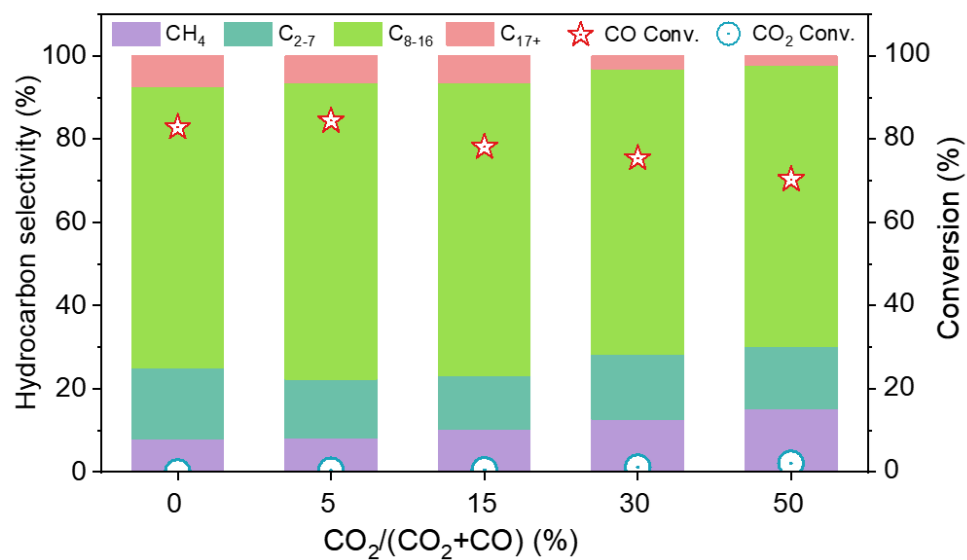


Supplementary Fig. 19 | CO conversion and hydrocarbon distribution at different WHSV. Reaction condition: 1.0 g of catalyst, H₂/CO = 2, P = 1.5 MPa, T = 230 °C.



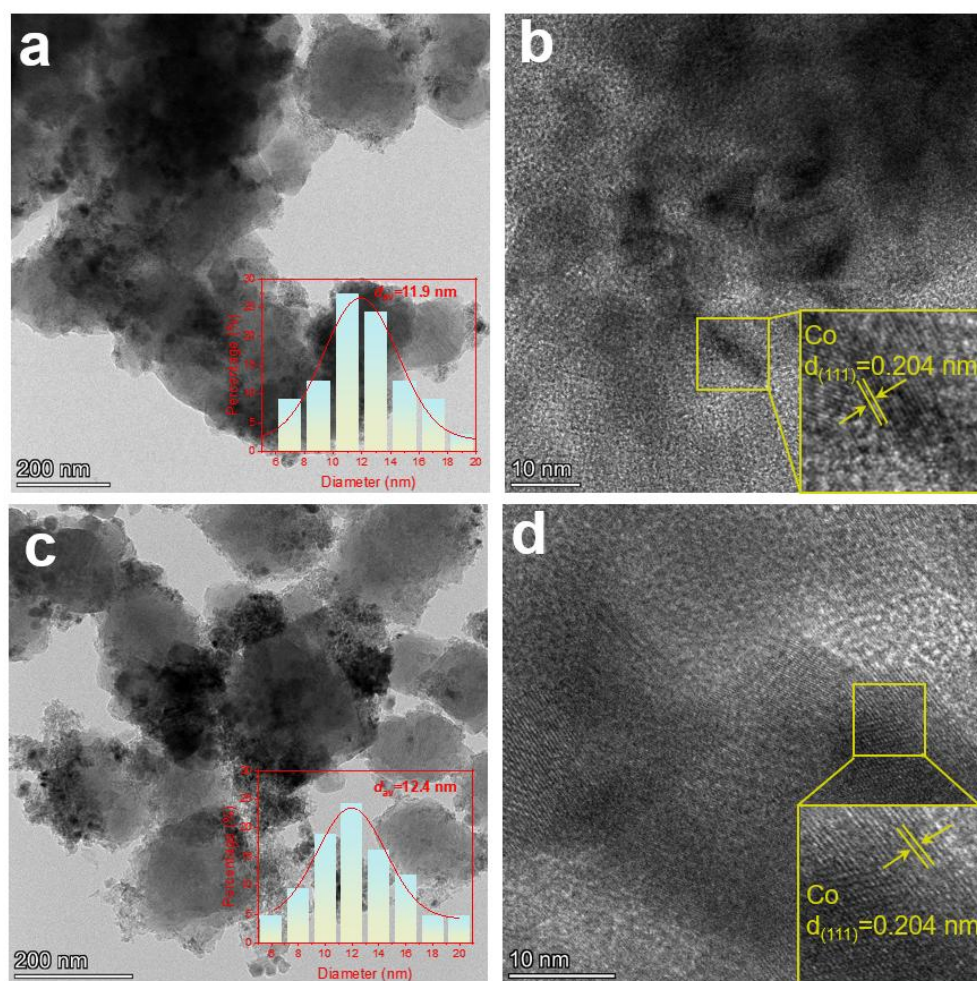
Supplementary Fig. 20 | CO conversion and hydrocarbon distribution at different H₂/CO ratios.

Reaction condition: 1.0 g of catalyst, P = 1.5 MPa, T = 230 °C, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

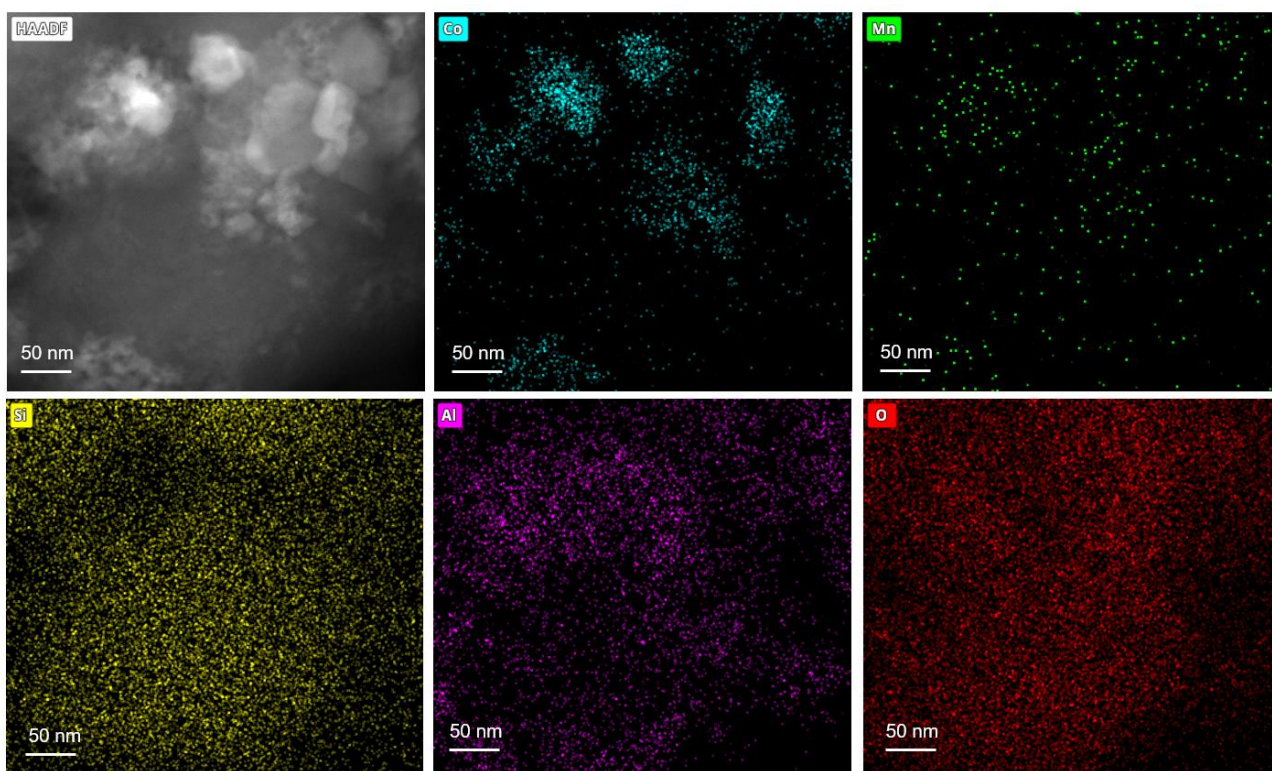


Supplementary Fig. 21 | Effect of CO₂ fraction in mixed-feed FT over CoMn/β_{0.2}. Reaction condition:

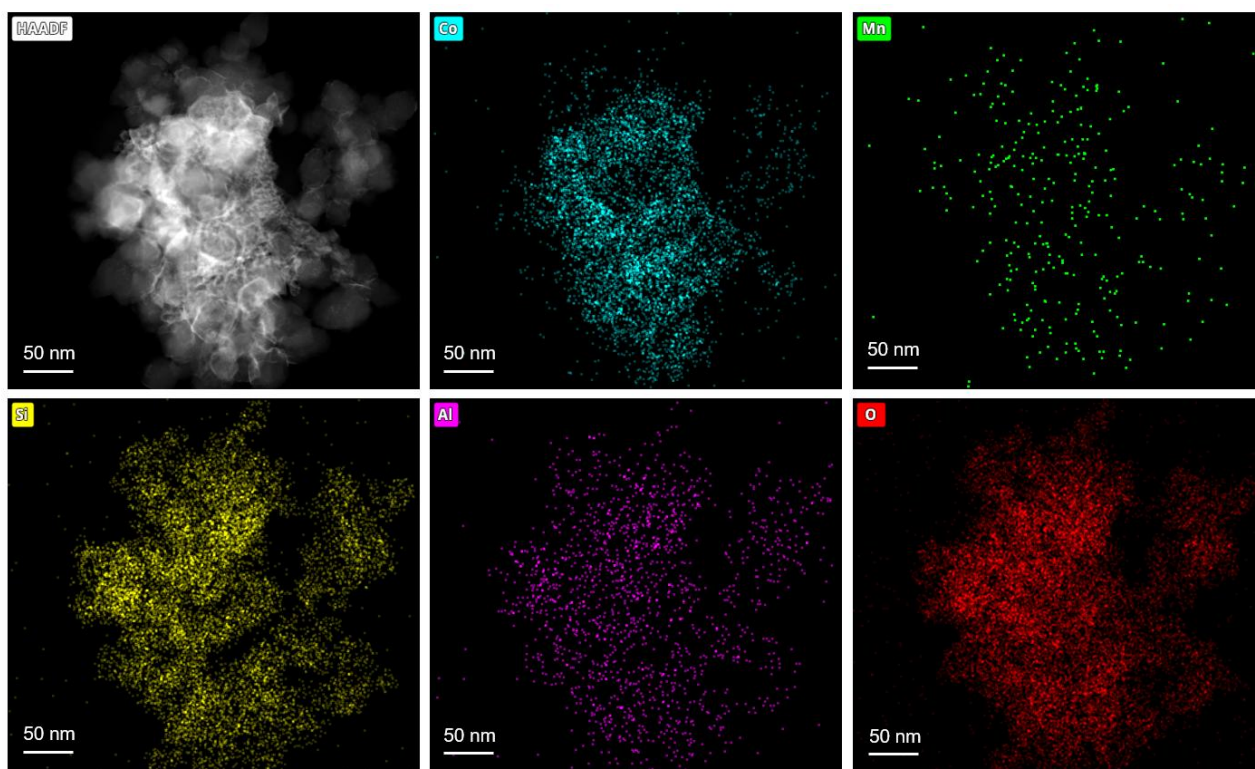
1.0 g of catalyst, H₂/CO_x = 2. P = 1.5 MPa, T = 230 °C, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.



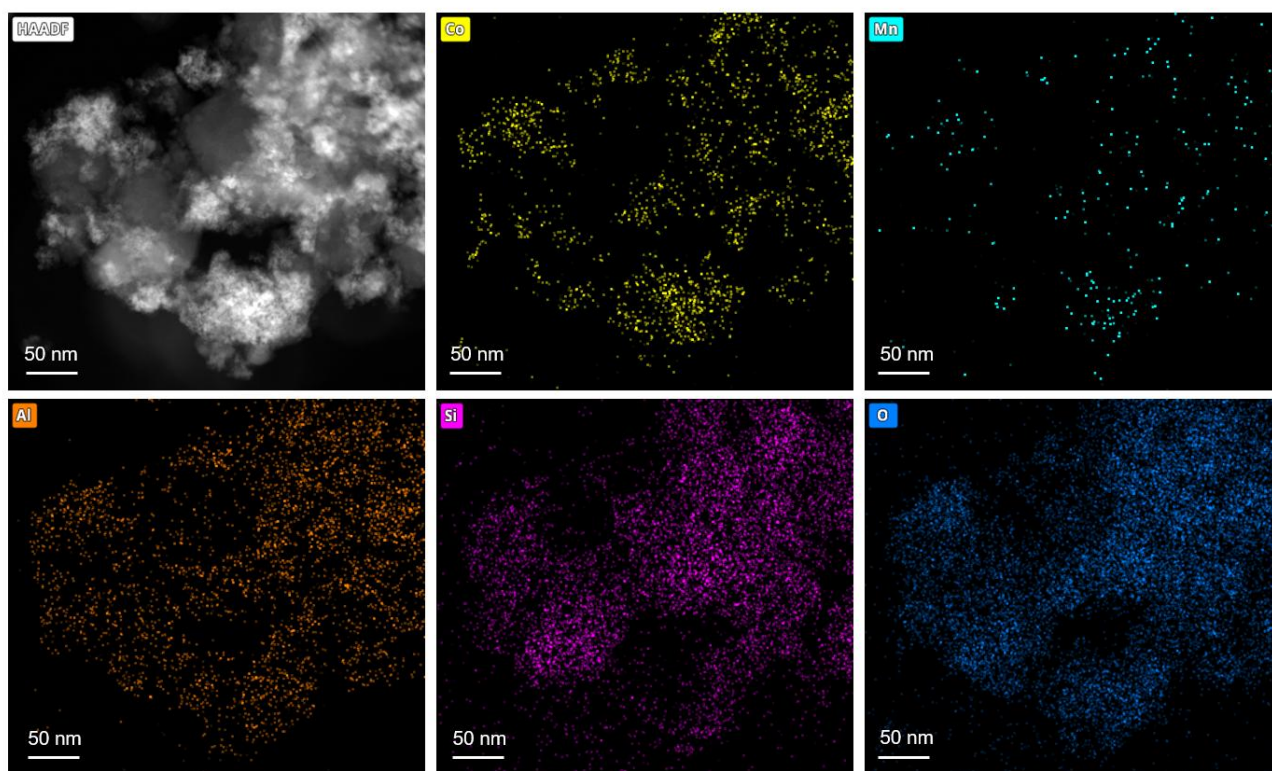
Supplementary Fig. 22 | (a, c) TEM images and (b, d) HRTEM images of the reduced and the spent CoMn/ $\beta_{0.2}$ catalyst.



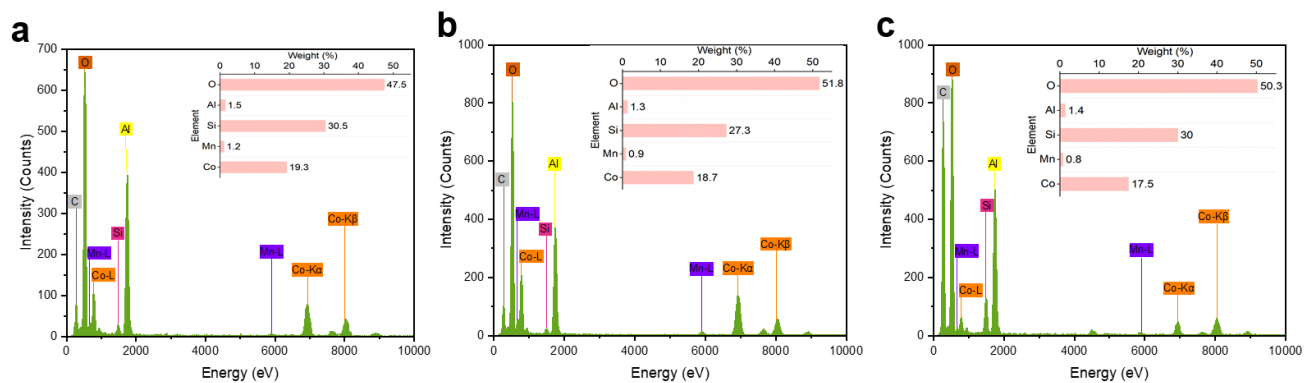
Supplementary Fig. 23 | HAADF-STEM images and the corresponding EDX mapping for the fresh unreduced CoMn/ $\beta_{0.2}$ catalyst.



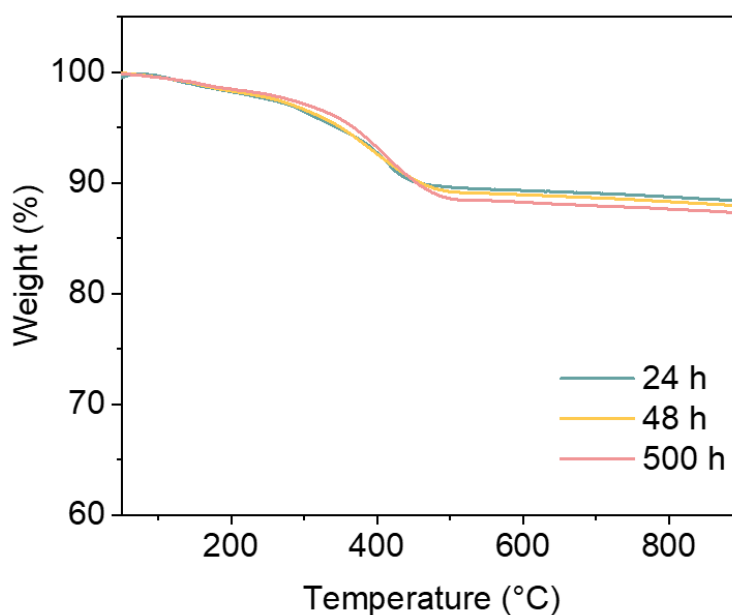
Supplementary Fig. 24 | HAADF-STEM images and the corresponding EDX mapping for the reduced CoMn/ $\beta_{0.2}$ catalyst after 48 h FT reaction.



Supplementary Fig. 25 | HAADF-STEM images and the corresponding EDX mapping for the CoMn/β_{0.2} catalyst after 48 h FT reaction.

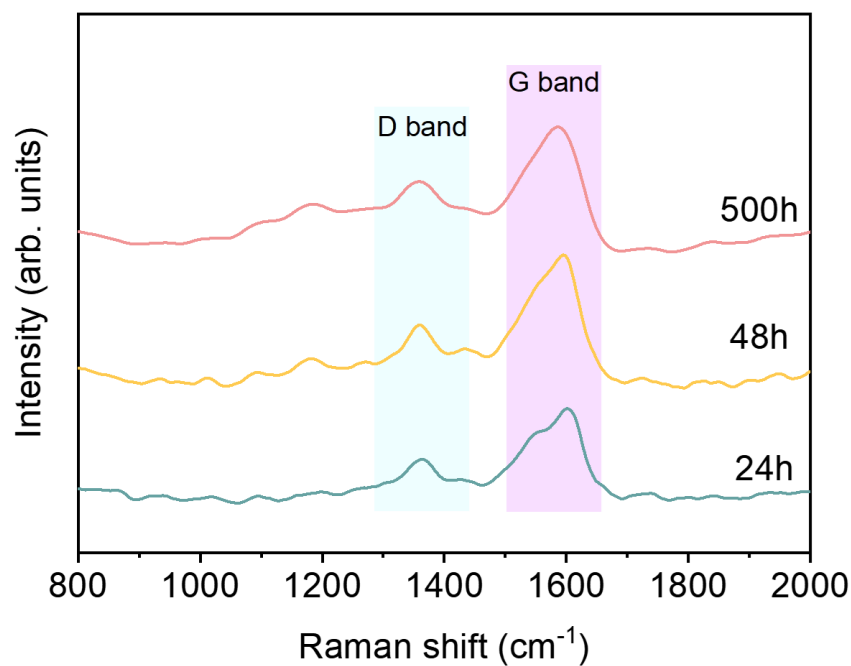


Supplementary Fig. 26 | EDX spectra of the (a) fresh, (b) reduced, and (c) spent CoMn/ $\beta_{0.2}$ catalyst after 48 h FT reaction.



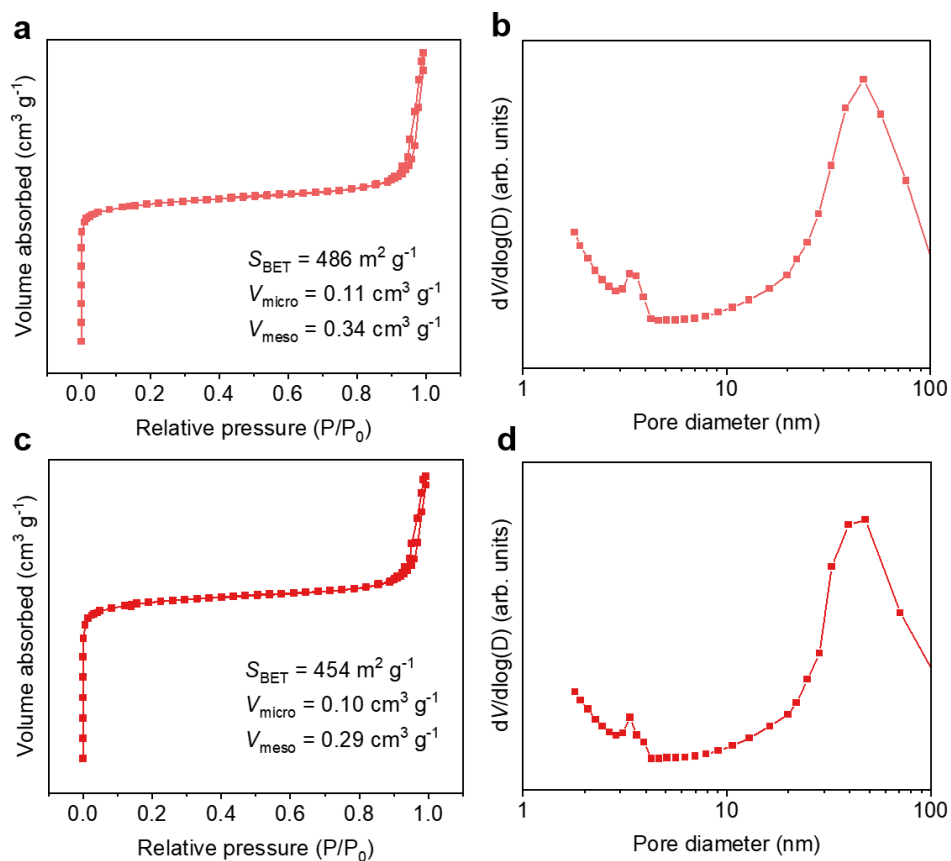
Supplementary Fig. 27 | TGA analysis of spent CoMn/β_{0.2} catalyst collected after different reaction durations.

Note: Thermogravimetric analysis (TGA) showed that the weight-loss curves of the spent CoMn/β_{0.2} catalyst collected after 24 h, 48 h, and 500 h of reaction exhibited a single distinct weight-loss step around 400 °C corresponding to the oxidative combustion of deposited carbon. The final residual weights of the three curves in the high-temperature region (> 600 °C) are nearly identical, indicating that the amount of surface carbon deposition remains stable across different reaction durations and does not increase significantly with prolonged operation.

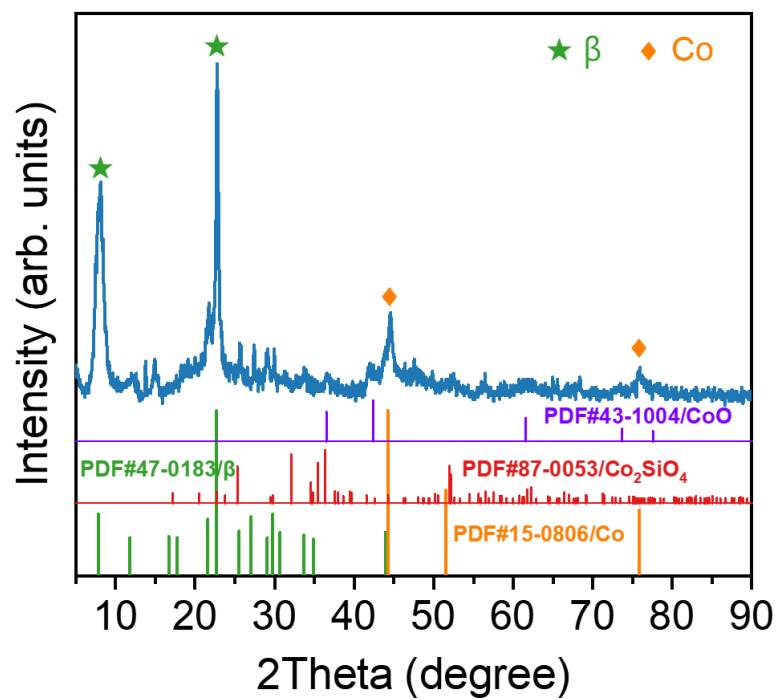


Supplementary Fig. 28 | Raman analysis of spent CoMn/ $\beta_{0.2}$ catalyst collected after different reaction durations.

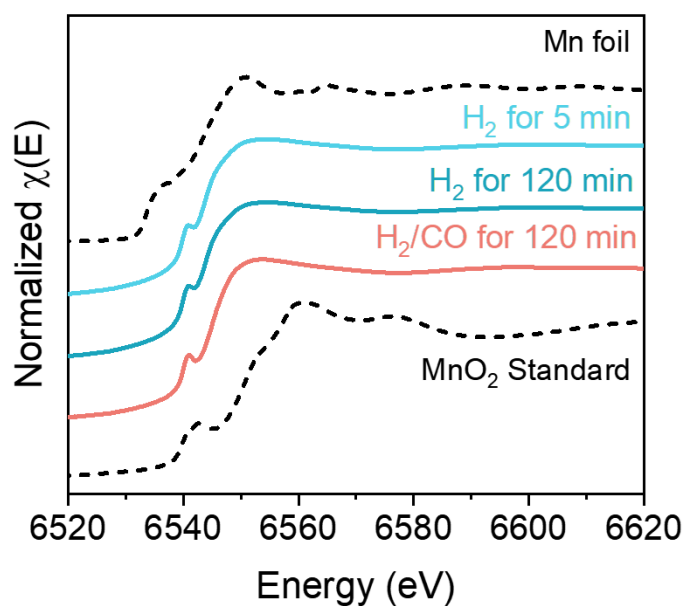
Note: The Raman analysis results indicated that after 24 h, 48 h, and 500 h reactions, the D band ($\sim 1350 \text{ cm}^{-1}$) and G band ($\sim 1580 \text{ cm}^{-1}$) peak positions and shapes of the carbon deposits on the spent CoMn/ $\beta_{0.2}$ catalyst surface exhibit high similarity, and no significant increase in intensity was observed over time. Furthermore, after 48 h and 500 h of reaction, the peak positions and shapes remained similar.



Supplementary Fig. 29 | (a, c) N₂ adsorption-desorption isotherms and (b, d) BJH adsorption pore size distributions of the reduced and spent CoMn/β_{0.2} catalysts.

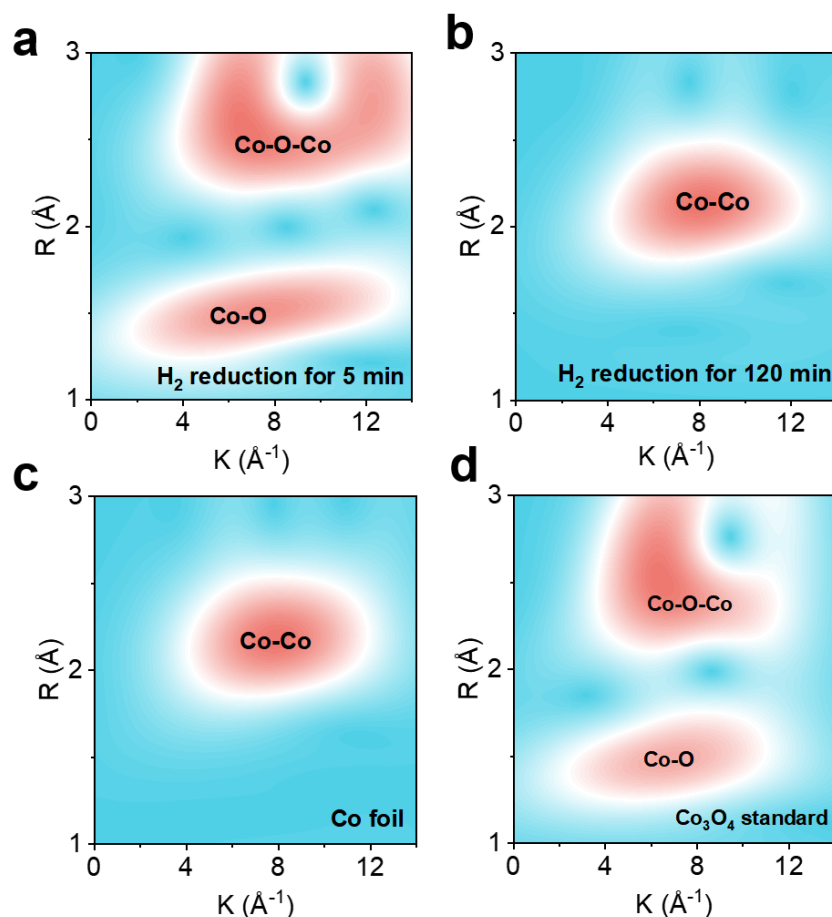


Supplementary Fig. 30 | XRD pattern of the spent CoMn/ $\beta_{0.2}$ catalyst.



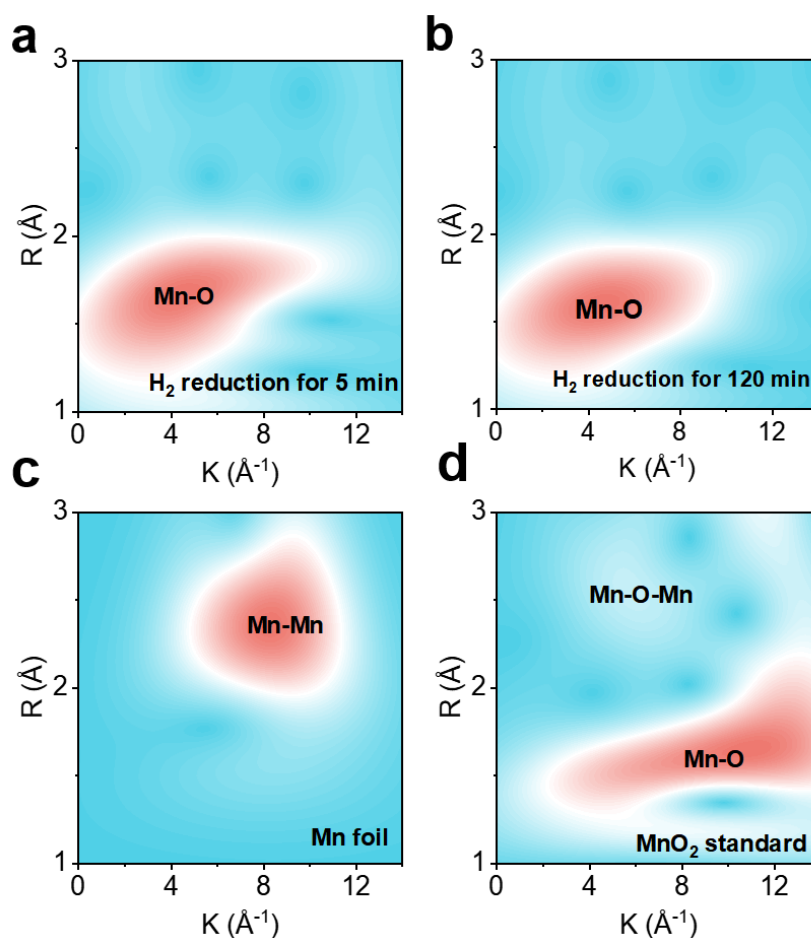
Supplementary Fig. 31 | Operando XANES spectra at the Mn K-edge under H₂ atmosphere at 400 °C and 0.1 MPa and H₂/CO reaction at 230 °C and 1.5 MPa.

Notes: The Mn K-edge Operando XANES spectra under both reduction (400 °C, H₂) and reaction (230 °C, H₂/CO) conditions remain consistently between the references of metallic Mn foil and MnO₂, indicating that the Mn species retained a partially oxidized Mn^{δ+} state without significant valence change. This demonstrates that the Mn^{δ+} species is structurally stable throughout reduction and reaction, and is strongly anchored on the β_{0.2} support in an atomically dispersed form.



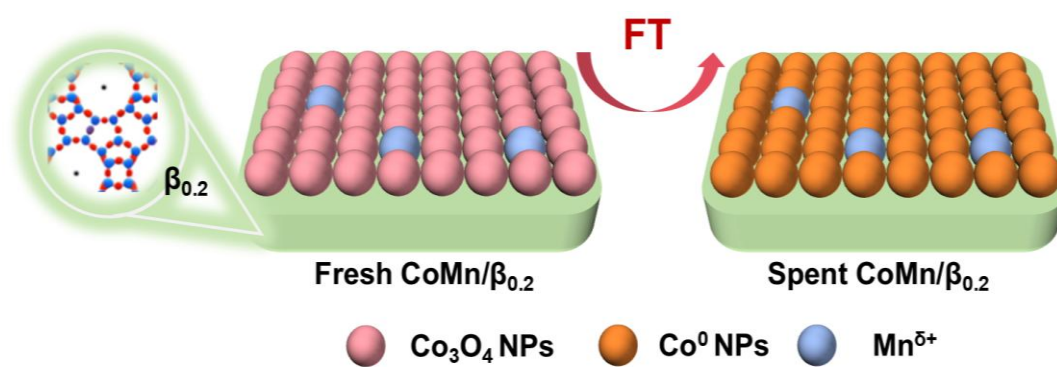
Supplementary Fig. 32 | WT analysis of Operando FT EXAFS spectra after 5 min (a) and 120 (b) min on stream of H₂ reduction, Co foil (c), and Co₃O₄ standard (d) for Co K-edge.

Notes: Wavelet-transform (WT) EXAFS analysis visually tracked the structural evolution of the active sites in the CoMn/ $\beta_{0.2}$ catalyst during reduction. The contour plot at 5 min showed coexisting Co–O and Co–O–Co signals, indicating incomplete reduction of Co₃O₄. After 120 min of reduction, only a prominent Co–Co scattering peak appeared, whose position and intensity match well with the Co foil reference, confirming the complete transformation of cobalt species into stable metallic Co⁰ nanoparticles.

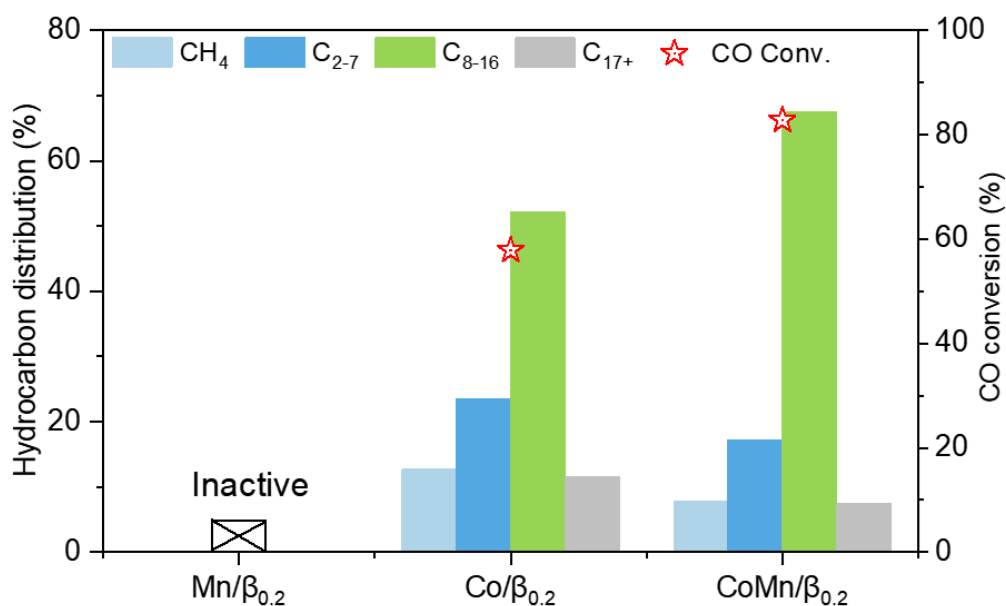


Supplementary Fig. 33 | WT analysis of Operando FT EXAFS spectra after 5 min (a) and 120 (b) min on stream of H_2 reduction, Mn foil (c), and MnO_2 standard (d) for Mn K-edge.

Note: WT analysis of the Mn K-edge revealed that the catalyst at the early reduction stage (5 min) and after full reduction (120 min), exhibiting only an Mn–O coordination peak with no Mn–Mn metallic bonding signal. This directly demonstrated that the $\text{Mn}^{\delta+}$ species remained atomically dispersed and valence-stable under working conditions, constituting a key component of the dual-site synergistic catalyst.

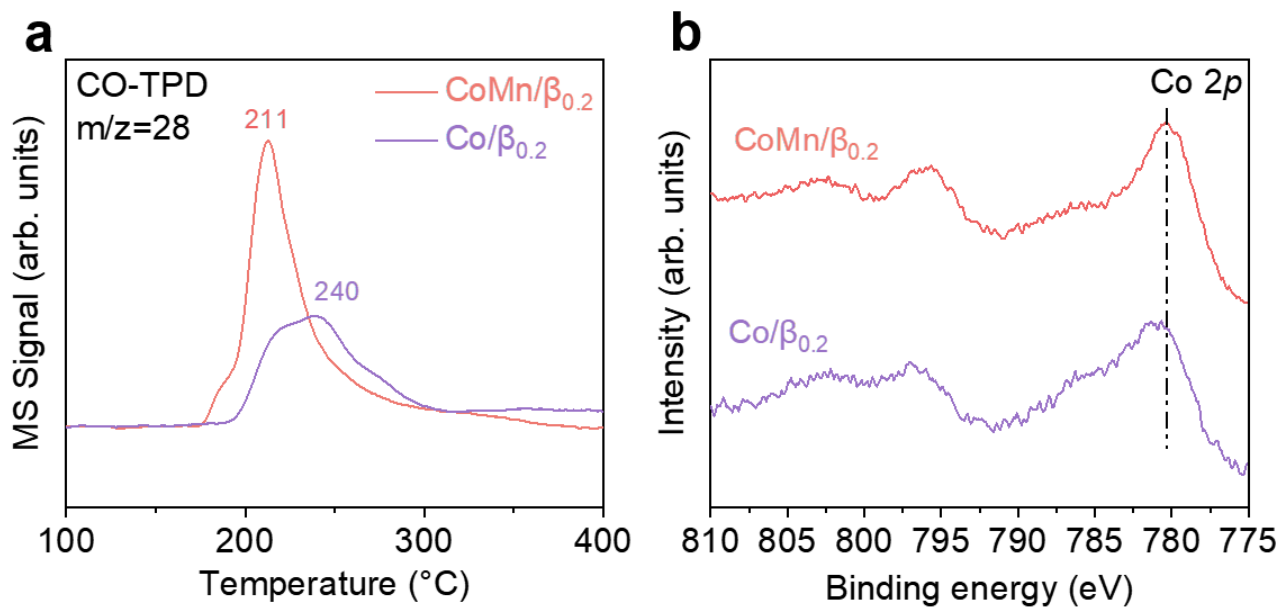


Supplementary Fig. 34 | Schematic illustration of structure evolution of the catalyst during FT.

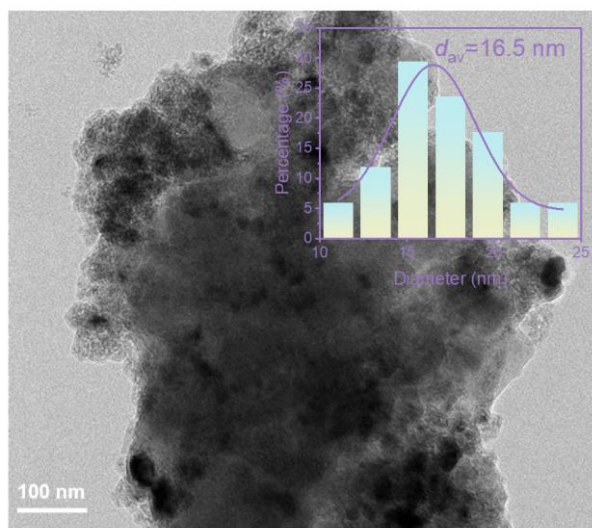


Supplementary Fig. 35 | CO conversion and hydrocarbon selectivity over various catalysts. Reaction condition: 1.0 g of catalyst, H₂/CO = 2, P = 1.5 MPa, T = 230 °C, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

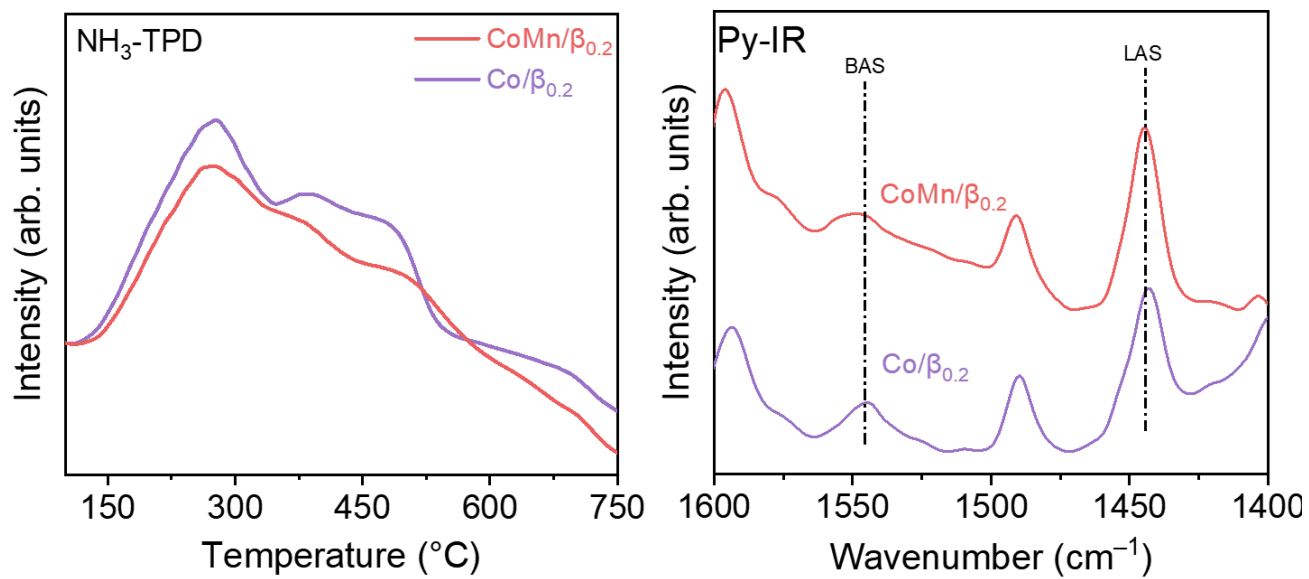
Note: To unambiguously verify the dual site synergy in the CoMn/β_{0.2} catalyst, we prepared the Mn only (Mn/β_{0.2}) and Co only (Co/β_{0.2}) catalysts using exactly the same synthesis method for comparison. Mn/β_{0.2} exhibits negligible CO conversion activity, while Co/β_{0.2} shows only moderate activity with high methane selectivity and limited yield of the target C₈₋₁₆ fraction. In contrast, the bimetallic CoMn/β_{0.2} delivers markedly enhanced CO conversion together with the highest selectivity toward the C₈₋₁₆ fraction. This result provides direct evidence that a pronounced synergy emerges only when stable Co⁰ nanoparticles coexist with atomically dispersed Mn^{δ+} sites, and that both components are essential for driving the high-performance FT synthesis.



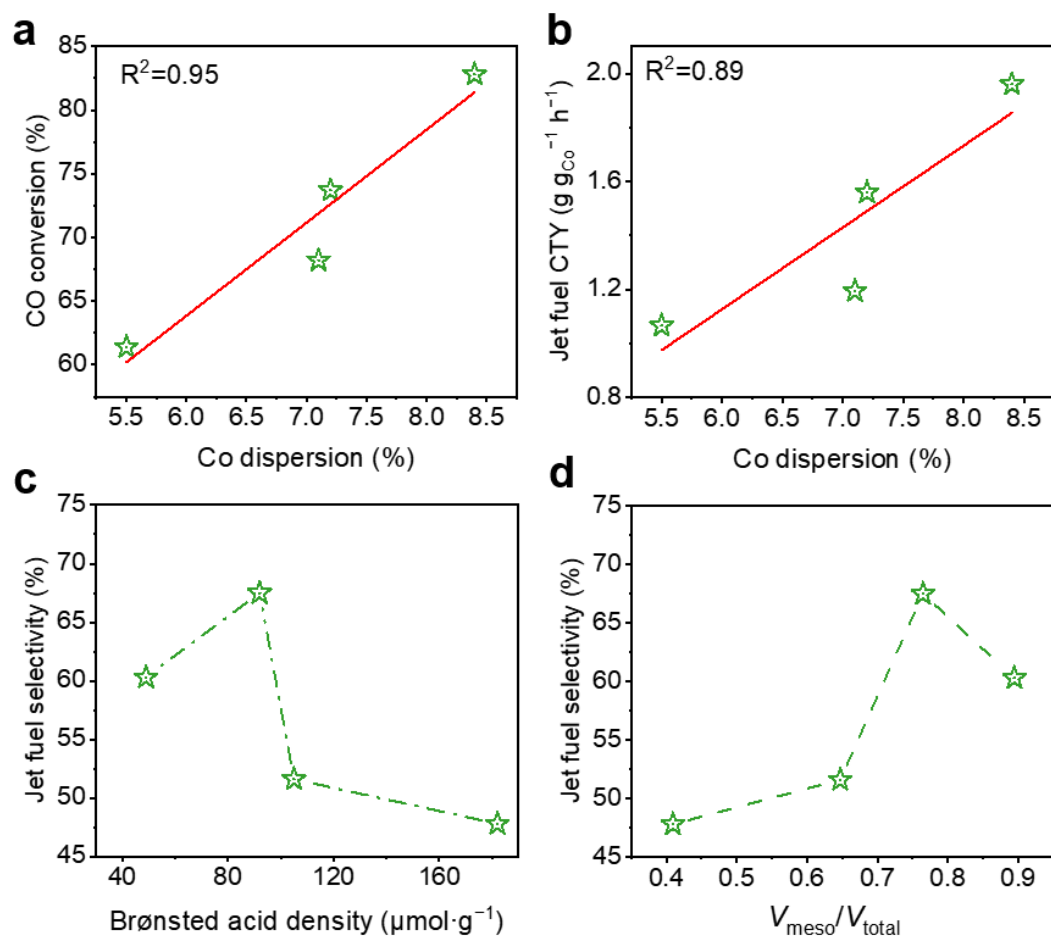
Supplementary Fig. 36 | (a) CO-TPD profiles and (b) XPS spectra of Co 2p of $\text{Co}/\beta_{0.2}$ and $\text{CoMn}/\beta_{0.2}$ catalysts.



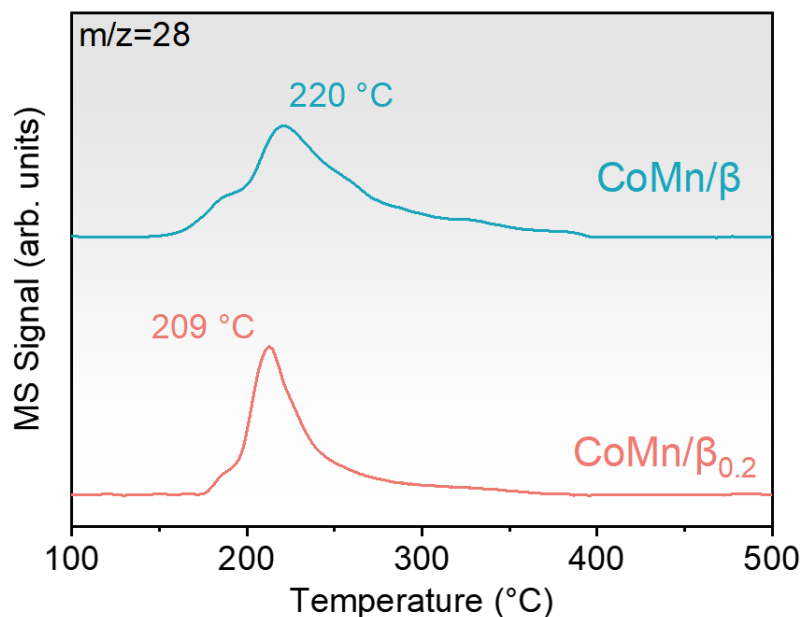
Supplementary Fig. 37 | TEM image of Co/β_{0.2}.



Supplementary Fig. 38 | (a) NH₃-TPD profiles and (b) Py-IR spectra of Co/β_{0.2} and CoMn/β_{0.2} catalysts.

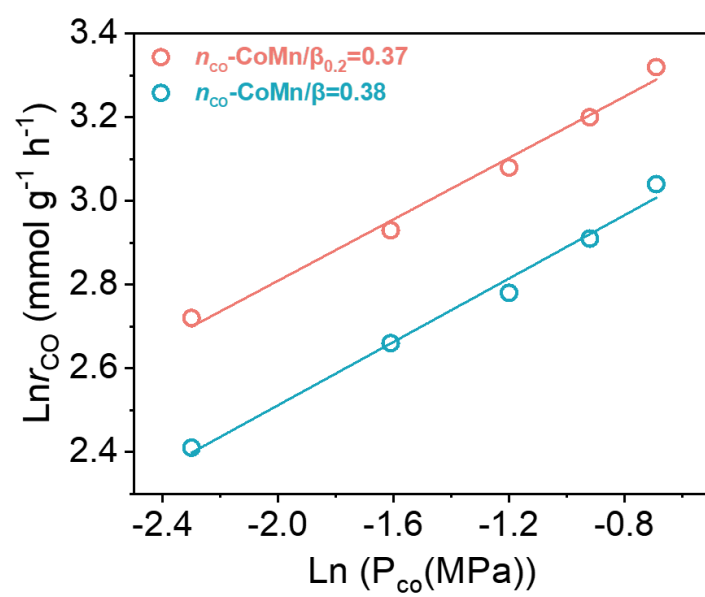


Supplementary Fig. 39 | (a) Correlation between Co dispersion and CO conversion. (b) Correlation between Co dispersion and jet fuel CTY. (c) Effect of Brønsted acid density on jet fuel selectivity. (d) Effect of mesopore volume ratio ($V_{\text{meso}}/V_{\text{total}}$) on jet fuel selectivity.

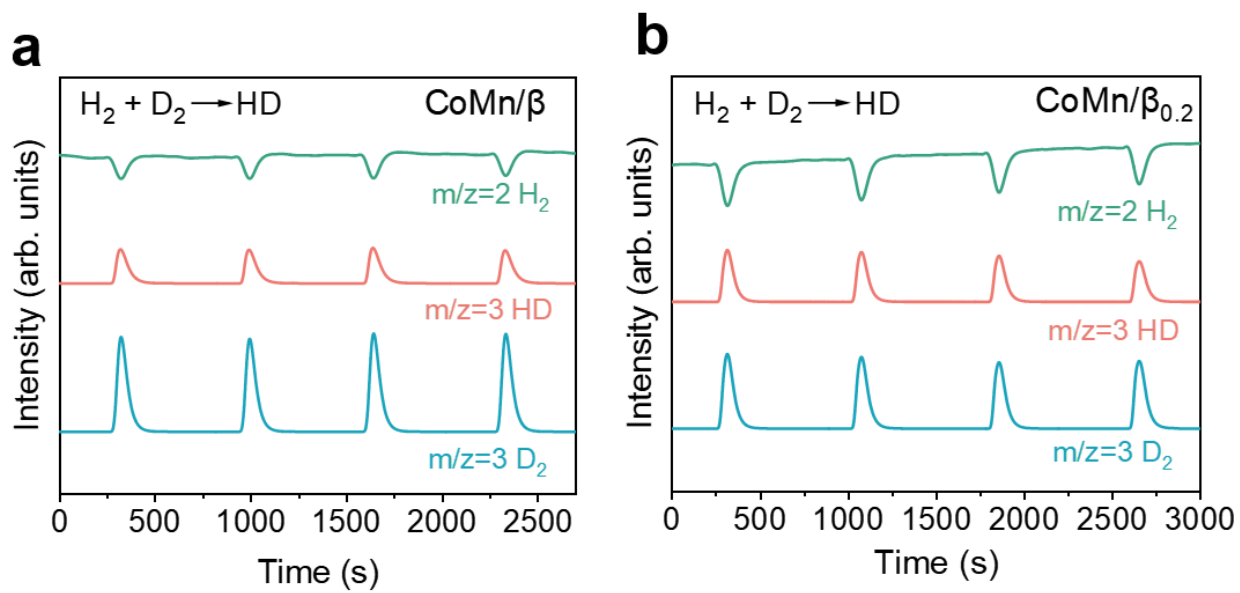


Supplementary Fig. 40 | CO-TPD profiles of the reduced catalysts.

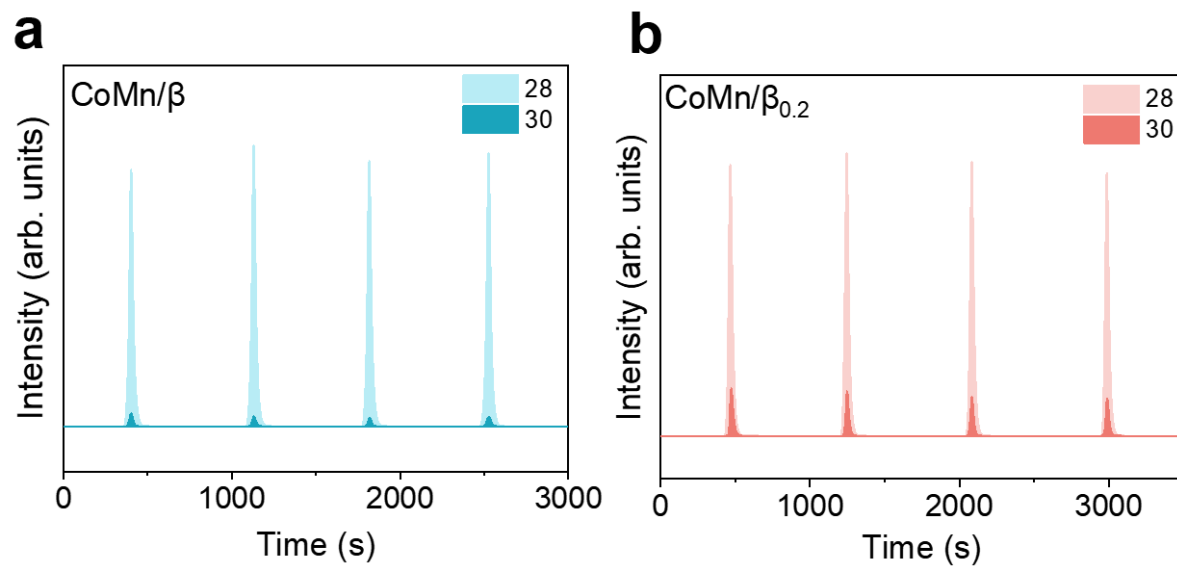
Note: CO temperature-programmed desorption (CO-TPD) of reduced CoMn/β and CoMn/β_{0.2} showed broad desorption features at 150–350 °C, attributed to recombination of surface C* and O* species on Co⁰ NPs^{2,3}. Compared to CoMn/β, the CO desorption peak for CoMn/β_{0.2} shifts toward lower temperature (from ~220 °C to ~209 °C) and becomes sharper. This indicates that the β_{0.2} support, modified by TPAOH etching, significantly weakens the binding energy between the Co⁰ NPs and CO, making CO easier to activate and desorb on the catalyst surface.



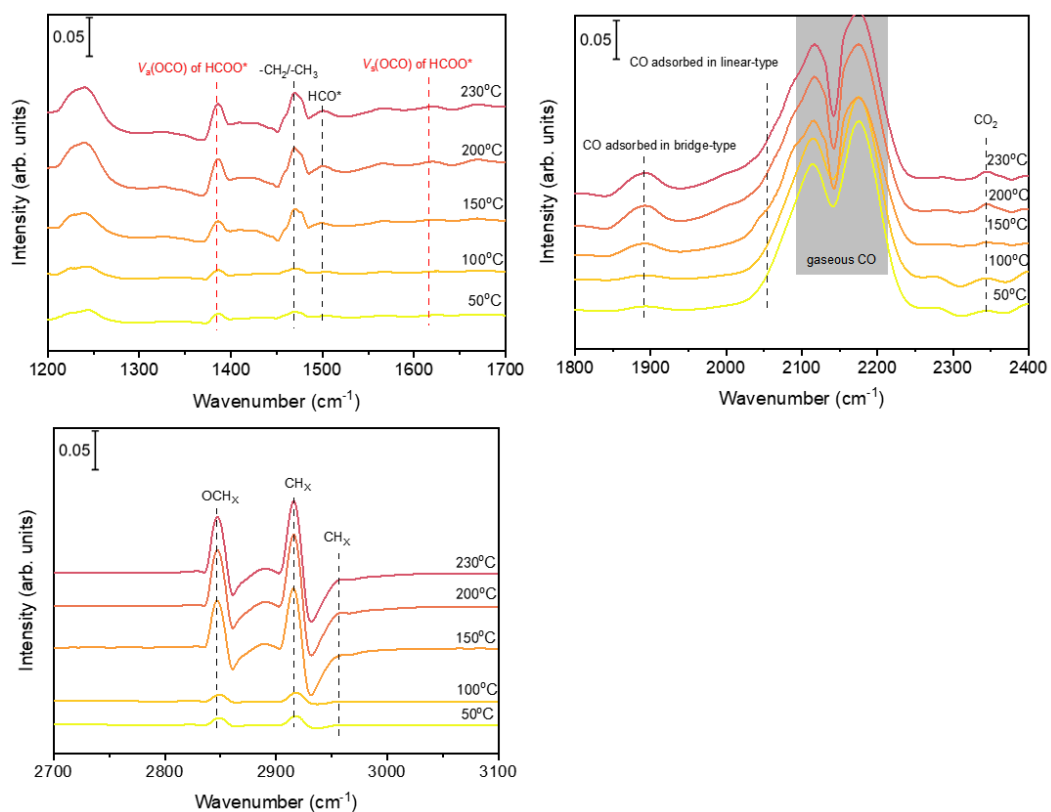
Supplementary Fig. 41 | Dependence of CO conversion on CO partial pressure.



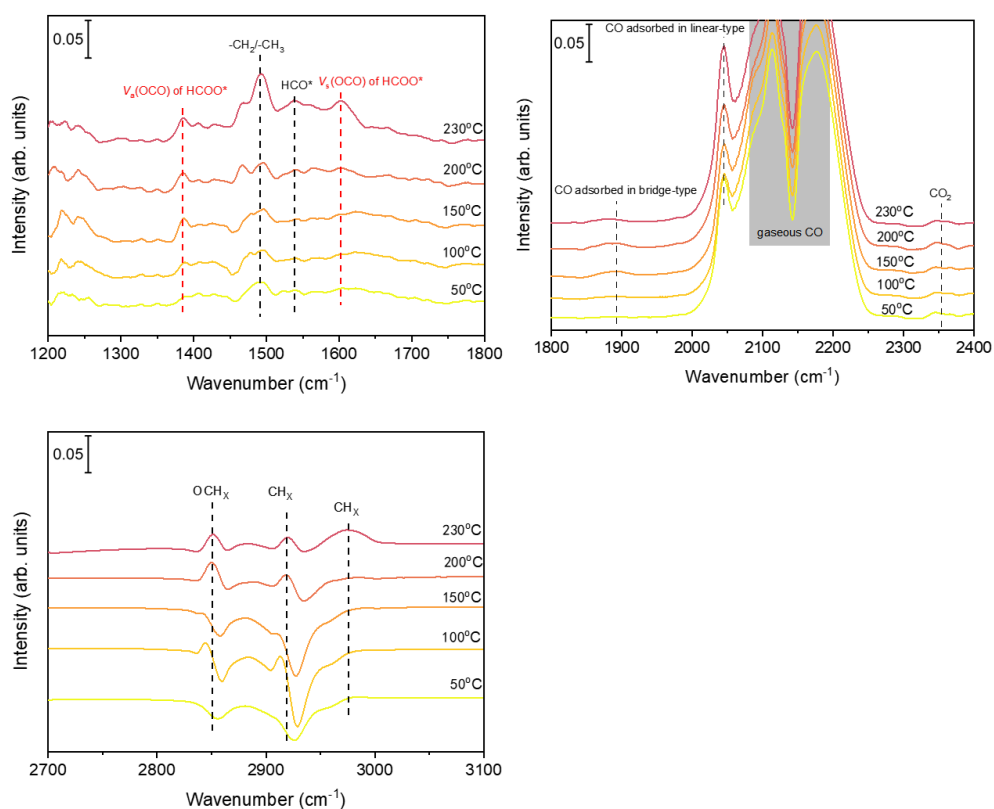
Supplementary Fig. 42 | H_2/D_2 isotopic exchange curves of CoMn/ β and CoMn/ $\beta_{0.2}$ catalyst.



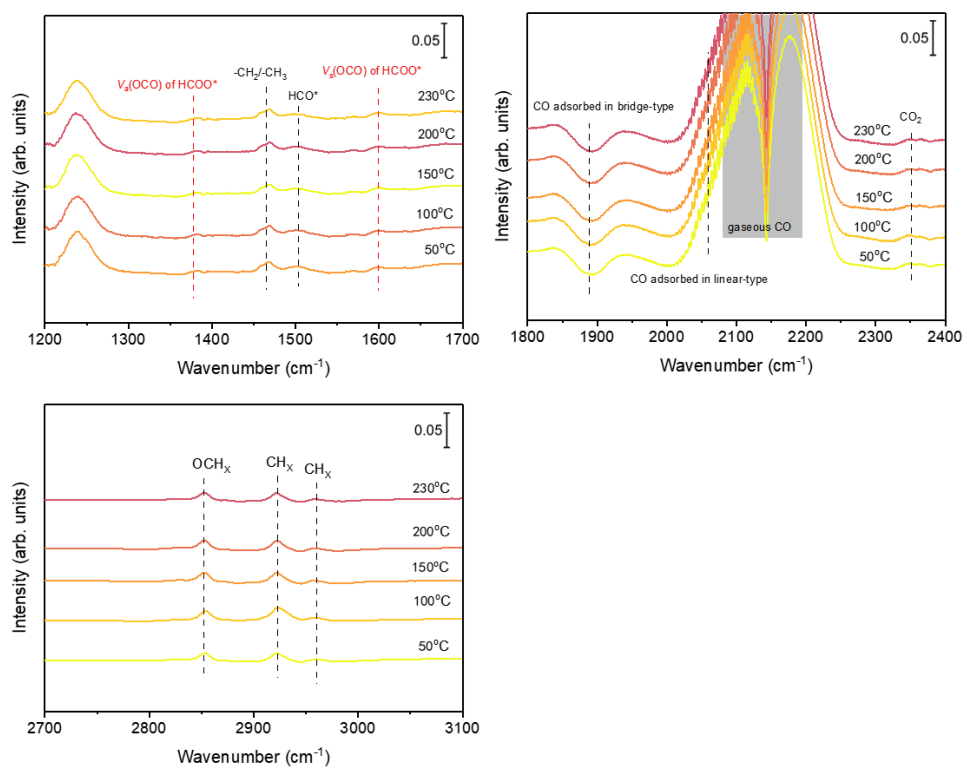
Supplementary Fig. 43 | C₂H₄ pulse hydrogenation result of CoMn/ β and CoMn/ $\beta_{0.2}$ catalyst.



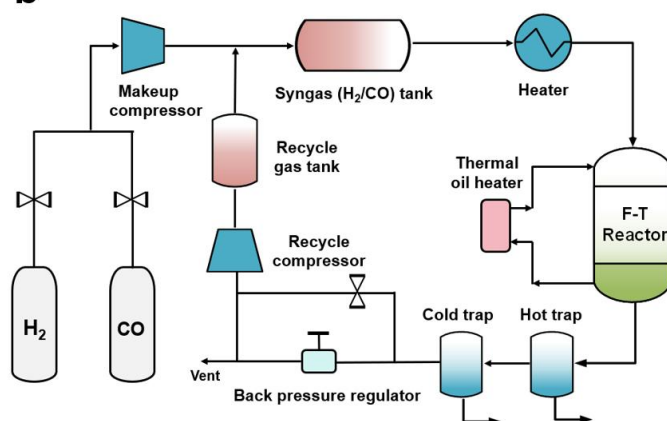
Supplementary Fig. 44 | Operando DRIFTS spectra of CoMn/β_{0.2} under a mixture of CO/H₂ (CO/H₂ = 1/2) from 50 to 230 °C at 1.5 MPa.



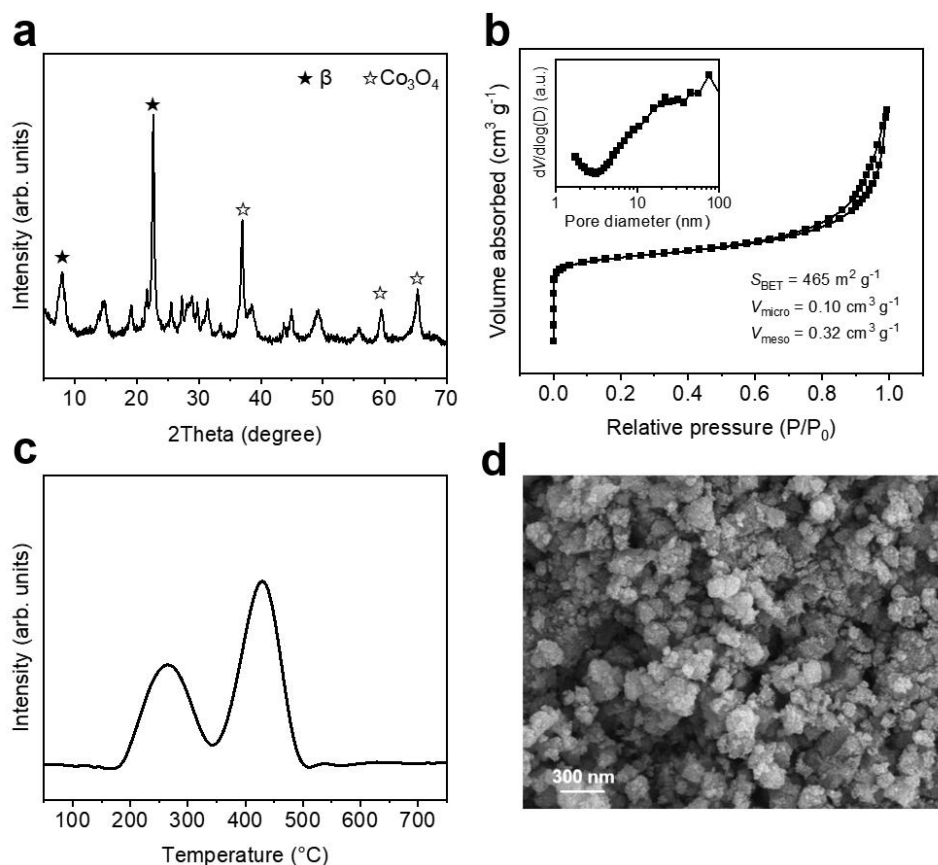
Supplementary Fig. 45 | In situ DRIFTS spectra of CoMn/β_{0.2} under a mixture of CO/H₂ (CO/H₂ = 1/2) from 50 to 230 °C at 0.1 MPa.



Supplementary Fig. 46 | Operando DRIFTS spectra of CoMn/β under a mixture of CO/H₂ (CO/H₂ = 1/2) from 50 to 230 °C at 1.5 MPa.

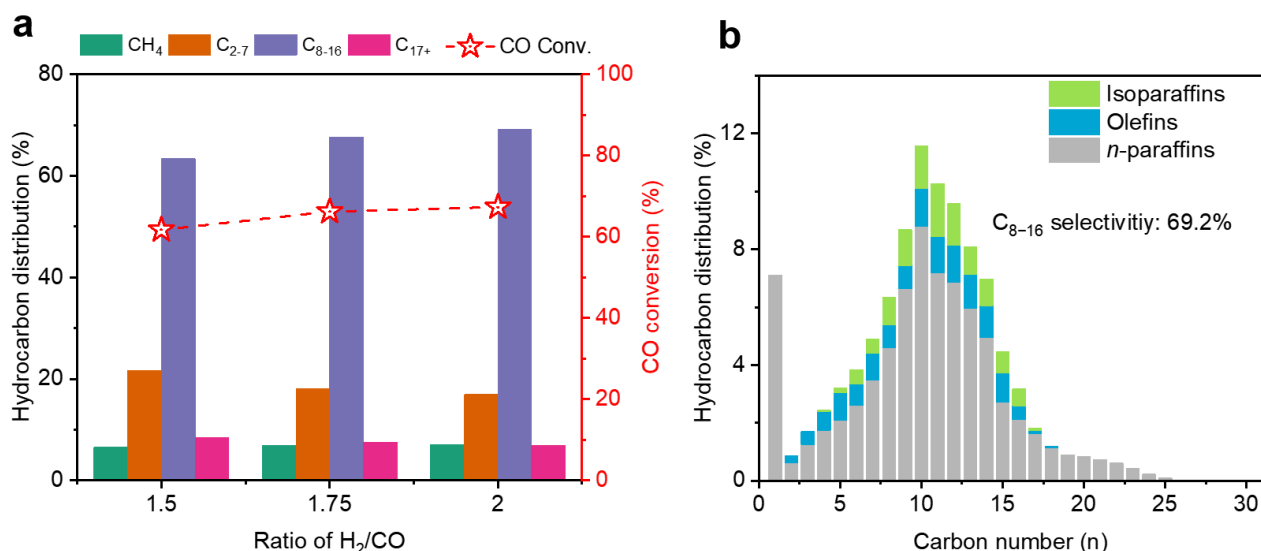
a**b**

Supplementary Fig. 47 | (a) Photograph of the pilot-scale fixed-bed reactor apparatus. (b) Simplified schematic process flow diagram of the reaction system.



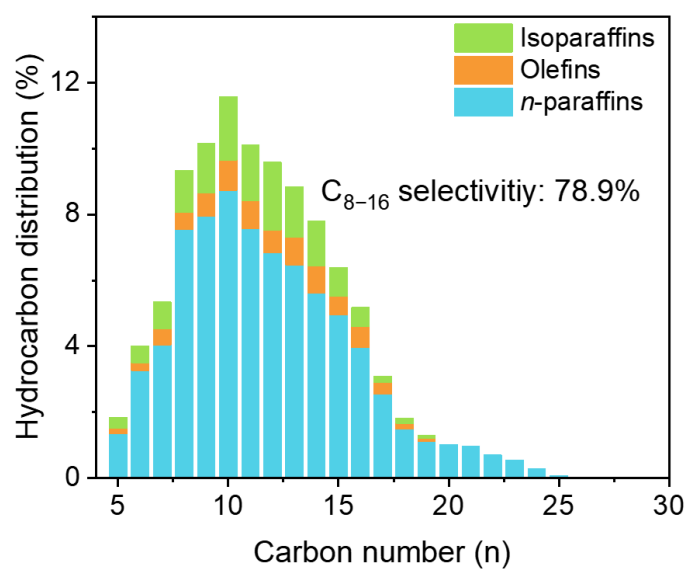
Supplementary Fig. 48 | (a) XRD pattern, (b) N_2 adsorption-desorption isotherm and pore size distribution (inset), (c) H_2 -TPR, and (d) SEM of pelletized $\text{CoMn}/\beta_{0.2}$.

Note: XRD verified the retention of the β zeolite framework and Co_3O_4 phase, N_2 sorption demonstrated that the hierarchical micro/mesoporous structure was maintained, H_2 -TPR confirmed the unchanged reducibility of Co species, and SEM showed the uniform particle morphology of the pelletized catalyst.

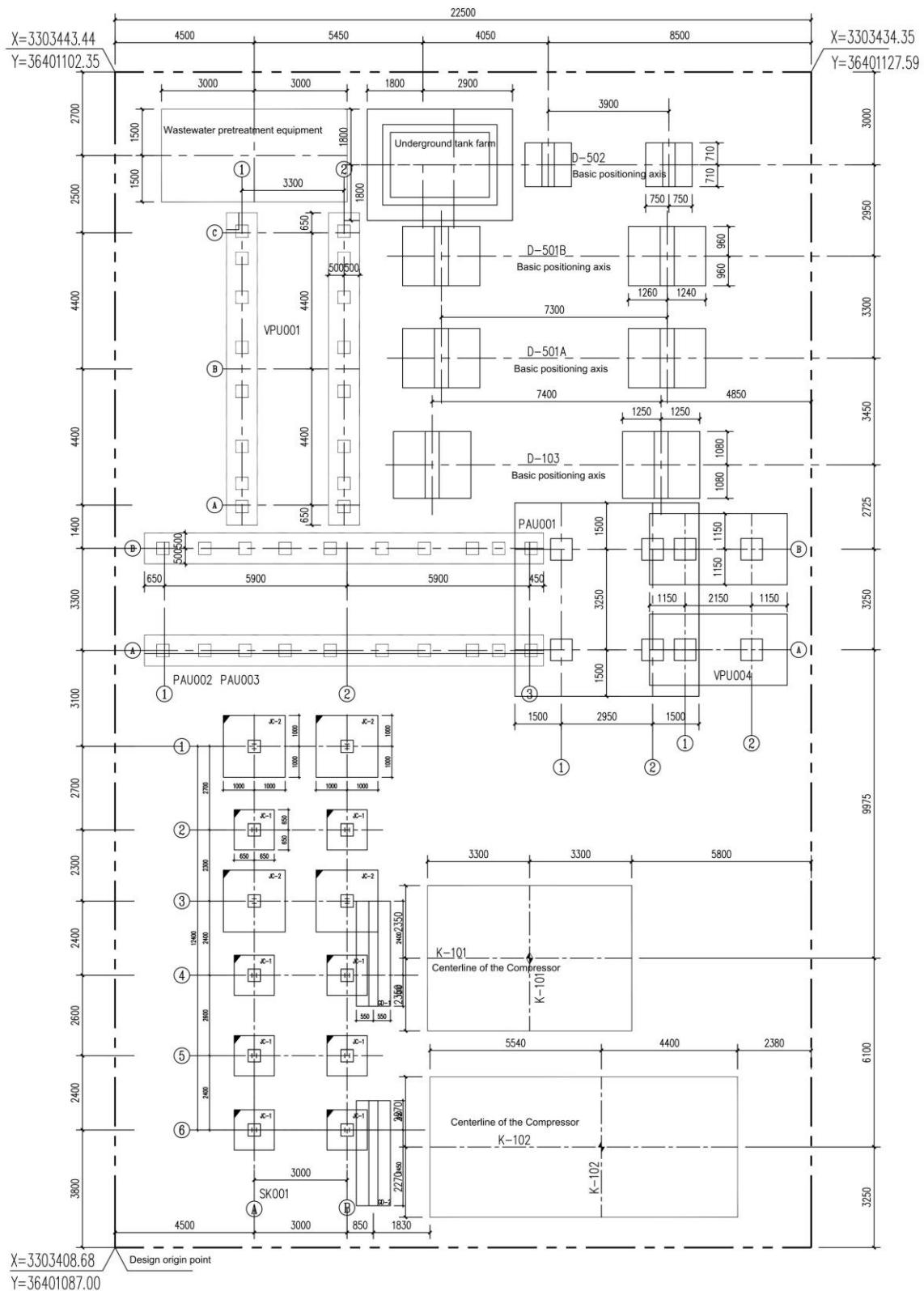


Supplementary Fig. 49 | (a) FT performance of the pelletized CoMn/β_{0.2} as a function of H₂/CO ratio in the pilot-scale fixed-bed reactor under industrial reaction conditions of 230 °C, 1.5 MPa, and 3000 mL·g_{cat}⁻¹·h⁻¹. (b) Carbon number distribution under P = 1.5 MPa, T = 230 °C, WHSV = 3000 mL g_{cat}⁻¹ h⁻¹, and H₂/CO = 2.

Note: To address the highly exothermic nature of the FT synthesis, we first employed a cautious strategy of gradually increasing the H₂/CO feed ratio during evaluation in a pilot-scale fixed-bed reactor. Under steady-state conditions with a H₂/CO ratio of 2:1, the pelleted catalyst achieved a CO conversion of 67.3% and a remarkable 69.2% selectivity towards C₈₋₁₆ hydrocarbons, corresponding to a jet fuel CTY of 2.537 g g_{Co}⁻¹ h⁻¹.



Supplementary Fig. 50 | Carbon number distribution of liquid products with jet fuel selectivity calculated based on liquid hydrocarbons under reaction conditions of 1.5 MPa, original feed gas = 800 mL·g_{cat}⁻¹·h⁻¹, recycle ratio = 3, and H₂/CO = 2.



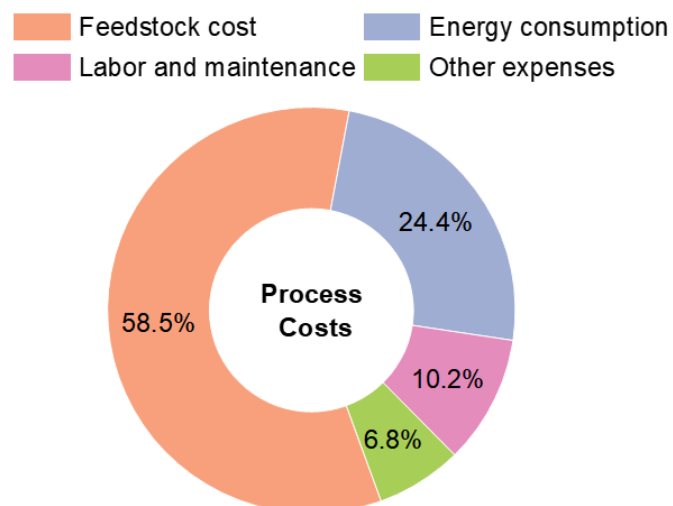
Supplementary Fig. 51 | Simplified plot layout of 1000 tonnes/year FT-to-jet fuel production facility.



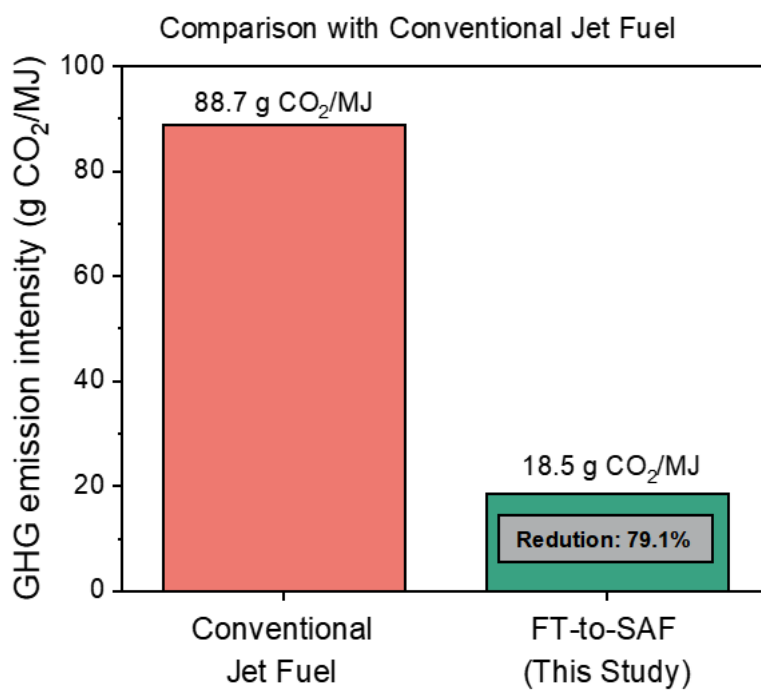
Supplementary Fig. 52 | Panoramic view of the 1000 tonnes/year FT-to-jet fuel production facility.



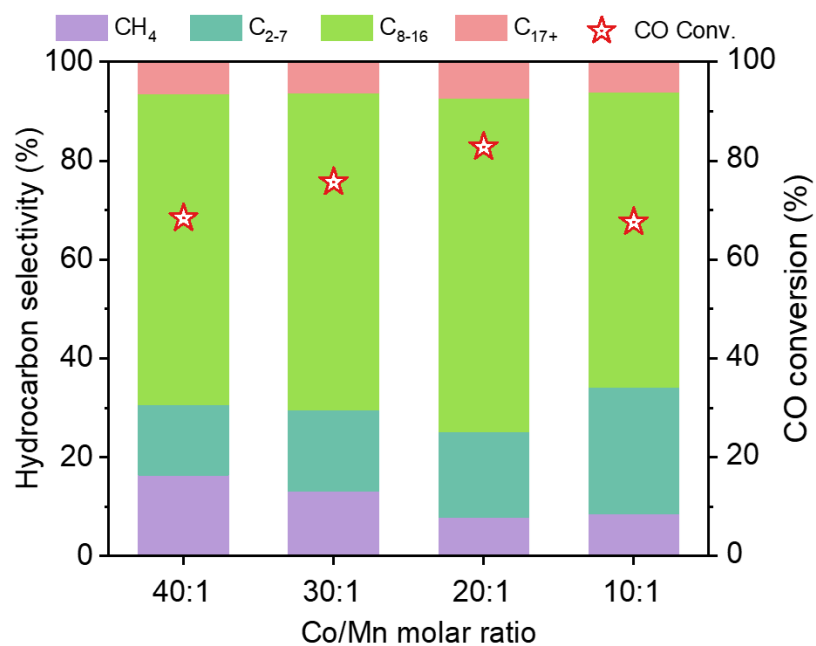
Supplementary Fig. 53 | Fischer–Tropsch jet fuel produced from the 1000 tonnes/year pilot unit.



Supplementary Fig. 54 | Operating expenses for the annual SAF products of 100,000-tonnes.



Supplementary Fig. 55 | GHG emissions contributions for the CoMn/ $\beta_{0.2}$ modeled cases FT-to SAF and Conventional jet Fuel.



Supplementary Fig. 56 | Performance of CoMn/ $\beta_{0.2}$ catalysts with different Co/Mn molar ratios at standard reaction conditions of 1.5 MPa, 230 °C, and WHSV=3000 mL·g⁻¹·h⁻¹.

Supplementary Tables and Notes:

Supplementary Table 1 | Summary of catalytic performances for direct synthesis of jet fuel via FT.

Catalyst	Cobalt loading (wt %)	T (°C)	P (MPa)	WHSV (mL·h ⁻¹ g ⁻¹)	H ₂ /CO Ratio	CO Conv. (%)	CH ₄ Sel. (%)	Jet fuel Sel. (%)	Jet fuel CTY ^c (g g _{Co} ⁻¹ h ⁻¹)	Reference
Co/Y-β-3	9	260	2.0	1000	2	83.7	18.7	40.4	0.782	4
10CB-MM	10	230	3.0	3120	2	34.6	14.0	40.2	0.903	5
Co/H-mesoBETA	9.6	240	2.0	4 ^a	2	32.8	14.7	41.0	1.166	6
10CoMn/Z5-0.5M	10	240	2.0	3000	2	53.4	18.8	50.2	1.674	7
Co-14/Z5	19.2	240	2.0	3000	2	57.4	20.9	43.5	0.804	2
Co/Y-meso-La	15	250	2.0	2240	1	40.0	9.6	72.0	1.344	8
Co/H-meso Y	10	240	2.0	4 ^a	2	17.8	16.8	39.5	0.585	6
Co/Al-SBA-15	15	230	1.0	4000	2	36.8	19.5	52.4 ^b	1.070	9
Co/Al-TUD-1	15	240	2.0	2400	1	47.2	7.8	51.3	0.806	10
Co@3Si&2Z5	~56	240	2.0	3600 ^d	2	85.1	6.3	75.2	0.823	11
Co@SA-0.1	20	240	2.0	3600	2	72.5	6.6	68.8	1.796	12
CoMn/β _{0.2}	18.1	230	1.5	3000	2	82.8	7.9	67.5	1.960	This work
CoMn/β _{0.2}	18.1	230	1.5	4500	2	74.7	10.6	63.8	2.508	This work

(a) The space velocity is given in terms of volume per hour (h⁻¹). (b) The selectivity of C₈₋₁₈ jet fuel. (c) Catalyst time yield (CTY) refers to the weight of jet fuel produced per gram of cobalt per hour. (d) not including ZSM-5 for space velocity calculation

Supplementary Table 2 | Textural Properties of the β and β_a zeolites.

Sample	S_{BET} ($\text{m}^2 \cdot \text{g}^{-1}$) ^a	S_{meso} ($\text{m}^2 \cdot \text{g}^{-1}$) ^b	V_{micro} ($\text{cm}^3 \cdot \text{g}^{-1}$) ^c	V_{meso} ($\text{cm}^3 \cdot \text{g}^{-1}$) ^d	$\text{SiO}_2/\text{Al}_2\text{O}_3$ ratio ^e
β	600	78	0.13	0.09	40
$\beta_{0.1}$	626	110	0.12	0.22	37
$\beta_{0.2}$	653	176	0.12	0.39	32
$\beta_{0.5}$	672	223	0.06	0.51	21

(a) BET surface area calculated by BET method; (b) BJH adsorption cumulative surface area of pores in 2–50 nm (c); the micropore volume evaluated by t-plot method; (d) the mesopore volume evaluated by the BJH method; (e) measured by XRF.

Supplementary Table 3 | The acidity distribution of the β and β_a zeolites.

Sample	NH ₃ adsorption ($\mu\text{mol}\cdot\text{g}^{-1}$)			Pyridine adsorption ($\mu\text{mol}\cdot\text{g}^{-1}$)		
	Weak	Medium/Strong	Total	Brønsted	Lewis	Brønsted/Lewis
β	170	235	405	182	189	0.962
$\beta_{0.1}$	153	194	347	105	109	0.963
$\beta_{0.2}$	145	124	269	92	96	0.958
$\beta_{0.5}$	127	116	243	49	51	0.960

Supplementary Table 4 | Physicochemical properties for CoMn/ β and CoMn/ β_a catalysts.

Catalysts	Cobalt loading (wt %) ^a	Reduction degree (%) ^b	H ₂ chemisorption amount ($\mu\text{mol g}^{-1}$) ^c	Co dispersion (%) ^d	Co diameter (nm) ^e
CoMn/ β	17.2	69.8	65.1	5.5	17.5
CoMn/ $\beta_{0.1}$	18.4	74.4	83.9	7.1	13.5
CoMn/ $\beta_{0.2}$	17.8	81.2	115.7	8.4	11.4
CoMn/ $\beta_{0.5}$	17.8	75.4	87.0	7.2	14.1

(a) Measured by ICP-OES; (b) measured by the H₂-TPR at 150-750 °C; (c) evaluated from the H₂ chemisorption measurement; (d) calculated from the H₂ chemisorption amount; (e) calculated using the following equation: Co diameter = 0.96/dispersion

Supplementary Table 5 | FT performance of the catalysts.

Catalysts	CO	CO ₂	Hydrocarbons selectivity				Jet fuel	Jet fuel
	conv.	sel.	(%)				yield	CTY
	(%)	(%)	CH ₄	C ₂₋₇	C ₈₋₁₆	C ₁₇₊	(%)	(g g _{Co} ⁻¹ h ⁻¹)
CoMn/SiO ₂	58.4	2.3	11.6	12.0	32.3	44.1	18.8	0.588
CoMn/Al ₂ O ₃	62.3	1.5	14.3	15.5	36.4	33.8	22.7	0.708
CoMn/ZSM-5	49.6	0.9	20.7	27.9	42.5	8.9	21.1	0.658
CoMn/Y	40.3	1.1	14.1	32.5	45.2	8.2	18.2	0.569
CoMn/ β	61.4	1.8	12.4	30.7	47.8	9.1	29.3	1.065
CoMn/ $\beta_{0.1}$	68.2	1.5	10.6	27.4	51.6	10.4	35.2	1.194
CoMn/ $\beta_{0.2}$	82.8	0.9	7.9	17.2	67.5	7.4	55.9	1.960
CoMn/ $\beta_{0.5}$	73.7	0.9	9.6	19.4	60.3	10.7	44.4	1.559

Reaction condition: H₂/CO = 2.0, P = 1.5 MPa, T = 230 °C, and WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

Supplementary Table 6 | Effects of reaction temperature on the FT performance of CoMn/ $\beta_{0.2}$ catalyst.

Temperature (°C)	CO	CO ₂	Hydrocarbons selectivity				Jet fuel	Jet fuel
	conv.	sel.	(%)				yield	CTY
	(%)	(%)	CH ₄	C ₂₋₇	C ₈₋₁₆	C ₁₇₊	(%)	(g g _{Co} ⁻¹ h ⁻¹)
210	40.4	0.8	6.1	13.5	69.7	10.8	28.2	0.988
220	62.6	0.9	7.3	15.7	68.3	8.7	42.8	1.500
230	82.8	0.9	7.9	17.2	67.5	7.4	55.9	1.960
240	85.3	1.1	13.3	21.8	57.9	7.0	49.4	1.732
250	91.5	1.3	14.4	24.7	54.1	6.8	49.5	1.736

Reaction condition: 210–250 °C, H₂/CO = 2.0, P = 1.5 MPa, and WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

Supplementary Table 7 | Effects of reaction pressure on the FT performance of CoMn/ $\beta_{0.2}$ catalyst.

Pressure (MPa)	CO	CO ₂	Hydrocarbons selectivity				Jet fuel	Jet fuel
	conv.	sel.	(%)				yield	CTY
	(%)	(%)	CH ₄	C ₂₋₇	C ₈₋₁₆	C ₁₇₊	(%)	(g g _{Co} ⁻¹ h ⁻¹)
1.0	72.4	4.2	15.5	28.9	48.7	6.9	35.3	1.237
1.5	82.8	0.9	7.9	17.2	67.5	7.4	55.9	1.960
2.0	85.6	0.9	8.4	17.9	62.9	10.8	53.8	1.889
2.5	90.2	1.3	7.6	17.1	58.3	17.0	52.6	1.845
3.0	90.7	1.2	6.2	16.5	54.8	22.5	49.7	1.743
4.0	92.3	1.2	6.0	12.2	54.0	27.8	49.8	1.748

Reaction condition: 230 °C, H₂/CO = 2, P = 1.0–4.0 MPa, and WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

Supplementary Table 8 | Effects of WHSV on the FT performance of CoMn/ $\beta_{0.2}$ catalyst.

Pressure (MPa)	CO	CO ₂	Hydrocarbons selectivity				Jet fuel	Jet fuel
	conv.	sel.	(%)				yield	CTY
	(%)	(%)	CH ₄	C ₂₋₇	C ₈₋₁₆	C ₁₇₊	(%)	(g g _{Co} ⁻¹ h ⁻¹)
1500	85.5	1.2	7.7	16.0	61.9	14.4	52.9	0.928
3000	82.8	0.9	7.9	17.2	67.5	7.4	55.9	1.960
4500	74.7	0.8	10.6	18.7	63.8	6.9	47.7	2.508
6000	70.5	0.7	12.1	21.8	60.1	6.0	42.4	2.972

Reaction condition: 230 °C, H₂/CO = 2, P = 1.5 MPa, and WHSV = 1500–6000 mL g_{cat}⁻¹ h⁻¹.

Supplementary Table 9 | Effects of the H₂/CO ratios on the FT performance of CoMn/β_{0.2} catalyst.

H ₂ /CO ratio	CO	CO ₂	Hydrocarbons selectivity				Jet fuel	Jet fuel
	conv.	sel.	(%)				yield	CTY
	(%)	(%)	CH ₄	C ₂₋₇	C ₈₋₁₆	C ₁₇₊	(%)	(g g _{Co} ⁻¹ h ⁻¹)
1.0	70.3	1.2	6.9	17.4	66.4	9.2	46.7	2.460
1.5	74.5	0.9	7.2	20.8	63.1	8.9	47.0	1.980
2.0	82.8	0.9	7.9	17.2	67.5	7.4	55.9	1.960
2.5	84.2	0.7	11.3	22.5	59.4	6.8	50.0	1.507
3.0	89.5	0.8	12.4	21.7	59.8	6.1	53.5	1.409

Reaction condition: 230 °C, H₂/CO = 1.0–3.0, P = 1.5 MPa, and WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

Supplementary Table 10 | FT performance of the pelletized CoMn/ $\beta_{0.2}$ as a function of H₂/CO ratio in the pilot-scale fixed-bed reactor.

H ₂ /CO ratio	CO	CO ₂	Hydrocarbons selectivity				Jet fuel	Jet fuel
	conv.	sel.	(%)				yield	CTY
	(%)	(%)	CH ₄	C ₂₋₇	C ₈₋₁₆	C ₁₇₊	(%)	(g g _{Co} ⁻¹ h ⁻¹)
1.5	61.7	2.4	6.5	21.7	63.4	8.4	39.1	2.161
1.75	66.2	2.2	6.8	18.0	67.7	7.5	44.8	2.476
2.0	67.3	2.2	7.1	16.9	69.2	6.8	46.6	2.573

Reaction condition: 230 °C, H₂/CO = 1.5–2, P = 1.5 MPa, and WHSV = 3000 mL g_{cat}⁻¹ h⁻¹.

Supplementary Table 11 | FT performance of pelletized CoMn/ $\beta_{0.2}$ with internal gas recycling as functions of the recycle ratio in the pilot-scale fixed-bed reactor.

Recycle ratio	CO conv. (%)	Hydrocarbons selectivity (%)			Liquid hydrocarbons selectivity (%)			Jet fuel CTY (g g _{Co} ⁻¹ h ⁻¹)
		CH ₄	C ₂₋₄	C ₅₊	C ₅₋₇	C ₈₋₁₆	C ₁₇₊	
1	75.4	6.5	11.8	81.7	16.0	74.4	9.6	2.896
2	83.7	6.8	10.7	82.5	14.9	76.7	8.5	3.478
3	92.4	9.0	6.1	84.9	11.3	78.9	9.8	5.154

Reaction condition: 240 °C, H₂/CO = 2, 1.5 MPa, and original feed gas = 800 mL g_{cat}⁻¹ h⁻¹. The recycle ratio represented the flow rate of the recycled tail gas over that of the original feed gas. The liquid hydrocarbons selectivity was estimated from the liquid product analysis.

Supplementary Table 12 | Key material flow rates, compositions, and properties for SAF production from biomass gasification-derived syngas over the CoMn/ $\beta_{0.2}$ catalyst.

Unit	S1		S2		S3		S4		S5		S6		S7	
Stream description	Biomass		High-pressure O ₂		Raw syngas		Shifted syngas		Purified syngas		SAF product		Naphtha by-product	
Component	kg/h	mol%	Nm ³ /h	mol%	Nm ³ /h	mol%	Nm ³ /h	mol%	Nm ³ /h	mol%	kg/h	wt%	kg/h	wt%
CO					69090.61	40.77%	30996.55	19.27%	30531.6	32.39%				
H ₂					25818.39	15.23%	63903.68	39.74%	62863.4	66.70%				
CO ₂					26862.43	15.85%	64699.87	40.23%	15.59	0.02%				
N ₂			9.06	0.03%	772.67	0.46%	772.62	0.48%	733.8	0.78%				
CH ₄					41.27	0.02%	41.26	0.03%	0.89	0.00%				
H ₂ S+COS					49.13	0.03%	48.68	0.03%	0	0.00%				
AR			111.69	0.37%	111.69	0.07%	111.68	0.07%	106.07	0.11%				
H ₂ O					46666.89	27.54%	244.58	0.15%	0	0.00%				
NH ₃					61.41	0.04%	0	0.00%	0	0.00%				
O ₂			30066.22	99.60%										
SAF											12500	100.00%		
Naphtha													1687.5	100.00%
C17+														
Biomass (s)	99922	90.00%												
H ₂ O(l)	11102	10.00%												
Pressure (MPa)	0		5		4		3.5		3.2		0.4		0.4	

Supplementary Table 13 | Fixed investment cost of SAF production from biomass gasification-derived syngas based on CoMn/ $\beta_{0.2}$.

No.	Item	Cost (Million USD)	Note
A	Fixed investment	396.29	B+C+D+E
B	Engineering cost	325.17	1+2+3+4+5+6
1	-Process production units	223.61	
2	-Utilities	61.67	
3	-Auxiliary production facilities	9.81	
4	-Storage and transportation facilities	25.00	
5	-Service facilities	5.08	
C	Overall plant systems	30.28	
D	Other cost	5.56	
E	Land cost	35.28	

Supplementary Table 14 | Economic analysis of SAF production from biomass gasification-derived syngas based on CoMn/ $\beta_{0.2}$.

No.	Item	Cost (Million USD)	Notes
A	Operating cost	113.89	1+2+3+4
1	-Feedstock cost	66.67	Refer to Supplementary Table 12
2	-Energy consumption	27.78	Electricity, steam, and H ₂ purification
3	-Labor and maintenance	11.67	
4	-Other expenses	7.77	Certification, management, insurance
B	Depreciation cost	23.81	Refer to Supplementary Table 13 and annual depreciation rate is 6.7%
C	Sales revenue	302.43	Refer to Supplementary Table 15
D	Gross profit	164.73	C-A-B

Supplementary Table 15 | Sales of products for SAF production from biomass gasification-derived syngas based on CoMn/ $\beta_{0.2}$.

No.	Product	Unit price (USD /t)	Sales volume (t)	Total amount (Million USD)
1	SAF	2950	100,000	295
2	By-products	550	13,500	7.43

Supplementary Table 16 | Economic analysis of SAF production from syngas based on CoMn/SiO₂.

No.	Item	Cost (Million USD)	Note
1	Raw material	51.55	Refer to Supplementary Table 17
2	Consumables (Catalyst, etc.)	0.48	
3	Labor-dependent	1.43	
4	Facility-dependent	2.74	Refer to Supplementary Table 18
5	Quality Inspection	0.03	
6	Utilities	3.19	Refer to Supplementary Table 18
7	Other cost	0.64	
8	Full cost	60.06	1+2+3+4+5+6+7
9	Sales revenue	170.77	Refer to Supplementary Table 19
10	Gross profit	110.72	9-8

Supplementary Table 17 | Consumption of raw materials of SAF production from syngas based on CoMn/SiO₂.

Item	Total consumption (t)	Unit price (USD/t)	Cost (Million USD)
Syngas	257760	200	51.55
	Total		51.55

Supplementary Table 18 | Detailed explanation of facility-dependent cost of SAF production from syngas based on CoMn/SiO₂.

No.	Item	Cost (Million USD)	Note
	Direct fixed capital cost	17.11	A+B+C
A	Physical cost	14.07	1+2+3+4+5+6+7
1	Equipment purchase cost	3.20	
2	Installation	6.39	
3	Process piping	1.28	
4	Instrumentation & insulation	0.96	
5	Electrical	0.64	
6	Buildings & Yard improvement	0.96	
7	Auxiliary facilities	0.64	
B	Total plant indirect cost	2.88	8+9
8	Engineering	1.28	
9	Construction	1.60	
C	Contractor's fee & Contingency	0.16	

Supplementary Table 19 | Sales of products of SAF production from syngas based on CoMn/SiO₂.

No.	Product	Sales volume (t)	Unit price (USD/t)	Total amount (Million USD)
1	Methane	750	571	0.43
2	LPG	2830	643	1.82
3	Naphtha	7400	1000	7.40
4	SAF	35480	2950	104.67
5	Diesel	14560	1071	15.59
6	Paraffin	33660	1214	40.86

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